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June 5, 6, and 7, 1979

The National Biomass Program

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THE NATIONAL BIOMASS PROGRAM

JUNE 5, 6, AND 7, 1979
COLORADO SCHOOL OF MINES
GOLDEN, COLORADO

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PROCEEDINGS
3RD ANNUAL BIOMASS ENERGY SYSTEMS CONFERENCE

INTRODUCTION

LARRY J. DOUGLAS
CONFERENCE TECHNICAL CHAIRMAN
SOLAR ENERGY RESEARCH INSTITUTE
GOLDEN, COLORADO

A national commitment to developing solar resources to furnish twenty percent of our total energy demands by the year 2000, requires increased emphasis applied to the near- and mid-term opportunities in solar energy. Biomass in its various forms is a renewable resource that can be converted to energy, fuels, and chemicals. The production and conversion technologies for biomass fall in the near and mid-term category, and therefore provides a very timely subject for review. The Department of Energy Biomass Energy Systems (DOE/BES) branch (formally Fuels From Biomass) supports a number of research projects relating to production of plant biomass, and conversion of biomass to fuels and energy intensive chemicals using biological and thermochemical methods.

The 3rd Annual Biomass Energy Systems Conference sponsored by DOE and hosted by SERI was held at the Colorado School of Mines, Green Center, June 5-7, 1979, and had as its major objectives the following:

1. Review the latest research findings from federally funded and other selected projects in the Clean Fuels For Biomass field.
2. Summarize the present engineering and economic status of Biomass Energy Systems.
3. Encourage interaction and information exchange among people working or interested in the field.
4. Identify and discuss existing problems related to ongoing research; explore opportunities for future research.

Biomass has been called the "sleeping giant" of the solar technologies, with a potential quad impact of 6-10 quads/year by the year 2000, however, the interest displayed at this conference and others held around the country this year suggests that awareness of the biomass potential is awakening. This conference served as a forum for discussing the problems and opportunities in biomass, and problems there are. The overriding problem was not how to grow or convert the biomass in a technological sense, but how to move the technology to fit into a framework that matches the economic realities of today and tomorrow.

As a final note, the conference posed some very difficult questions relating to economics and process development problems...however, it has been said that "a well defined problem is often the key to the door of opportunity".

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SUMMARY OF DISCUSSION SESSIONS

The Wednesday evening discussion sessions were held as an open forum reviewing the problems and opportunities in the five main areas in Biomass Energy Systems:

1. Biomass Refining (fermentation)
2. Anaerobic Digestion
3. Biomass Production
4. Thermochemical Conversion
5. Biomass As A Fuel (combustion)

These discussions provided a useful input for evaluating the direction of the current program both from the perspective of D.O.E. contractors and interested participants outside of this group. I wish to personally thank all the participants in these sessions for a valuable contribution and to those who could not attend these sessions. We are including highlights from each group.

BIOMASS REFINING

About forty participants took part in the discussion/workshop on biomass refining (fermentation). The discussion was lively and listed below are some of the main arguments with statement answers.

Why Produce Ethanol From Biomass?

Pro:
- Ethanol burns as a liquid fuel extender.
- Technology is well developed.

Con:
- Separation energy is high.
- Conversion to chemical feedstocks show better economics.

General Discussion:

- Ethanol needs to be produced for less than $1.00/gal. to be competitive with methanol.
- Maleic acid offers an attractive alternative to producing acetic acid.
- Membrane separation may offer some economic benefits.

Acid vs. Enzyme For Hydrolysis

- Acid hydrolysis is faster.
- Enzyme hydrolysis produces fewer degradation products.
- Single pass acid hydrolysis only converts 55 percent of cellulose.
- Acid hydrolysis good for removing hemicellulose.
- Enzymatic hydrolysis compatible with combining fermentation and hydrolysis.

Why 100 Percent Ethanol?

- Minimizes problems of water in gas lines.
- 85 percent may burn more efficiently.
- Add emulsifier to prevent falling out (separation) of water from gasahol.

General Comments

- With a feedstock cost of $30.00/ODT and no by-product credit ethanol starting point for materials cost is $1 to $1.25/gal.
- Can small farm units work economically? Projected cost for small ethanol plant is $3-$7/gal.
- Should feedstock determine process design, i.e., certain lignocellulosic materials because of their contents of lignin, hemicellulose and cellulose may require different pretreatments and/or processing procedures.
- Development of continuous fermentation systems needed to reduce capital cost and plant size.
- Development of by-product utilization vital to economics of ethanol production.

ANAEROBIC DIGESTION

Approximately 25 persons attended the discussion session which Chairman Don Wise focused on the question of how to effectively digest crop residues to methane. A wide variety of opinions were expressed and a number of suggestions for a crop residue program were made. The questions discussed included:

- Short-term applications vs. long-term. Is it realistic to try and commercialize crop residue systems now, or will it only be feasible many years from now, when gas prices are much higher.

- Environmental Effects: Do we know what long-term effects to expect if we remove large quantities of crop residues from the soil? Will we be able to recycle the nutrients?
- Alternate Conversion Processes: Assuming we decide to convert large amounts of crop residues to energy, should we hydrolyze the residues and produce alcohols, burn the material to produce heat, or digest the residues to methane? What are the trade-offs? Should the hydrolysis process be used to first make alcohols, and the digestion used to further convert the effluent from an alcohol plant?
- Biomass Production: Should farmers be encouraged to grow corn which produces less grain and more residues, since we have a surplus of grain?

BIOMASS PRODUCTION

- Weeds were designated as one of the prime problem areas in silviculture energy farming. U.S.F.S., Rhinelander, Wisconsin is conducting an extensive field research effort on this problem and some promising results are being obtained.....a pre-emergent herbicide coupled with the planting of a leguminous cover crop seems to offer good control. Research will be continued.
- Lack of a sustained wood supply is troubling Minnesota authorities. It appears that there is insufficient wood available in that state to serve in competition with pulp and paper as a lasting resource. Minnesotans are looking toward cattail as an energy crop.
- A number of aspects of the Gasahol question were discussed. It was generally agreed that biomass supply problems would also become prominent in regard to the production of alcohol.

THERMAL-CHEMICAL CONVERSION

About forty of the attendees in thermal-chemical conversion processes for biomass met Wednesday evening to discuss institutional and technical problems relevant to thermal conversion. The discussion was wide-ranging and sometimes heated but a number of interesting suggestions were made. The following are representative. (See institutional comments in SERI's Gasification Workshop Proceedings.)

- While air gasifiers are already being developed in the range 1-20 MBtu/hr. by private money, there is a need for larger gasifiers up to 120 MBtu/hr. or even 500 MBtu/hr. (30 dry tons/hr.). This size would be especially useful for retrofitting boilers in the paper industry and some rural utilities; yet the cost of development is such that no group has yet undertaken these larger boilers. If we are serious about using biomass to replace gas and oil this would be a very significant market - possibly 2 quads/yr. - need a mechanism for developing this size. It could possibly be done by a consortium of paper companies - or matching federal funds.
- Closely related is the need for a method of helping private capital do innovative development on a large scale. First installations typically cost 2-4 times as much as subsequent units. Who is to pay this cost? No industry is stepping forward to bear the cost. A possible role for the federal government is to fund the risk and development portion of these costs, that amount above expected sale price. In return, all records that are not proprietary should be made available to others working in the same area. There is a precedent for cost-sharing in developing coal processes; this should be studied by the biomass community for possible implementation or modification.
- There has long been a prejudice in U.S. coal circles against "foreign technology", Lurgi, Winkler, Koppers Totzek, Davy Gasifiers, etc.). This puts us in the unfortunate position of having to develop more complex processes, possibly of greater merit, but with no experience of the simple processes. Several participants hoped that the same prejudice would not be extended to foreign biomass gasifiers. Since there has been no experience in the U.S. with biomass gasification up to a few years ago and since there have been a number of gasifiers built abroad, (Imbert, Duvant, Davy, Century) we would try to build on this technology rather than circumvent it.
- There was some discussion of the proper role of government in developing biomass thermal conversion systems. All agreed that at a minimum, the governmental role was to supply and diseminate scientific and technical information to those working in the field. Several specific areas where further information is needed were discussed. Biomass occurs universally at a small scale, but as various processes require larger concentrations, the number of possible sites diminishes so that few sites can supply more than 1000 tons/day. A survey of possible larger sites having residues or other large supply bases, would be very useful for those working in thermal processing. A related question is that of the appropriate scale for various processes. Air gasification seems to be applicable at very small scale - liquefaction only at the largest scale. Another study which the attendees suggested would be the cost benefit of various end products from biomass. How much more is a liquid boiler fuel worth than gas? An automotive fuel? Chemical feedstocks? These need to be available for rating various thermal conversion processes.
- Finally, a large barrier to private money development of biomass thermal conversion

processes is the continuing uncertainty of future fuel costs. With conflicting directives coming from COE on the use of coal/gas/oil there is obviously a high risk involved in developing alternate fuels not yet tested in the market. Only by establishing a floor immune to daily policy changes on synthetic fuel prices can the risk of development be removed and private capital attracted.

BIOMASS UTILIZATION AS A FUEL (COMBUSTION)

Primary Points Brought Out During Discussion Group:

- Methods to improve first low efficiencies for high moisture biomass fuels.
- Areas where additional biomass combustion technology is needed.
  - Methods to improve first low efficiencies for high moisture biomass fuels.
  - Retrofit combustion technologies to allow existing oil, gas, and coal combustors to use biomass fuels.
- Only limited types of biomass fuels are currently being used.
- Fuel (biomass) transportation costs are a major factor in the overall economics.
- Trade-off between economies of scale for larger biomass burning plants and fuel transportation costs not defined.
- All biomass burning facilities must be examined on a site-by-site basis to determine -
  - available fuel resource
  - competing uses for the fuel resource
  - properties of the fuel (moisture, ash, heat release, etc.)
  - best types of combustors (fluid bed, spreader stoker, suspension, etc).

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STATEMENT OF HON. RICHARD L. OTTINGER,

New York Congressman-U.S. House of Representatives

THIRD ANNUAL BIOMASS ENERGY SYSTEMS CONFERENCE

Golden, Colorado

June 5, 1979

Good morning. It gives me great pleasure to be out here in Golden, away from those Potomac shores so dangerous to any vision of what is really happening in the world.

It is very rewarding to see so many people gathered here, all of whom are so interested in, committed to, and working on utilizing biomass for energy.

I wish I could be bearing good tidings of all the wonderful things we in Washington have done for biomass lately, but unfortunately I cannot. I have just finished combing through National Energy Plan II, that wonderful, unidentified document with the white cover that might lead one to think it's a dirty book (it may well be, come to think of it, considering some of the pollution likely to result from some of the proposed projects contained in NEP II). And so far as I can determine there is no mention of biomass whatsoever! I should hasten to add that my not-so-friendly reaction to NEP II reflects my own, personal views.

Also, at the end of last week, the Department of Energy submitted its annual report to the Congress; that document contains three short paragraphs about biomass, and that's all.

It is clear that biomass retains a low priority on the laundry list of initiatives this Administration seems wedded to pursuing, and in my own view this is extremely misguided—both because of the lack of emphasis on biomass and because of the overemphasis on technologies which, in my own opinion, are too costly in economic, social, and environmental terms and not likely to contribute to the sort of energy future which I think this Nation should be pursuing. Any of you who have been familiar with my efforts in Congress over the past five years are well aware that my preference is for emphasis on getting off oil imports by pursuing a course of intensive energy conservation and productivity and the use of renewable resources. Biomass, of course, fits right in with my scheme for our energy future.

As I look back over the past four or five years since we began providing funds for biomass, I find a confused, uncoordinated, non-coorperative series of non-initiatives, not only within the Department of Energy and its predecessors, but within other Federal agencies as well. There has been little coordination internally within DOE, and certainly little cooperation between DOE and other agencies involved. As an example of the internal problems, the Department's Fuels from Biomass program had no less than four administrators during a mere nine months ending this past January!

It is noteworthy that there is a plethora of other agencies with biomass programs and if, by dint of some new efforts at cooperating, DOE would work with them, we would have superb opportunity to move biomass to the position of stature is deserves. In addition to DOE, which itself has some 10 divisions with one role or another in biomass, there are programs in the Department of Agriculture, the Environmental Protection Agency, the Department of Defense, the Agency for International Development, the National Science Foundation, the Tennessee Valley Authority, the Department of Commerce and the National Aeronautics and Space Administration! It's almost enough to make your head spin.

Of most of those agencies I will say no more, although it seems to me absolutely critical that at least three of them—the Department of Energy, the Department of Agriculture and the Environmental Protection Agency—begin at once to improve their interagency cooperation and coordination. After all, with DOE worrying over energy in general, and the EPA with jurisdiction over wastes and pollution, and the Agriculture Department in charge of forestry and farming, some joint efforts could bring together most of the basic resources available for biomass energy production.

The crucial thing is to think of biomass resources as being available both from active production for the purpose—of wood or farm crops—and from secondary production resulting from biproducts from other endeavors--such as timber, crop, feedlot and human waste.

It seems to me in many ways that we are not just looking at the traditional interagency rivalry over turf, but we are in a sense looking at the existence of two basic economies: the taxpayer economy and the business economy. Where there is a "waste" problem, such as sewage "waste," solid "waste," feedlot "waste," woodlot "waste," or agricultural or timber "waste," we tend as a people simply to regard these as "waste" problems, and to deal with them as best we can, and at taxpayer expense. That would be the taxpayer economy. In the business economy, on the other hand, we have "resources," which would be the products extracted for public use and for profit—before the "wastes" are generated. It seems to me that we should be trying to look at the traditional "resources" and the "wastes" as part and parcel of the same thing. Indeed, "wastes" need to be moved over into the "resources" area, and shifted into the "for profit" column. In this use of the words "for profit," I am considering overall profits to all Americans in the broadest possible sense.

With that approach, then wastes become extremely important resources, especially since so many of them are in the truest sense renewable resources. If we regard the sources of biomass fuels only as things we must grow deliberately, we will be overlooking numerous available resources which currently only drain tax moneys to be dealt with. It may not be that feedlot waste or sewage sludge can make major "Quad" contributions to our national energy needs, but the sum of their smaller contributions has to be significant and, moreover, their reuse would help to solve a problem. An added appeal to utilizing farm and human wastes is that, unless I'm hopelessly naive about the nature of biological systems, they are not likely to diminish in quantity even in the face of abnormal occurances such as floods and droughts.

I might say here that one of the gravest problems I have perceived within the Department of Energy is the apparent inability to see that a host of small quantities of energy produced can add up to a major contribution. The obvious penchant within DOE is only to look for the big bang for megabucks, and that is what leads our policy-makers always to be seeking huge production facilities, whether new oil refineries, big coal gasification and liquification facilities, power plants or whatever. And even when a technology such as biomass production does emerge, the DOE tendency is to try to convert that technology into a national panacea.

This, it seems to me, is a serious mistake. We must get away from an insistence that we must pursue energy production as we have known it for forty years. DOE must learn that it does not necessarily make energy sense to rely solely upon the energy distribution grids as we now know them, or to prop up those grids with infusions of high-cost energy conversion processes simply because those grids exist. My perception is that this is the DOE's whole thrust, and that explains the emphasis on high-Btu, pipeline quality gases and liquids from coal and shale. And to my mind that misses the best opportunities for such energy resources as biomass, which opportunities lie in using the resources where they are found, either directly or through converting them to the most sensible fuels for the regions in which they are.

Under such an approach, we might make methanol from coal in the Southeast and Midwest, burn directly or gasify wood in the Northeast, Southeast and Northwest, make a biofuel from peat in Minnesota, North Carolina, Maine, New York and other areas where there are peat bogs, perhaps make gaseous fuels from sewage in the major cities, make ethanol or methanol from agricultural wastes and/or crops in the Midwest farm belt, make petrochemical feedstocks from various plant forms found in the Southwest and elsewhere, make biogas from feedlot wastes in Texas, Oklahoma, Colorado and the other major meat producting areas, and so on.

We have recently seen an incredible consciousness-raising regarding the use of alcohol fuels as a displacement for motor gasoline. That's nice, but need not be the only answer. Where there are captive fleets of vehicles, by companies, the Postal Service, the military, other biofuels surely could be used successfully. Only if the displacement fuels, such as gasohol, were to be promoted for general use nationally would we need a standard for quality and Btu content. It is conceivable that New Yorkers could utilize a 10% additive of methanol from sewage which, if some drove across the country, could be displaced by a methanol/ethanol combination from crops or agricultural wastes in other parts of the country. Similarly, fuels from biomass for stationery use would need to be compatible only within the region, unless such fuels were to be used for appliances and other national commodities, in which case we would need a standard there, too. For stationary use within a region, it is quite probable that low- and medium-Btu gas from a wide variety of products could be effectively utilized.

All of this, of course, would require a new world-view on the part of the DOE, and a department from our traditional dependence upon a relatively small number of energy distributors. The appeal in this for me is an inevitable increase in local and individual self-reliance which would serve extremely well that mood in the country which appears to be growing in strength, that is the desire for increased controls over our own lives and activities.

In conclusion, I believe we have become mired in an old vision as regards energy. It is time for new ideas, new approaches which I believe this audience exemplifies. Rather than the "naysaying" we hear so much of on our Potomac shores, the "we can't get there from here" attitude so prevalent in the Department of Energy, we must move in the direction of supporting the best of American ingenuity. All across the country it is being demonstrated that "we can get there from here." People are taking up the cudgels to do something about energy—and they are acting on their own and in groups in spite of, unfortunately, rather than because of the Federal Government. It delights me to see people moving out in front of us in Washington, chuckling over those who come in and say, "Hi, I'm from the Department of Energy, and I'm here to help you." It doesn't wash, and you here all know it.

As one final note, the Subcommittee which I chair on the House Science and Technology Committee will be conducting oversight hearings on the biomass programs later on this year. I would welcome any observations any of you have on how DOE is doing—or not doing—and any suggestions you might have for new legislative initiatives.

I commend you all for being here at this third biomass conference, and I thank you for your indulgence.

SERI AND THE BIOMASS PROGRAM

Dr. Clayton S. Smith
Solar Energy Research Institute
Golden, Colorado

ABSTRACT

The development of biomass as a resource base for fuels and energy saving chemicals will require the identification of appropriate supply, conversion technology and product applications for commercialization. Biomass does not presently enjoy a suitable technology base. SERI is presently developing the staff and facilities necessary to conduct analyses, selected technology research and development, and commercialization assistance on biomass options. SERI is also providing Technical Program Management for the aquatic production, anaerobic digestion and fermentation components of DOE's Biomass Energy Systems Program.

As a framework within which to discuss SERI's involvement in the Biomass program, I would first like to discuss the nature of the technology.

NATURE OF THE TECHNOLOGY

"Biomass" is the monolithic term applied to what is perhaps the most technically, economically, and socially complex of the solar options. This deceptively simplistic term hides the true nature of the resource and the routes to its use as a renewable energy source. Proponents agree that Biomass offers the potential for significant quad impact; yet a clear strategy for achieving this promise has not emerged. If Biomass is to realize this goal, a comprehensive program must be developed which takes into account the nature of the resource.

Biomass is distributed in supply and diverse in its physical and chemical characteristics. It is generally bulky and expensive to transport, thus has a limited economic collection area, quite unlike coal, petroleum, or natural gas. Thus, it must be produced, converted, and probably utilized on a local basis. Use of Biomass for fuel or energy saving chemicals is often competitive to food and fiber production, yet complementary situations do exist.

Generally the application drives the need for technology development. In nearly all energy related applications, Biomass must compete with coal, petroleum, and natural gas as the resource base. In many cases it may not be competitive; in some it clearly is competitive. Successful deployment of Biomass as an energy alternate must ferret out such situations.

A systems approach must be used to identify such ortunities. The following major components must considered:

1. The supply of Biomass, which may be forest, crops, aquatic, or waste material.
2. The conversion technology required to change Biomass feedstock to a useful energy product.
3. The product, which may be heat, electricity, or fuels and chemicals.
4. The application or market for the product. With Biomass, the application is likely to be near the supply, such as the combustion of corn stover for grain drying.

We believe that the cornerstone of the Biomass program must be the identification and development of Biomass Energy Systems. To properly identify viable Biomass Energy Systems requires a highly competent team of diverse backgrounds and skills working closely together.

Certain biomass energy systems will emerge as leading candidates for development. Development includes market development, microeconomic and macroeconomic analysis and other assessments in addition to technology development.

Depending upon the state-of-knowledge of the system, the emphasis might be with market and user development, as with wood (forest residues) for direct combustion; with technology development, as with wood conversion to methanol or with applied research, as with certain biological conversion schemes. Thus, the program will have near-term and long-range components.

THE BIOMASS TECHNOLOGY BASE

Although Biomass has been used for centuries for heating, its use as a fuel and chemical feedstock has been largely ignored in the present age of petroleum. Petroleum has spawned the birth of an impressive array of technologies related to its discovery, production, conversion and utilization. Thus, there is an enormous resource of highly trained and experienced scientists and engineers devoting their careers towards advancing the petroleum art. Of course, there is a broadly based commercial sector whose very existence is committed to this technology.

At one time coal sponsored such a following; except in a minor way, Biomass has not, outside of food, fiber and lumber. For Biomass to provide a major contribution to our fuels and chemical needs, such a base must be developed. We believe that the following points should be given important consideration.

- As a resource, Biomass is in competition with fossil resources and will be for many years. Appropriate applications for Biomass must be identified. This must be done from a market need viewpoint rather than force-fitting biomass based technology to the wrong applications.

- A technological and industrial base using biomass feedstock for fuels and chemicals must be developed. Any plan to develop this base must recognize the personal risk and resultant conservatism assumed by corporate executives against breaking new ground. It is important to obtain their commitment, however, in order to initiate the development of a technology base.

- In order to obtain such participation, sufficiently attractive investments must be identified. Initially, this is likely to be achieved by technology which produces petrochemical replacements rather than lower valued fuels. Acetic acid, ethanol, phenol and similar commodity chemicals are potential entries for Biomass. Once commercial participation is obtained, expansion of the technology and of biomass use will occur if the economics justify.

- There is an enormous resource of individuals and small businesses, many of whom are technically unsophisticated but who have the ingenuity, dedication and spirit to develop our biomass resources. These groups need the technical support that an institute like SERI can provide.

SERI's institutional objectives in Biomass are designed to provide a search, identification and development capability to help initiate the widespread utilization of Biomass.

Specifically, SERI is developing the staff and facilities to:

- Conduct a comprehensive analysis of Biomass Energy Systems.

- Conduct an effective research and development program on selected projects.

- Assess economic, social and environmental factors.

- Assist in technology transfer to the private sector.

- Provide program management for certain elements of the DOE Biomass Program.

SERI's Biomass Program Responsibilities can be divided into two separate activities: External Program Management and In-house Biomass Program. Beginning in 1979 and continuing in FY80, SERI has technical program management responsibility for the following elements of the DOE Biomass Energy Systems program: Anaerobic Digestion, Aquatic Production and Fermentation including the cellulose to alcohol fermentation Process Development Unit (PDU). In addition, support will be given to other selected program elements such as near-term wood utilization. To carry out this responsibility, SERI is establishing a separate branch for Biomass Technical Program Management. SERI currently has contracting capability.

SERI's in-house program in FY80 will be approximately 50% committed to Technology Research and Development and 50% to other activities. These other activities include Energy Resource Assessment, Systems Studies, Social Science and Economic Studies, Commercialization Assistance, Market Development and related support activities.

SERI's in-house and R/D capability in Biomass related sciences and engineering will experience significant growth in 1979-80. Our goal is to develop a strong interdisciplinary team in genetics, microbiology, biochemistry, biochemical engineering, process engineering, electrochemistry and certain other disciplines. We plan to conduct basic and applied research and technology development on selected opportunities.

Research objectives will be driven by specific needs identified through analysis of potential Biomass energy systems. In addition to specific objectives an exploratory activity will continue to identify new areas of opportunity. It is our hope that such opportunities will then become a part of the national program.

Some examples of SERI's input to the national effort are as follows. We have issued a significant review entitled: "Photobiological Production of Hydrogen, A Solar Option" by Seibert, Weaver and Lien. A comprehensive survey of gasification technology is in editing and other surveys are being developed. Through such surveys and by developing recognized capability, SERI expects to perform a significant role.

Due to the distributed nature of Biomass, the interfacing with food and forest products sectors, and the potential for significant environmental impacts, a substantial effort in social, environmental and macroeconomic analysis is needed. Such analyses are part of SERI's analysis and assessment role.

Wide use of Biomass will involve new participants not necessarily well versed at this time in the technology or applications. The distributed and regional nature of Biomass will tend to discourage the larger, more sophisticated potential users (other than forest products industries, and certain chemical oriented firms). Thus, many potential Biomass users such as farmers, co-ops, small institutions, etc. will require identification, education, and technical assistance. Such activities are carried out by SERI's Technology Commercialization Division.

As stated earlier, "Biomass" is a complex resource. Successful use of this resource will require a multitude of skills and perspectives. As a society

are involved with major shifts in resource bases
applications of our technology. Success de-
nds on the intelligent integration of many forces.
Without it the potential for failure or even harm
is great; with it the potential for benefit is
significant.

SERI is developing for a lead role in the Biomass arena.

NOTES

Session II A

BIOMASS PRODUCTION BY MARINE AND FRESHWATER PLANTS

John H. Ryther, LaVergne D. Williams, M. Dennis
Hanisak, Richard W. Stenberg and Thomas A. DeBusk
Woods Hole Oceanographic Institution
Woods Hole, Massachusetts 02543
and
Harbor Branch Foundation, Inc.
RFD 1 - Box 196
Fort Pierce, Florida 33450

ABSTRACT

Studies were continued during 1977-78 on the growth and yields in culture of the red seaweed Gracilaria tikvahiae. Partial control of epiphytes was achieved by nutrient removal, shading, and/or biological agents. For the first time, a single clone of the alga was grown continuously throughout the year without replacement. Yields in large (2600 liter) aluminum tanks averaged 21.4 g dry wt/m^2.day, equivalent to 31 tons/acre.year.

Yields of the freshwater macrophytes Lemna minor (common duckweed), Eichhornia crassipes (water hyacinth), and Hydrilla verticillata have also now been measured throughout the year with mean yields of 3.7, 24.2 and 4.2 g dry wt/m^2.day (5.4, 35.3 and 6.1 dry tons/acre.year) respectively. Yields of duckweed and water hyacinths in the cultured units have averaged roughly three times those of the same species growing in highly eutrophic natural environments.

INTRODUCTION

Research has been carried out for the past three years at the Harbor Branch Foundation, Inc. (a nonprofit research organization located on the Indian River between Fort Pierce and Vero Beach, Florida) to develop methods for the artificial cultivation of seaweeds and freshwater weeds, to determine maximum potential biomass yields of such plants (expressed as ash-free dry weight produced per unit of area and time) on a year-round basis in Central Florida, and to investigate the optimal yields that are possible with low cost and non-energy-intensive culture systems that are cost effective both economically and with respect to energy input:output ratio. Methods of estimating the growth and yields of aquatic biomass by monitoring diel nutrient uptake are also under investigation. In addition, because freshwater macrophytes may lose to the atmosphere through evapo-transpiration large quantities of water, which may be in critically short supply in certain areas, the rate of water loss through that mechanism by different species of such ts throughout the year is also in the processes valuation. Finally, because the nutrients ired to produce biomass would otherwise repre- a major cost factor, studies are also in prog- to investigate the feasibility of recycling the liquid residues from the anaerobic digestion of the aquatic biomass back into the culture system as a nutrient source.

SEAWEED CULTIVATION

Seaweed research was initiated with a screening program designed to evaluate growth and biomass production of all macroscopic algal species that could be obtained in adequate quantity in the central Florida area.

The culture system used for the screening was designed and constructed from four 6-meter long, 0.4 m diameter PVC pipes that were longitudinally sectioned and divided into 0.75 m (50 liter) compartments by means of plywood partitions. Each section was provided with a calibrated flow of enriched seawater by means of a manifold fed from a headbox, and also provided with a non-clogging overflow drain. Compressed air was fed into the bottom of each compartment through holes drilled along the bottom of the pipe connecting to an air line (a sectioned one-inch PVC pipe) cemented to the outside of the main pipe. Thirty-two individual growth assay chambers were produced in this way.

The growth chambers were located out-of-doors in full sunlight. Seawater was taken from the Harbor Branch Foundation ship channel which connects to the Indian River, a shallow lagoon of the Atlantic Ocean. No attempt was made to control water temperature, which ranged annually from 12°-34°C, or salinity, which ranged from 20-34‰, in the incoming seawater. Seawater was pumped into a reservoir tank holding several days supply for the experimental chambers. Prior to its use, the stored seawater was enriched with the desired concentrations of nitrogen and phosphorus, normally provided as sodium nitrate and monosodium (dibasic) phosphate at a ratio of 10:1 by atoms of N:P.

Weighed amounts of seaweeds were stocked in the experimental chambers to give the desired density. At intervals of 5-10 days, depending upon growth rate, the algae was removed from the chamber, shaken vigorously to remove water, and weighed. Establishment of the relationship between drained wet weight and dry weight was determined carefully on replicate samples by oven drying at 90°C for 48 hours.

A total of 42 species of seaweeds indigenous to the coastal waters of Central Florida were assessed this way. That number includes six green algae (Chlorophyta), two brown algae (Phaeophyta) and the remainder red algae (Rhodophyta). The latter group included 11 species or varieties of the large genus Gracilaria, of which over 100 species have been described, as well as the closely related Gracilariopsis sjoestedtii.

Of these, the most successful and suitable species to date has been Gracilaria tikvahiae (formerly G. foliifera). Growth of G. tikvahiae in the small, intensively-operated culture system, with strong aeration and over 20 culture-volume exchanges per day of enriched seawater, resulted in biomass yields throughout the year that averaged 34.8 g dry wt/m^2.day (equivalent to 127 dry metric tons/hectare.year, about half of which is organic (1,2,3). Yields were found to be directly proportional to seawater exchange rate, between one and 30 culture volumes/day, for reasons that are apparently not related to nutrient or CO_2 limitation at the slower exchange rates. Maximum yields occur at relatively low nutrient concentrations, 10-100 μmoles N/l as NO_3^- or NH_4^+ and 1-10 μmoles/l $PO_4^{\equiv}$-P together with essential trace metals, and a starting seaweed density of 2-4 kg wet weight/m^2 culture surface area, harvested back to that density every one to two weeks. Growth of Gracilaria occurred throughout the year at Ft. Pierce, Florida, with a maximum mean weekly yield of 46 g dry wt/m^2.day at the end of July and a minimum of 12 g/m^2.day in late January, when the water temperature in the culture fell to 12°C.

During the first two years, epiphytization, the overgrowth of the cultured seaweed with undesirable species of filamentous marine algae, was the major problem in the cultivation of seaweeds. That problem necessitated the periodic discard of the cultures when they became seriously infested, and their replacement with newly-collected material - an expedient that was clearly not feasible in any large-scale culture operation.

Various chemical control methods, including enzymes to prevent the attachment of the epiphytes and algicides that might prove selectively toxic to the epiphytes, were tried with limited or no success. The economic cost-benefits of such treatment, even if successful were judged to be unacceptable in any event.

However, limited but adequate epiphyte control was achieved during the past year (1977-78) with a combination of physical-biological control methods consisting of: (1) temporary (5-10 days) shading of the infested seaweed, (2) temporary (5-10 days) cessation of nutrient enrichment, and (3) introduction of the herbivorous snail Costoanachis avara, which feeds selectively on the epiphytes.

In December, 1977, a specimen of G. tikvahiae was collected from the Indian River near the Harbor Branch Foundation which has subsequently been designated "Oslo Road Clone A" or "ORCA". That clone has since been grown continuously in culture for one complete year, using the epiphyte control methods described above.

In a larger, less energy-intensive culture system than used to obtain the yields described earlier (an aluminum tank of 2.4 m^2 surface area and 2600-liter volume exchanged four times per day), a mean yield over the 12-month growth period of 21 g dry wt/m^2 (76 dry tons/ha.year or 38 ash-free dry tons/ha.year) was obtained (Fig. 1). This has been the first time that a single clone of seaweed has been maintained and grown continuously over an entire year, and this achievement is considered to represent a significant landmark in the seaweed culture efforts of this project.

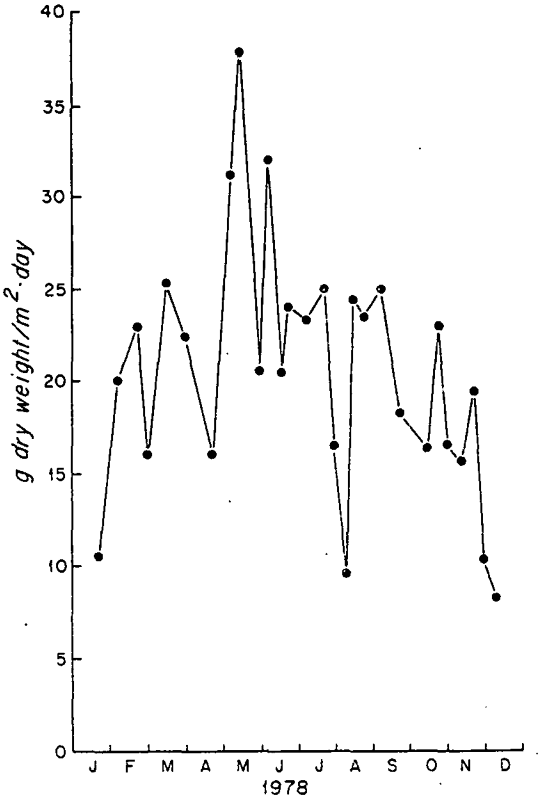

Fig. 1. Mean weekly yields of Gracilaria tikvahiae ORCA clone during 1978.

Attempts have been made, beginning in the summer of 1978, to grow Gracilaria and several other species of seaweeds, including three species of the floating brown alga Sargassum, in floating Vexar-mesh trays suspended in several locations in the Indian River, in Fort Pierce Inlet which connects the Indian River with the Atlantic Ocean, and in the intake canal of the Florida Light and Power Company's Hutchinson Island nuclear power plant. The latter receives its water from an intake line located 2000 feet offshore at a depth of 35 feet e Atlantic Ocean, and the seawater has a much r, more constant salinity and more oceanic roperties and better water quality in general than he more brackish and relatively polluted Indian iver. Furthermore, seawater is pumped through the ntake canal at a rate that provides water exchange hrough the seaweed cultures in the trays of the rder of 10,000 times per day. Nutrient concentra-ions were, however, lower in the intake canal than n the Indian River or in the experimental systems t Harbor Branch Foundation, averaging about 1.0 mole N/l and about 0.3 μmole P/l.

ost of the new species inoculated in the trays ailed to grow at all. *G. tikvahiae* and one spe-ies of *Sargassum* (*S. filipendula*, an estuarine orm), grew initially at moderate to high rates 10-25 g/m^2.day) in the Inlet and in the intake anal, but after two to three weeks, growth de-lined and eventually stopped entirely and the lants became heavily epiphytized and necrotic. cean cage culture of these seaweeds without addi-ional nutrient enrichment, even at very rapid ater exchange rates, therefore does not appear to e feasible.

RESHWATER MACROPHYTE CULTURE

he freshwater macrophytes *Eichhornia crassipes* water hyacinth), *Lemna minor* (common duckweed), nd *Hydrilla verticillata* have now been grown for oughly 18 months at Harbor Branch Foundation. hese plants are grown in 25,000-liter (30 m^2) PVC-lined earthen ponds and/or in 2.2 x 0.8 x 0.2 m oncrete tanks through which enriched well water is assed at exchange rates that have ranged from 0.06 o 2.0 culture volumes per day. Water exchange is pparently much less important with the freshwater lants than with the seaweeds. Not only will they olerate much slower circulation rates in general, ut increasing the exchange rate does not improve yields, except for some slight enhancement with the more rapidly-growing water hyacinths (1).

Initially, only nitrogen and phosphorus were added, at concentrations that varied inversely with the water exchange rates from 100 to 1500 μmoles N/l and from 10 to 150 μmoles P/l. After approximately six months it was found that the most rapidly growing plant cultures (i.e., water hyacinths) became flaccid and chlorotic in appearance, followed by reduction and eventual cessation of growth. Subsequent experiments revealed that these plants required enrichment with nutrients other than N and P, and the culture medium was subsequently supplemented with a commercial trace metal mix[1]. The latter is now used routinely in all freshwater and marine plant culture experiments. Although the addition of trace metals did not affect the growth of the other freshwater plants, that of the water hyacinths significantly increased, so earlier yield data for that species were therefore disregarded and monitoring of the annual growth cycle of water hyacinth was re-started in December, 1977.

Culture experiments with the freshwater plants were similar in most respects to those with the seaweeds except that much less energy-intensive culture methods were used. The cultures were not aerated and, as mentioned above, water was circulated through them much more slowly.

Growth of the plants was measured as the increase in weight per unit area and time, determined by direct weighing. Consistent wet weight values were obtained by draining the plants for specific time intervals. At each weighing, a plant sample was removed, weighed, dried for 58 hours at 60°C and then reweighed providing a ratio of dry matter in the plant to wet or fresh weight. From this relationship, productivity was calculated and expressed as mean grams dry weight/m^2.day for the growth period in question.

The water hyacinths in the culture units were grown in Vexar-mesh cages ranging in size from one to 2.3 m^2. At intervals of approximately one week, the cages with the contained plants were lifted from the water using a rope and hand winch suspended from an A frame over the culture, with a spring scale between the end of the line and the Vexar cage. The culture was allowed to drain for four minutes and weighed. Individual plants were removed to return the population to its starting density (i.e., at the time of the previous weighing and harvest), and the cage was returned to the water.

Duckweed was grown in ponds and vaults with the plants covering the entire water surface. Each week, the plants were netted from the water, transferred in handfuls to a dry container (allowing the excess water to drain), weighed by balance or spring scale, harvested back to their starting density, and returned to the water.

Hydrilla in natural stands is found rooted to the bottom, so it initially appeared impossible to weigh the plants without destroying the culture. After unsuccessful attempts to grow *Hydrilla* in a free-floating fashion, it was discovered that apical sections of the plants would grow when woven through a Vexar-mesh screen suspended above the pond bottom. The Vexar screens and attached plants were then periodically removed, at intervals of one to two weeks, allowed to drain for 10 minutes, weighed, and returned to the water. Yields of *Hydrilla* grown in this fashion were found to be equal to those of plants rooted in both sand and mud (1).

Unlike water hyacinth and duckweed, that can reproduce by budding off new plants, *Hydrilla* grows vegetatively from apical meristems which concentrate near the waters surface in dense mats. New growth is generally harvested by cutting off sections of the individual plants some distance below these growing tips. However, this was found to arrest further growth for periods of 10 days to two weeks until new meristematic growth tips were regenerated. In practice, this meant that harvested (cut) *Hydrilla* did not grow about half the time, a procedure that would seriously underestimate the growth potential of the species. *Hydrilla*

[illegible]niland Nutri-Spray (Chase and Co., Sanford, FL [illegible]71) containing 2% Fe, 0.1% Cu, 0.75% Zn, 0.75% Mn, 0.02% B, 0.01% Mo, 1.5% S.

was therefore not harvested back at all during the year that its growth was monitored, although it was recognized that the culture may have exceeded its optimal density for best growth during part of the year.

In the cultures of water hyacinth and duckweed, on the other hand, it was possible to vary the density of the culture and to return the culture to a constant starting density following each weighing by removing an appropriate number of plants equivalent to the incremental growth since the previous harvest. As had been found earlier with the seaweed (4), specific growth rate (grams increase/gram.day or % increase/day) of the freshwater plants is extremely variable and inversely proportional to the density (g/m^2) of the plant culture. Yield or productivity (g/m^2.day), which is the product of specific growth rate and density, is therefore also a function of density.

High projected yields reported in the literature for both seaweeds (5,6,7) and freshwater plants (8,9,10) that are based on separate measurements of specific growth rate and crop density are therefore invalid and may be seriously misleading.

Figures 2 and 3 show the relationship between specific growth and yield of both duckweed (Lemna minor) and water hyacinth (Eichhornia crassipes) as functions of plant density. In both cases, the same exponential decrease of specific growth occurs with increasing density, resulting in a bell-shaped curve of yield as a function of density. The latter results from the fact that at low densities all of the incident solar radiation per unit area is not absorbed by chloroplasts, while at high densities, above that at which all of the light is absorbed by the plants, no further photosynthesis is possible but respiration continues to increase with increasing density leading to a decreasing net plant production (photosynthesis-respiration).

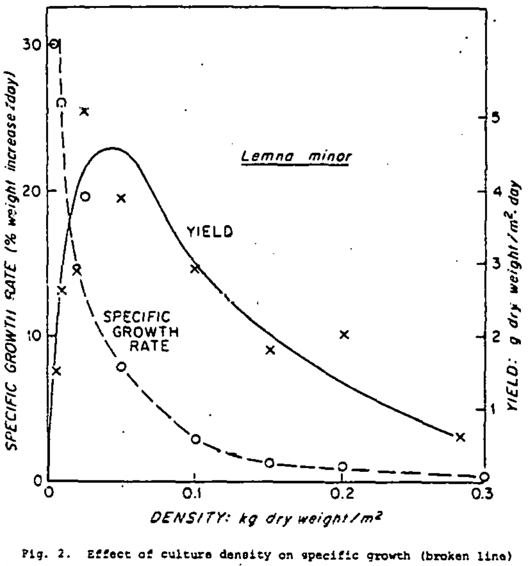

Fig. 2. Effect of culture density on specific growth (broken line) and yield (solid line) of Lemna minor.

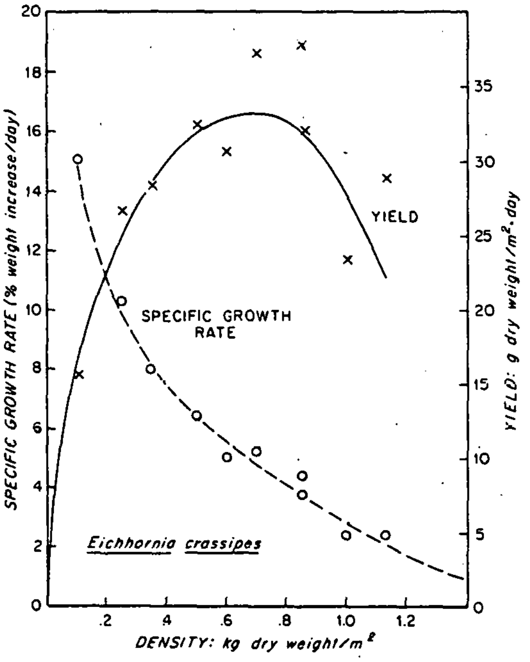

Fig. 3. Effect of culture density on specific growth rate (broken line) and yield (solid line) of Eichhornia crassipes.

Specific growth of Lemna was twice that of Eichhornia at low densities, but the exponential decay is extremely rapid at increasing densities resulting in a low peak productivity at the optimum density of 0.02 kg dry wt/m^2 (0.2 kg wet wt/m^2). In Eichhornia, on the other hand, the decrease in specific growth was slower and peak yields occurred at densities ranging from about 0.5 to 1.0 kg dry wt/m^2 (10-20 kg wet wt/m^2), with resulting yields 6-7 times those of Lemna.

The difference in productivity between the duckweeds and the water hyacinth is probably one of plant form and habit. Duckweeds are essentially two-dimensional organisms. When they have completely covered the water surface, which occurs at a very low plant density, they cannot grow further without overlapping and self-shading. There is no way that they can expose more photosynthetic tissue per unit of area to sunlight. A possible solution to that problem might be to remove the incremental growth as quickly as it is produced. However, increasing the harvest rate from one week to three days did not appear to improve yields in our cultures, though Culley et al. (11) has reported a significant increase in the production of duckweeds by harvesting its new growth daily.

Hyacinths, on the other hand, grow as high as one meter above the water surface, providing a high leaf area index (12) which enables the plants to intercept solar radiation efficiently in dense stands. Thus, the plant is capable of high specific growth rates at much higher densities than is the case with the duckweeds.

ring the establishment of what appeared to be l culture conditions with respect to nutrient a, nutrient concentration, water exchange rate, e density, and general method of growing the ants, the three species of aquatic macrophytes re grown for an entire year, weighing and, with o of the three species, harvesting the cultures ck to a constant starting density, every one to veral weeks (depending upon season and growth te) throughout the year.

nual yields of duckweed, water hyacinths, and drilla are shown in Fig. 4, each point representg mean daily productivity over periods of one to veral weeks (depending upon season and growth te). The mean annual productivity of water yacinth was 24.2 g/m^2.day, with a range of 5.3 to 4.9 g/m^2·day. Water hyacinth is by far the most roductive, trouble-free, and generally the most uccessful and suitable freshwater macrophyte grown o date as a candidate species for a biomass plantation. Its mean annual production of 24.2 dry wt /m^2.day is equivalent to 88 dry metric tons/ha. ear or 35 dry tons/acre.year, 82% of which (72 ons/ha.year or 28 tons/acre.year) is organic mater, a yield greater than any other photosynthetic rop reported in the literature (e.g., 13), and omewhat greater than that reported above for racilaria grown in the small, intensive culture ystem (63.5 ash-free tons/ha.year or 25.4 tons/ cre.year.

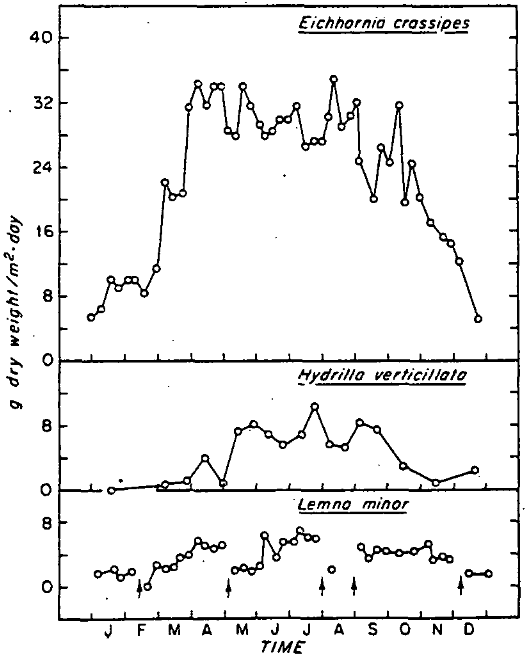

Fig. 4. Mean daily yields of Eichhornia crassipes, Hydrilla verticillata and Lemna minor throughout the year at Fort Pierce, Florida.

he productivity of duckweed for the year showed a ange of 0.1 to 7.0 g/m^2.day and an annual mean of .7 g/m^2.day, equivalent to 13.5 dry metric tons/ ear or 5.4 tons/acre.year. However, the duck- cultures were not continually maintained; at four times during the year, the cultures were overgrown by the filamentous alga Hydrodictyon, necessitating replacement of the culture with additional plants. The culture also had to be replaced once because heavy winds literally blew the plants out of the pond. (See arrows, Fig.4 for replacement times.)

Because of the relatively poor growth of the common duckweed, Lemna minor, its growth in the concrete tanks was compared with that of the related giant duckweed, Spirodela polyrhiza, in an identical culture unit and grown under the same conditions over a period of several weeks. Yields of the two species were almost identical (Table 1) so further work with the latter was not pursued.

Table 1. Yields of Lemna minor (common duckweed) and Spirodela polyrhiza (giant duckweed) in concrete tanks at a water exchange rate (residence time) of 0.5 days.

Dates (1978)	Mean yield (dry wt/m^2.day) L. minor	S. polyrhiza
1/17-1/25	2.4	3.3
1/25-2/1	2.3	1.5
2/1-2/8	1.9	1.8
2/8-2/21	2.0	2.3
2/21-3/1	2.2	3.4
3/1-3/14	1.8	3.2
3/14-3/28	6.1	5.8
3/28-4/14	3.6	3.8
4/14-4/24	7.9	5.9
Mean (1/17-4/24)	3.3	3.4

The mean annual yield of Hydrilla was 4.2 g dry wt/m^2.day (15.3 dry metric tons/ha.year or 6.1 tons/acre.year) slightly better than duckweed but far from that of water hyacinths.

Both duckweed and water hyacinth have their photosynthetic and gas exchange mechanisms exposed to the air and are not dependent upon the CO_2 dissolved in freshwater, a factor that may limit photosynthesis and growth in many submerged freshwater plants including Hydrilla.

Another major problem with Hydrilla grown at the latitude of central Florida is that plants flower in the fall, begin to store organic matter in root tubers, and virtually cease vegetative growth throughout the winter. Our experiments demonstrated that some growth would occur in winter if water temperatures were maintained at an elevated temperature of 25°-30°C, but for practical purposes Hydrilla must be considered as a seasonal crop in most if not all of mainland United States.

Finally there is the problem, discussed earlier, that Hydrilla must be harvested by cutting the plant below its terminal meristematic tissue, setting back further growth until new growing tissue is regenerated.

Experiments were initiated in July, 1978 with the cultivation of the pennywort (Hydrocotyle umbellata). This is an emergent plant similar in its habit and appearance to water hyacinth, and it has been grown in Vexar-mesh baskets immersed in the concrete tanks exactly as with the hyacinths. Yields from July 3 through November 14, 1978 averaged 15.9 g dry wt/m^2.day (Table 2). During the time it has been cultured, pennywort has been found to be considerably more productive than the other freshwater macrophytes tested with the exception of water hyacinths.

Table 2. Yields of pennywort (Hydrocotyle umbellate) grown in concrete tanks at a water exchange rate (residence time) of 0.5 days with influent NO_3-N of 50 µmoles/1 and $PO_4^{=}$-P of 5 µmoles/1.

Dates	Mean yield* (g dry wt/m^2.day)
7/3-7/14/78	9.3
7/14-7/21	14.6
7/21-7/28	21.6
7/28-8/11	16.9
8/11-8/18	20.8
8/18-8/25	16.7
8/25-9/1	15.4
9/1-9/6	29.7
9/6-9/15	22.3
9/15-9/22	9.7
9/22-10/6	5.4
10/6-10/17	4.6
10/17-10/25	12.7
10/25-10/31	23.2
10/31-11/14	15.4
Mean (7/3-11/14/78)	15.9

*Constant starting density not maintained.

Its advantage may lie in the fact that it is reputedly more cold tolerant than the tropical water hyacinth and may therefore be adaptable to a more temperate climate. Conceivably, through better growth performance in winter, it could have a better annual yield than hyacinths in many parts of the United States. It will be grown for an entire year at Harbor Branch Foundation to evaluate its annual yield. A disadvantage of pennywort is the fact that it does not bud off separate plants, as does the water hyacinth and the duckweeds, but the new plants remain firmly attached to the parent stock through an intricately connected and interwoven root system. Harvesting therefore would require virtually cutting the plants apart, a practice that has not yet been attempted in the culture of the species to date.

There being little or no available data on the natural productivity of freshwater macrophytes, there was no way to assess the significance of the yield data that was being obtained in the artificial culture systems that had been developed at the Harbor Branch Foundation. For that reason, measurements were begun in the spring of 1978 of the yields of natural populations of water hyacinth and duckweed, comparing these with yields of the same species taken from the same populations and transplanted i the Harbor Branch Foundation culture systems.

In both cases, the natural populations occurred in areas that were heavily enriched from the run-off from agricultural and cattle rearing operations.

A natural stand of water hyacinths growing in a fire ditch adjacent to the Kissimmee River near Okeechobee, Florida, and a pond containing Lemna minor north of Ft. Pierce, Florida, were the sites chosen for field studies. Water hyacinths from the fire ditch location were collected and placed in 1 m^2 Vexar cages. Three such cages were left with the natural population in the fire ditch, and another three were returned to the HBF and placed in a pond through which enriched sewage was passed with a residence time of 0.5 days. Lemna was collected from the pond north of Ft. Pierce and placed in 1 m^2 PVC enclosures having nylon screened sides. Three of these enclosures were left in the field, and three were returned to the HBF and cultured in a fashion similar to that of the hyacinths. Initially, growth of the two (natural and laboratory) populations were roughly the same for each species, but during summer and fall yields of the natural populations of both duckweed and hyacinths fell to about one-third that observed at the laboratory ponds, presumably due to the lack of controlled enrichment and water exchange in the field (Fig. 5). Mean yields of cultured vs. natural stands of the duckweeds from 5/11-11/3/78 were 4.8 and 1.4 g dry wt/m^2.day and for cultured vs. natural stands of water hyacinths from 3/30-11/3/78, 36.3 and 10.4 g dry wt/m^2.day.

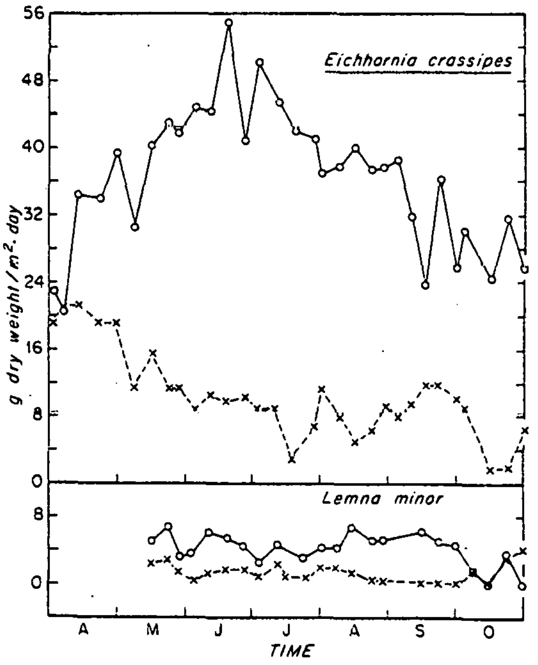

Fig. 5. Mean daily yields of Eichhornia crassipes and Lemna minor in artificial cultures at Harbor Branch Foundation (solid lines) and in natural population in the field (broken lines) near Fort Pierce, Florida.

: studies were continued so as to obtain data ın entire year, but were not available at the of this writing. The results for the year uggest that production of the freshwater macrohytes observed at the Harbor Branch Foundation ulture facility may closely approach the maximum otential yield for the species.

STIMATING GROWTH THROUGH NUTRIENT UPTAKE

lthough it is possible to monitor the growth of quatic macrophytes in small, experimental systems y weighing the entire plant populations, such a rocedure would be obviously impossible in any arge-scale biomass farm. An alternative and more ractical approach for larger units would be to onitor the uptake of some essential plant nutrient nd relate that to biomass production. Such a ethod is possible if the nutrient is assimilated t a relatively constant rate, represents a constant relationship to total biomass, and is neither ost to nor gained in the water by any processes ther than plant assimilation. Nitrate would ppear to satisfy these requirements if no other orm of nitrogen is presented to the plant cultures. xperiments were therefore carried out in which ater hyacinths were grown in concrete tanks hrough which enriched well water containing aproximately 50 µmoles/l of nitrate as a sole nitroen source was circulated at two volume exchanges er day.

he stocking density of water hyacinths was usually 0 kg wet wt/m^2. At approximately weekly intervals, he cages containing the water hyacinths were lifted ut of the tanks, drained for four minutes, weighed ith a spring scale, and, after removing the increental growth, returned to the tanks. Primary prouctivity by this harvest technique was calculated s the increase in dry plant weight per unit area nd time. The dry weight of *Eichhornia* was conidered to be 5% of its wet weight.

n order to estimate primary productivity from easurements of nitrate removal, the effluent from ne of these tanks was analyzed for its NO_3-N conent in a continuous basis over a period of 24 ours with an automated nutrient analyzer system. n Autoanalyzer Model I proportionating pump pumped amples from the effluent of the tank through an n-line reagent filter and then through a copper-admium column which reduced the NO_3-N in the samle to NO_2-N. The resulting solution was pumped hrough a flow-through cell in a Bausch and Lomb pectronic 100 spectrophotometer which transmitted n output signal that was recorded on a OmniScribe 3-5000 recorder. The entire apparatus was enclosed n a wooden box (1.28 x 0.65 x 0.50 m) which could e readily moved around from one sampling location o another. Input and output tubing went through n opening in the box by means of a PVC pipe (2.54 m in diameter) that was fitted with an elbow to revent precipitation from entering into the apparatus. A 60 watt light bulb was kept on inside the ox to facilitate monitoring the apparatus at night [illegible] also to aid in humidity control. The tank was ted with compressed air through holes drilled a PVC pipe (2.54 cm in diameter) placed on the bottom of the tank in order to insure complete mixing within the tank and to prevent any time lag in observing changes in NO_3-N concentration. Although it was not necessary, a new cadmium column was prepared for each run. Standards, reagent blanks, and influent NO_3-N levels were monitored at the beginning and end of each diel run. For each run, a composite sample was made from three plants for an analysis of the internal nitrogen content of the plants with a Perkin-Elmer Model 240 Elemental Analyzer.

From the continuous 24-hour record of the effluent NO_3-N concentration, primary productivity, based on diel nitrate uptake by the plants was calculated from the equation:

$$P = \frac{1.4 \times 10^{-5}(C_i - C_o)\ (V)\ \ (T)}{(A)\ (N)} \tag{1}$$

where C_i = the NO_3-N concentration (µM) in the influent into the concrete tank, C_o = NO_3-N concentration (µM) in the effluent from the tank, V = the number of volumes (liters) per turn-over, T = the number of turnovers per day, A = the area occupied by the cage of plants, N = the internal nitrogen content of the plants (g N/g dry weight), 1.4×10^{-5} converts µmole NO_3-N to g N. The differences between influent and effluent NO_3-N levels were obtained by integrating the area under the diel curves and determining the amount of NO_3-N removed from the water over 24 hours. All diel graphs were plotted on a scale from 0000 to 2400 to facilitate comparison of experiments that began at different times of the day.

Eleven diel experiments were conducted from January to July 1978 for *Eichhornia crassipes*. An example is illustrated in Fig. 6. Based on these data, there was little evidence of a diel cycle in NO_3-N uptake.

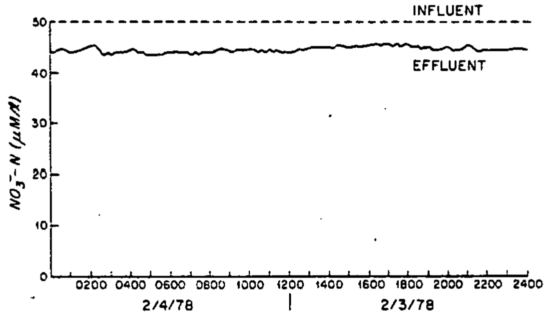

Fig. 6. Diel pattern of effluent nitrate-nitrogen concentration from water hyacinth culture. Influent concentration shown as broken line.

From these data, primary productivity rates were estimated using equation (1). These estimates (Table 3) agreed favorably with those obtained with the harvest method, being on the average 12.33% less. Interestingly, the best agreement between the two methods was in July when productivity was the highest of the study. At that time, estimates of primary productivity from diel NO_3-N uptake measurements were slightly higher than those made with the harvest method. The worst agreement was on April 6-7, 1978, which was an exceptionally

Table 3. Comparison of estimates of primary productivity for Eichhornia crassipes as calculated from measurements of diel NO_3-N uptake (P_D) and from direct harvesting (P_H).

Dates of diel 1978	Primary productivity (g dry weight.m^{-2}.day^{-1}) P_D	 P_H	 P_D/P_H
1/5-1/6	7.00	7.26	96.42
1/7-1/8	5.75	7.26	79.20
1/24-1/25	7.01	7.47	93.84
2/3-2/4	3.18	4.85	65.57
4/3-4/4	12.24	16.77	72.99
4/4-4/5	12.84	16.77	76.57
4/5-4/6	13.47	16.77	80.32
4/6-4/7	5.89	16.77	35.12
7/14-7/15	29.49	28.01	105.28
7/15-7/16	28.32	28.01	101.11
7/27-7/28	36.57	34.37	106.40
Average	14.71	16.76	87.77

inclement day. That day was the last of four consecutive days in which diel studies were conducted, the first three of which demonstrated good day-to-day reproductivity although the individual diel patterns did vary somewhat.

WATER LOSS THROUGH EVAPO-TRANSPIRATION

A major constraint to the possible use of freshwater macrophytes in a large-scale energy plantation could be the consumption of water, in short supply over much of the earth's surface. It was found earlier that the freshwater plants, unlike the seaweeds do not require a rapid exchange of water to insure their maximum growth. Experiments had not previously been conducted, however, to determine the relative growth of these plants in completely stagnant cultures. Such an experiment was therefore begun in July 1978.

Both duckweed (Lemna minor) and water hyacinths (Eichhornia crassipes) and (Hydrilla verticillata) were grown in 1000-l polyethylene cylinders that received no flow of water but were enriched once a week with a concentrated solution that provided the same weekly ration of N, P, and trace elements as did the flow-through systems described above. The weeds were weighed weekly and harvested back to their starting densities.

Growth in the stagnant culture was initially poor, but after modification of the enrichment procedure involving spraying the trace element mix onto the emergent hyacinths and duckweeds, they approached those obtained at the same time in the flow-through cultures.

Water loss through evapo-transpiration (straight evaporation in the submerged Hydrilla culture) was then monitored, adding weekly the measured amount lost the previous week, and the losses were compared to that of a control cylinder containing no plants (Table 4).

Table 4. Water loss from cultures of Eichhornia crassipes, Lemna minor, and Hydrilla verticillata relative to water loss from a control container of water (plant culture/control) and related meteorological data.

Dates (1978)	Eichhornia	Lemna	Hydrilla	Total Rainfall (mm)	Mean Sunlight (langleys/d.)	Air Temperature (°C) Mean	Max	Min
7/18-7/25	1.72	1.04	--	62.7	553	29	34	24
7/25-8/1	1.24	0.97	--	91.6	417	27	33	21
8/1-8/8	1.60	1.04	1.20	63.2	499	28	33	23
8/8-8/14	1.49	0.99	1.06	57.2	551	28	34	22
8/14-8/22	1.35	0.86	1.06	0	536	29	32	26
8/22-8/29	1.77	0.98	1.00	36.8	567	28	34	22
8/29-9/5	1.31	1.03	1.30	2.5	488	27	33	22
9/5-9/12	1.61	0.92	1.10	7.6	422	28	34	22
9/12-9/19	1.27	0.92	1.16	6.3	434	27	33	22
9/19-9/26	--	--	--	15.7	409	27	33	22
9/26-10/4	--	--	--	21.6	389	27	33	22
10/4-10/11	1.78	--	1.10	27.9	375	26	--	--
10/11-10/18	2.58	0.93	1.37	53.3	308	24	30	16
10/18-10/25	1.52	0.88	1.03	27.9	343	22	27	17
10/25-10/31	2.55	--	1.33	55.9	388	25	30	21
10/31-11/7	1.39	1.18	0.97	39.4	380	21	28	11
11/7-11/14	--	--	--	91.9	--	24	29	18
11/14-11/21	1.78	0.92	1.11	0.8	--	24	29	18
11/21-11/28	1.51	0.87	1.20	0	--	22	29	14
11/28-12/5	2.26	1.15	1.76	10.2	--	23	31	14
Mean	1.69	0.98	1.18					

Of the three species, water hyacinth lost the most water, as would be expected from its greater exposure to the air and relatively greater metabolic activity. Duckweed behaved essentially like water alone. The slightly greater loss in the Hydrilla culture compared to that of the control probably resulted from greater light absorption and/or restriction of vertical convective circulation in the dense plant culture. Little correlation could be seen visually between water loss and productivity of the plants or between water loss and the various meteorological factors measured. However, the data have not been statistically analyzed.

Evapo-transpiration loss from the water hyacinths has been surprisingly low - 1.69 times on the average and 2.58 times at most the loss from evaporation alone. As explained earlier, however, the hyacinths were not growing well in the stagnant culture during the first part of the experiment and, by the time the problem had been corrected, the season for the best growth of the species had passed. The experiment is therefore being continued so as to obtain valid data for water loss of water hyacinth throughout the year at the Harbor Branch Foundation location, and relevant meteorological data will be collected at the same time in an attempt to identify the factor(s) that control water balance in the species.

ELEMENTAL COMPOSITION OF FRESHWATER MACROPHYTES

All of the freshwater macrophytes cultured at Harbor Branch Foundation and in the field are routinely analyzed for their content of ash, volatile solids, carbon and nitrogen. Mean compositions of six species are shown in Table 5. Note hat the percent volatile solids in the enriched atory cultures of duckweed and *Hydrilla* is ficantly lower than in the natural stands, hyacinths are the same at both locations, nd cultured pennywort (*Hydrocotyle*) has a higher olatile solids content than the natural popula-ion, though the latter is based on only two amples.

Table 5. Percent of ash, volatile solids, carbon and nitrogen in freshwater macrophytes grown at Harbor Branch Foundation (HBF) and in natural populations of the same species.

Species	Ash		Volatile Solids		Carbon		Nitrogen	
	HBF	Field	HBF	Field	HBF	Field	HBF	Field
Eichhornia crassipes	19	19	81	81	35	35	2.2	1.5
Lemna minor	33	22	67	78	31	34	1.8	2.0
Hydrocotyle umbellata	15	22	85	78	39	39	2.1	2.4
Hydrilla verticillata	28	19	72	81	37	37	3.4	3.6
Salvinia rotundifolia	37	28	63	72	31	34	1.9	2.4
Azolla caroliniana	13	--	87	--	42	--	3.0	--

ECYCLING DIGESTER RESIDUES

ive 125-liter digesters measuring approximately 5 x 45 x 80 cm were constructed from 0.6 cm sheet lastic. Filling and emptying ports were made of 3 cm PVC pipe with screw-cap ends and two smaller 2.5 cm PVC pipes with valves were also provided or removal of the liquid residues. One side of ach digester contained a plexiglass window for isual observation. Gas lines led from the tops of he digesters to inverted, submerged 50-gallon drum anometers, where the gas was collected and its olume monitored.

wo of the digesters have been used for water hya-inth fermentation, two for *Gracilaria* fermentation, nd the fifth kept in reserve in case of failure f any of the other four cultures. Operation of he digesters was carried out at ambient tempera-ure during the summer and early fall. Beginning n November, 1978, the digesters have been kept artially submerged in a large circular water tank, he water of which has been kept at 30°C by an im-ersion heater.

ater hyacinths

he water hyacinth digesters were started with fer-enting dairy manure; the manure was gradually re-laced with water hyacinths. The studies herein escribed were initiated 4 months after the initial tartup of the digesters.

ne water hyacinths require fine shredding or chop-ing prior to digestion. A Sears-Roebuck electric ard and garden shredder is used for that purpose, roducing a greenish-black slurry that has the con-istency of thick mud. The water hyacinths must be hopped immediately after harvesting and removal rom the water, as even partial drying renders the lant material tough and fibrous and resistent to hredding, and soaking in water does not reconsti-ute the plant flesh to a suitable form for chopping.

ne hyacinth digesters were loaded three times a at 0.8-1.0 g volatile solids/liter digester volume (2.4-3.0 kg chopped hyacinths/digester/day). At the time they were loaded, an equivalent volume of liquid residue has removed from the digester. The digester contents were not stirred, and the solid fraction floated at the surface above the liquid fraction. To prevent the solids from plugging the discharge port, a screen was inserted above which the solids were trapped. Stratification of the solid phase was, however, probably a deterrent to more complete digestion of the organic matter and some change in design, perhaps including gentle agitation, would appear desirable.

The water hyacinth digesters have now been in operation for over eight months, producing on the average 0.4 l gas/g volatile solids (24 g wet wt plant material), at 60% methane. The heat of combustion of water hyacinth is 3.8 Cal/g dry weight (14), or 4.6 Cal/g ash-free dry weight (volatile solids), as hyacinths are 17% ash (15). Assuming that of pure methane to be 8.8 Cal/liter (1000 BTU/SCF), the efficiency of conversion has averaged 46% (0.4 l/g x 0.6 l methane/l gas x 8.8/4.6). The pH has been maintained at 7.0 to 7.3. Higher digester loadings were found to depress pH and reduce methane content of the gas.

The liquid residue withdrawn from the hyacinth digesters contained, on the average, approximately 430 mg nitrogen and 33 mg phosphorus per liter. Roughly half the nitrogen is NH_4^+-N, the remaining half is dissolved organic nitrogen of unknown identity.

To investigate the suitability of the digester residue as a nutrient source for growing water hyacinths, three cultures of the plants, maintained at a density of 10 kg wet wt/m^2, were established in 750 l, 2.28 x 78 x 64 cm (1.8 m^2) concrete tanks. Two of the three cultures were operated in a "batch" mode with one complete exchange of well water per week. One of these cultures received no enrichment. The second received enrichment with the nutrient medium normally used to grow water hyacinths. Approximately half the nitrogen and phosphorus was assimilated each week from the enriched culture. However, after one month, both of the above cultures became chlorotic and unhealthy in appearance, so both were thereafter sprayed weekly with additional trace-element mixture which restored them to a normal appearance.

The third culture received five liters of liquid residue from the water hyacinth digester three times a week, a total of approximately 6450 mg N and 495 mg P per week. The water in that culture was not exchanged nor did the plants receive the trace-element spray or any other form of enrichment.

The water hyacinths were contained in a Vexar-mesh basket in the tanks and the basket containing the entire culture was hauled out of the concrete tanks once a week, allowed to drain for four minutes, and weighed. Incremental growth was removed from the culture which was then returned to the water. Growth, expressed as mean daily yield of dry wt/m^2, was calculated for each one-week interval assuming dry weight = 5% of wet weight. Data for the period August 11-December 28, 1978 indicate that growth in the digester residue has averaged 54% more than in the chemically-enriched culture and four times that of the unenriched control (Table 6).

Table 6. Yields of water hyacinth (Eichhornia crassipes) grown in unenriched water, in chemically-defined enrichment medium, and in the liquid residue from anaerobic digestion of water hyacinths.

Dates 1978	Mean yield (g dry weight/m^2.day) Unenriched	Chemical medium	Digester residue
Aug 11-17	27	33	40
Aug 17-25	17	28	30
Aug 25-Sept 1	6	10	22
Sept 1-8	12	9	21
Sept 8-15	8*	11*	24
Sept 15-22	7	24	30
Sept 22-29	5	14	22
Sept 29-Oct 6	2	21	23
Oct 6-13	4	8	17
Oct 13-20	5	15	20
Oct 20-27	1	7	14
Oct 27-Nov 3	< 1	19	17
Nov 3-10	6	20	23
Nov 10-17	< 1	9	18
Nov 17-24	1	11	12
Nov 24-Dec 1	0	18	20
Dec 1-8	0	5	11
Dec 8-15	0	3	9
Dec 15-21	0	1	9
Dec 21-28	0	2	11
Mean	5	13	20

*Began spraying foliage with trace-element mix.

An approximate balance of the nitrogen recycled through the culture-digestion-culture system was made. Over the 20-week experimental period, the digester was loaded with a total of 235 kg wet weight of water hyacinths. This biomass is 11.75 kg dry weight and contained 255 g N. During that same time period, a total of 285 l of liquid effluent was removed containing 430 mg N/l or 123 g N. Addition of that effluent to cultures produced 2800 g dry weight of water hyacinths containing 80 g N.

In summary, of the 255 g nitrogen loaded into the digester, 123 g (48%) were recovered in the liquid effluent, of which 80 g (65%) were reassimilated by the water hyacinths, an overall efficiency of 31%. Most of the remaining nitrogen loaded into the digester appears to be in the undigested, solid residue. This solid residue is a potential source of nutrients, being rich in nitrogen (approximately 4% on a dry weight basis). Preliminary results indicate that this solid residue can serve as a slow-release fertilizer for cultures of water hyacinths. By using the residue as nutrient source and by increasing the rate of digestion, the recycling efficiency of the present system could be significantly enhanced.

Gracilaria

Several attempts were made to convert an anaerobic digestion substrate from dairy manure to Gracilaria, but these were all unsuccessful. Extremely low pH (< 5) was reached and maintained in the digestion mixture, indicating the presence of persistent organic acids that presumably destroyed the relatively sensitive methanogenic bacteria. The conclusion was reached that successful digestion of Gracilaria would require a bacterial culture acclimated to the marine environment and to the tissues of marine organisms. Accordingly, a new attempt was made, using highly organic, anaerobic marine sediment collected from an area where Gracilaria and other seaweeds had been observed to accumulate on the bottom.

The 125-liter digester was loaded with 20 kg of wet sediment, 5 kg of fresh Gracilaria, and 80 liters of seawater. The digestion mixture was maintained at approximately 30°C and received no agitation. Within two days, the pH dropped from 7.4 to 6.6 and thereafter remained in the 6.3-6.6 range. After six days, gas evolution began, and on the seventh day, the evolved gas became combustible, indicating the presence of methane and the commencement of complete digestion.

A major advantage in the digestion of Gracilaria is that the seaweeds do not need to be chopped, shredded, or otherwise processed prior to loading - a fcctor tht could substantially reduce operating costs relative to water hyacinths and other species that do require such processing.

Two Gracilaria digesters have now been in operation for approximately two months, normally loaded three times a week at the rate of 0.75 g volatile solids per liter of digester volume (ca. 2.0 kg wet wt Gracilaria/digester) per day. Gas production was in the range of 0.2-0.4 liters/g volatile solids at 60% methane. Assuming a heat of combustion of 4.5 Cal/g ash-free dry wt (16), the conversion efficiency (or volatile solids reduction) ranged from 23 to 36%. The Gracilaria digesters appear to be very sensitive to temperature, particularly below 25°C. However, gas production in the seaweed digester has improved steadily since successful digestion commenced and was exceeding that of the water hyacinths (in liters/g volatile solids) at the time of this writing.

In the only other published information on seaweed digestion (17), reported gas yields containing up to 63.4% methane ranging from 0.29 to 0.53 liters/g volatile solids and with a conversion efficiency (i.e., volatile solids reduction) ranging from 29.7 to 56.8%. The seaweed, kelp (Macrocystis pyrifera), was chopped or powdered, dried, and frozen prior to digestion and the latter took place at 35°C. However, the results were very similar to those reported above for unprocessed Gracilaria.

Because of the initial difficulties in fermenting Gracilaria, experiments have only recently been started on recycling the digester residues as a nutrient source for growing the seaweeds. Cultures have been started in 50-liter containers (sectioned 0.4 dia. PVC pipes), that are aerated to maintain the seaweed in suspension. There is no flow of seaweed through the cultures, but the water is exchanged once a week. One culture is enriched with 1500 μmoles/l NO_3-N, 150 μmoles/l $PO_4^{\equiv}$-P, and 5 ml/l trace element mix (i.e., the seawater enrichment normally used to grow Gracilaria). The other culture receives one liter of liquid Gracilaria digester effluent three times per week. The analysis of the Gracilaria digester residue is not yet available, so comparable nutrient loading of the two cultures is not yet possible.

- ³-te, the mean yield of *Gracilaria* in the chem-
y-enriched seawater culture was 2.5 g dry wt/
y, that in the digester-residue enriched sea-
was 7.1 g dry wt/m^2.day, almost three times
s great. That experiment will be continued and
ne recycling efficiency determined, as in the
ater hyacinth experiment described above.

UTURE RESEARCH

uture research will be largely restricted to the
reshwater macrophyte, water hyacinth (*Eichhornia*
rassipes) and the red seaweed *Gracilaria tikvahiae*.
xperiments will be carried out at three levels or
cales of operation.

A) Small (50-500 l) scale studies of the growth,
utrient uptake and other aspects of the physiology
f the two aquatic plant species so as to obtain a
etter understanding of their biomass production and
ne environmental factors that limit and control
roduction, to understand and control problems in
neir mass artificial cultivation and to help maxi-
ize the output:input ratio of mass culture systems
ith respect to both economic and energy criteria.

3) Meso-scale (20,000 l) pond experiments to
nvestigate different low-cost, non-energy-inten-
ive culture methods specifically, for the seaweed
racilaria with respect to yield, performance and
eliability throughout the year, and relative free-
om from epiphytization and other chronic problems
f mass seaweed culture. These experiments will
ave the objective of developing and selecting one
r more methods for the larger, pilot-scale demon-
tration experiments.

C) Large-scale (1/4 acre, 850,000 l) pond cultures
f both *Gracilaria* and water hyacinths to demon-
trate feasibility and performance of mass culture
ystems, to determine annual yields under mass cul-
ure conditions, and to permit realistic cost:bene-
it analysis with respect to both energy and
conomics.

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NOTES

BIOENGINEERING ASPECTS OF INORGANIC CARBON SUPPLY TO MASS ALGAL CULTURES

Joel C. Goldman
Woods Hole Oceanographic Institution
Woods Hole, Massachusetts 02543

ABSTRACT

The study involves a thorough investigation of the inorganic carbon requirements of algae (both micro and macro species) under mass culture conditions. Major research topics considered include: 1) the growth kinetics of various freshwater and marine algae under inorganic limitation at different pH levels; 2) comparisons of growth with bubbled CO_2 and HCO_3^- alkalinity as inorganic carbon sources at varying pH levels; 3) effects of pH on algal growth under non-carbon limiting conditions; 4) effects of mixing in inorganic carbon limiting and non-limiting situations on algal growth; 5) growth kinetics under carbon saturation but light limitation. Based on results to date, it appears that certain marine microalgae appear to be far more tolerant of high pH (> 9) than other species. HCO_3^-, if supplied in excess and with proper pH control, can meet the inorganic carbon requirements of marine algae. Efficiency of use is 90% at HCO_3^- concentrations four times greater than in natural seawater. Based on the growth kinetics of fresh water green algae under HCO_3^- limitation, virtually 100% assimilation is possible at all growth rates up to just before cell washout. Hence, maximum yields under carbon limitation occur at very high growth rates. In addition, it has been shown that the supply and mass transfer rates of gaseous carbon dioxide rather than the concentration of carbon dioxide in the gas mixture is the critical factor in ensuring an adequate supply of inorganic carbon.

BACKGROUND

Among the many alternative energy sources being considered to meet future demands, the photosynthetic conversion of radiant to useful chemical energy (bioconversion) is receiving widespread attention [15]. Clearly, all energy on earth has evolved, in one way or another, from the ultimate source, solar radiation, primarily through the storage of fossil fuels over geologic time. Now the rate of energy consumption far exceeds the rate of supply; the prime determinant, then, in deciding the attractiveness of a new alternative source is the rate at which it can be produced. This implies that an energy balance for a particular process is favorable; i.e., the amount of useful energy produced exceeds the energy expended during production.

The photosynthetic production of energy has particular appeal because it is the most basic of energy-storing and life-supporting processes. One of the major problems, however, is that solar energy, although virtually infinite in total capacity, strikes the earth at a very low flux, about 1.6 or less gr cal m^{-2} $min.^{-1}$, hence requiring very large collection systems for capturing the required energy. The other major problem is that photosynthetic conversion efficiences are very low, and under the most ideal conditions the most efficient plants can convert at best about 10-12% of visible solar radiation to stored chemical energy as organic matter [19]. In reality, photosynthetic conversion efficiences of natural terrestrial and aquatic systems are considerably lower, and seldom exceed 1-2%, primarily because other factors such as light availability, nutrients, water, etc. are limiting. In fact, it is estimated that only about 0.1% of the solar energy striking the earth is converted to organically bound energy in the form of plant material [15].

Therefore, it is the prime goal of any bioconversion scheme to maximize possible photosynthetic efficiencies and resulting yields by forcing the only uncontrollable growth factor, light, to be limiting. This means that all the required nutrients for growth must be supplied in excess.

Aquatic plants, primarily micro- and macroalgae, are among the most efficient converters of radient energy, and conversion efficiencies under laboratory conditions with low incident radiation have been reported to be ca. 20% [21]. In addition, algal cultures can, in theory, be maintained indefinitely and thus are not dependent on seasonal growth. For these reasons mass algal cultures are being considered as candidate bioconversion systems in the DOE Biomass Energy Systems Program [6, 7].

The potential applications for algal cultures are widespread, as seen in Fig. 1. Early interest in algal mass culturing centered around the possibility of converting single-celled algae to human, and/or animal, protein supplements until W. J. Oswald and co-workers at the University of California Berkley expanded on this theme by demonstrating that algal systems could be used

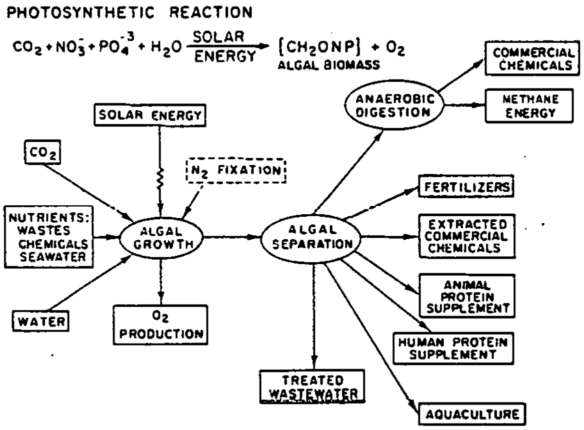

Fig. 1. Process Flow Diagram of Algal Mass Cultures Indicating Materials Required and Potential Applications (from [6]).

for treating wastewater [18] and producing methane via solar energy conversion and anaerobic digestion, as seen in Fig. 1 [17].

The latter concept has been suggested as a dual-functioning process to simultaneously treat wastes and produce energy [2]. However, as shown by Goldman and Ryther [13], on a national level the nutrients available from total U.S. domestic wastewater discharges, even when completely converted to algal biomass, would represent just a small fraction of the total nutrients required for bioconversion. For example, to produce 1% of the U.S. energy demand projected in 1990 would take a population equivalent of 1 billion to meet the nitrogen requirements in the bioconversion scheme depicted in Fig. 1 [13].

As suggested by Goldman and Ryther [13], any large-scale algal bioconversion process will have to include some form of nutrient recycle to be energy and economically efficient. The open type bioconversion scheme in which nutrients and water are continuously supplied from an external source and eventually wasted is uneconomical and may not be possible due to a lack of available nutrients. Thus, either the semi-closed scheme in which nutrients, and possibly water, are recycled after the methane conversion step, or the completely closed scheme in which both nutrients and water are recycled, will have to be employed if bioconversion is to make a viable contribution to U.S. energy supplies.

The three primary nutrients required for algal growth are the inorganic forms of carbon, nitrogen, and phosphorus. The requirements for these nutrients by algae are reasonably known and the stoichiometric relationship between them is often reported as $C_{106}N_{16}P_1$, after the work of Redfield [20], who showed that the chemical composition of marine phytoplankton was typically in these proportions. However, it is now well established that this chemical ratio is achieved only under certain environmental conditions and that the proportions of phosphorus and nitrogen in an algal cell can deviate widely from the above stoichiometry when limitation by one or the other nutrients exists [14]. In recent work in the author's laboratory it was demonstrated that under phosphorus-limited growth of the marine chrysophyte Monochrysis lutheri in continuous culture the C:P and N:P ratios (by atoms) varied from over 1000:1 and 100:1 at low growth rates (the region of severe phosphorus limitation) to ca. 100:1 and 10:1 at 95% of the maximum growth rate (the region of non-nutrient limitation) (Fig. 2). Hence, it is clear that under non-nutrient limitation the Redfield equation of $C_{106}N_{16}P_1$ is a good approximation of the chemical composition of algae because the chemical ratios of different algae, both fresh water and marine, appear to vary in the same proportions.

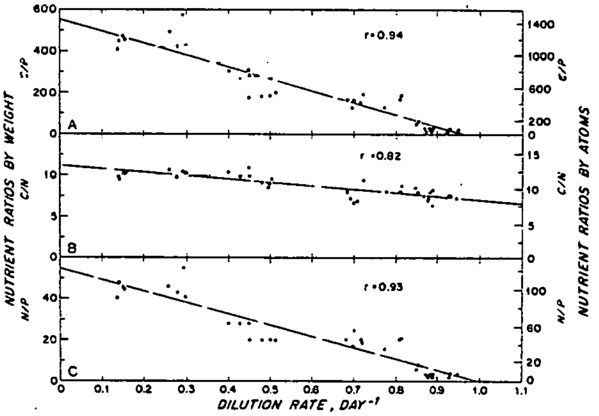

Fig. 2. Effect of Dilution Rate (Growth Rate) on Nutrient Ratios for Monochrysis lutheri Grown in Phosphorus-limited Continuous Cultures (Modified from [14]).

Fortuitously, these are precisely the conditions that are achieved in algal mass cultures when light is made the limiting growth factor. The requirements for carbon, nitrogen and phosphorus in algal mass cultures can then be estimated fairly simply. Assuming that carbon represents about 50% of the organic matter in algae (with a heat of combustion of $\sim$5.5 cal$\cdot$gr^{-1} of ash-free dry weight) [8], the nitrogen and phosphorus requirements in one gram of ash-free algae would be 0.5 x (16/106) = 0.07 gr for nitrogen and 0.5 x (1/106) = 0.005 gr for phosphorus. The total requirements for these two nutrients could then be calculated by multiplying the above unit values by the total yield of algae anticipated, providing a 5-10% excess to ensure that neither nutrient ever becomes limiting. Nitrogen would be supplied as NH_4^+ or NO_3^- and phosphorus as PO_4^{-3}. If these nutrients were to be recycled from the anaerobic digestion portion of a bioconversion process, then it would be necessary to ensure that substantial oxidation of the digested residue containing the recycled nitrogen and phosphorus occurred before the nutrients were added back into the algal growth system.

requirement for inorganic carbon is a far e complex problem, however. Even though the ual quantities of organic carbon produced via photosynthesis can be calculated in the same manner as above, the total amount of inorganic carbon required is much more difficult to calculate. This is because inorganic carbon is distributed among the chemical species CO_2 (aqueous), H_2CO_3, HCO_3^-, and CO_3^{-2} in an exceedingly complex chemical equilibrium system which is controlled by two parameters, alkalinity and pH. In natural fresh and marine waters this chemical system constitutes the main buffering system; losses of inorganic carbon through photosynthesis result in the destruction of buffering capacity, leading to a rise in pH, which can adversely affect algal growth in a number of ways [5].

Normally, the transport of CO_2 from the atmosphere cannot keep pace with algal assimilation of CO_2 during intense algal growth and a rise in pH to over 10 is not uncommon in eutrophic natural waters and mass culture systems [8]. Thus, in algal mass cultures, to avoid the combined problems of inorganic carbon limitation and pH rise, inorganic carbon as gaseous carbon dioxide is usually supplied via some aeration scheme or by creating sufficient turbulence so that sufficient CO_2 can be transferred from the atmosphere. Mixing, to some degree, can enhance CO_2 transport from the atmosphere; but, because of the very low concentration of CO_2 in the atmosphere (0.03%), the transport gradient is always small and CO_2 mass transfer has, in the past, been considered to be ineffective unless very turbulent mixing was employed. However, mixing is required in algal mass cultures for several other reasons: to prevent settling and subsequent decay of organic matter, to prevent thermal stratification, to break down diffusion gradients of essential nutrients which could develop at the cell surface in very intense mass cultures (this is particularly true of seaweeds which are large cells and have long diffusion paths), to prevent epiphyte buildup on the surface of seaweeds, and most important, to provide uniform cell exposure to light because self-shading of cells exists in thick cultures.

The technology and resulting economics of providing adequate CO_2 and mixing in algal mass cultures is amazingly undeveloped considering the substantial research effort now underway to mass culture various freshwater and marine algae for bioconversion applications [6]. The problem of delineating the requirements for carbon dioxide is exceedingly difficult and the major questions still to be addressed are the relative importance of mixing (for solving non-carbon related problems), pH control, and quantity and source of inorganic carbon necessary for maximizing algal yields. Therefore, the major objective of my ongoing research is to address the question of inorganic carbon supply to algal mass cultures primarily from a scientific basis. With a firm understanding of the chemical-biological interactions involved in carbon availability, rational decisions can be made regarding the engineering design of inorganic carbon supply systems for large-scale cultivation systems.

Coupled with the problem of inorganic carbon supply is the question of maximizing algal yields from the standpoint of sunlight availability. As I demonstrated earlier [7], for a given level of available sunlight there is a unique combination of algal growth rate (controlled by the medium flow rate and culture dimensions) and culture depth that will give the maximum yield. Hence, if inorganic carbon is supplied either completely as part of the aqueous medium, or supplemented via external gaseous CO_2 systems, compatibility must be established between the conditions necessary to achieve maximum efficiency of inorganic carbon utilization and yield optimization. Thus, an additional research objective is to establish the conditions of culture operation that will allow for optimum yields by forcing light to be the sole limiting growth factor while at the same time providing for the most economic and efficient supply of inorganic carbon.

RESEARCH IN PROGRESS

The research involves a thorough investigation of the inorganic carbon requirements of algae (both micro and macro species) under mass culture conditions. The major research topics considered included: 1) the growth kinetics of various freshwater and marine algae under inorganic limitation at different pH levels; 2) comparisons of growth with bubbled CO_2 and HCO_3^- alkalinity as inorganic carbon sources at varying pH levels; 3) effects of pH on algal growth under non-carbon limiting conditions; 4) effects of mixing in inorganic carbon limiting and non-limiting situations on algal growth.

The studies are currently being carried out under laboratory conditions with continuous algal cultures and controlled temperature and light intensity (see [3]). Algae are grown in defined medium with pH control using non-carbon buffers. A variety of algal species that typically have been identified as being dominant in outdoor mass cultures are being examined under the above conditions.

Inorganic carbon measurements: Precise measurements of the various forms of carbon are critical to the ongoing study. One of the important considerations is that a good mass balance for carbon be achieved. Thus, if the carbon in the influent medium to the cultures is comprised essentially of inorganic carbon, then that concentration of carbon should equal the sum of particulate organic carbon (algal biomass), dissolved organic carbon (excreted by the algae), and unused inorganic carbon remaining in the culture.

Particulate carbon analysis on filtered samples is carried out routinely here at the Woods Hole Oceanographic Institution with a Perkin-Elmer 240 elemental analyzer. Protocols for using this instrument for particulate carbon analysis (and simultaneous particulate nitrogen measurements) have been established over many years of operation, and the technique is highly reliable. The technique involves combustion of particulate carbon to CO_2 in the presence of oxygen and other catalysts and subsequent measurement of this CO_2 by thermal conductivity.

Dissolved inorganic carbon has traditionally been measured by infrared spectroscopy and a number of commercial instruments are available for this purpose. Dissolved organic carbon analysis usually first involves oxidation of the organic carbon to CO_2 and subsequent measurement of the generated CO_2. For freshwater several oxidation techniques such as high-temperature combustion, UV-irradiaton and persulfate oxidation have been used successfully. In contrast, analysis of dissolved organic carbon in seawater has proved to be considerably more difficult because of the refractory nature of many seawater organic components. Thus, no completely satisfactory technique exists at present.

Recently I obtained a Dohrman DC-54 Ultra-low Level Total Organic Carbon Analyzer with funds obtained through a grant from the Ocean Industries Program (OIP) of the Woods Hole Oceanographic Institution. This instrument was originally obtained for measuring dissolved organic carbon in sea water. In principle, a sample is first acidified to pH = 2, followed by sodium persulfate addition. The sample is then introduced into a chamber where it is initially sparged with pure helium in a closed system. All inorganic carbon present is converted to CO_2 and evolved from the liquid phase and trapped in a lithium hydroxide-filled tube. The sample is then passed through a quartz coil wrapped around a UV lamp. The combination of UV exposure and persulfate oxidation, in principle, leads to oxidation of all organic carbon to CO_2. The sample is then passed to a second chamber, sparged again with helium, but now the CO_2 generated is first converted to methane over a nickel catalyst and the resulting methane measured precisely in a flame ionization detector. The instrument has a precision of $\pm$ 10 ppm (or $\pm$ 2%) and an accuracy down to $\sim$50 $\mu g \cdot \ell^{-1}$.

In our current studies we have found the instrument to give rather poor recovery of dissolved organic carbon in seawater ($\sim$ 60-80%). This has been attributed to insufficient UV lamp wattage and sample exposure time according to the studies of Collins and Williams [4]. The main problem seems to be incomplete oxidation of refractory compounds in natural seawater. However, it is hoped that the instrument will be useful in quantifying excreted organics of algae which probably are not particularly refractory. Currently I am engaged in an intercalibration of the instrument against other techniques sponsored by the Canadian Research Council at Dalhousie University, Halifax, Nova Scotia. The results of this survey will be available in the Spring, 1979 and will provide an accurate measure of the instrument's efficiency in recovering dissolved organic carbon as CO_2.

The same instrument, however, has been extremely useful for measuring dissolved inorganic carbon (DIC). I have modified the instrument for DIC measurements by eliminating the lithium hydroide trap and persulfate addition, and by turning off the UV lamp. Thus all DIC sparged out upon acidification is reduced to methane and measured by the flame ionization detector. As seen in Fig. 3, the instrument displays a linear calibration between 0-20 $mg \cdot \ell^{-1}$ DIC and is extremely accurate even in the $\mu g \cdot \ell^{-1}$ range.

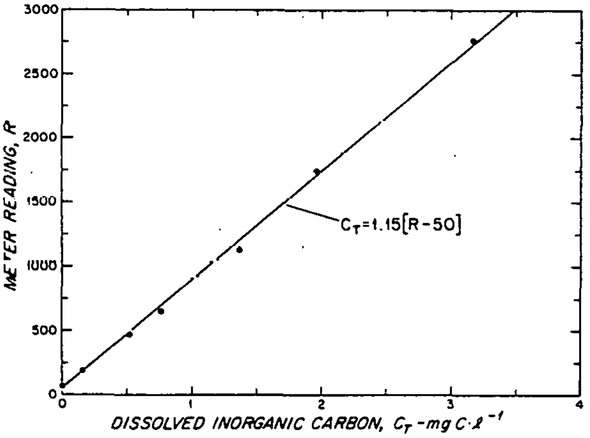

Fig. 3. Calibration Curve for Dohrmann DC-54 Ultra-low Total Carbon Analyzer. Instrument Span Setting = 367. Although Not Shown Here, Calibration is Linear to 20 $mg \cdot \ell^{-1}$ Dissolved Inorganic Carbon.

Effect of pH on marine algal growth: Five marine algal species listed in Table 1 were grown in inorganic carbon limited continuous cultures maintained at a dilution rate of 0.5 day^{-1} and an influent total inorganic carbon concentration of 6.3 $mg \cdot \ell^{-1}$ (supplied solely as $NaHCO_3$). There was no gas phase as the cultures were sealed off from the ambient environment. The cultures were maintained at 20°C with continuous visible light of 0.06 $ly \cdot min^{-1}$. One set of cultures

Table 1. EFFECT OF pH ON STEADY STATE BIOMASS FOR VARIOUS MARINE MICROALGAE GROWN IN INORGANIC CARBON LIMITED-CONTINUOUS CULTURES (6.3 $mg \cdot \ell^{-1}$ DIC) AND AT A DILUTION RATE OF 0.5 DAY^{-1}.

SPECIES	UNBUFFERED pH	CELL COUNT (10^5/ml)	BUFFERED pH	CELL COUNT (10^5/ml)
CULTURE MEDIUM	8.60		7.75	
P. TRICORNUTUM	9.39	6.5	7.88	9.3
M. LUTHERI	8.65	W.O.[a]	7.89	7.3
D. TERTIOLECTA	8.99	0.7	7.88	3.5
S. COSTATUM	8.60	W.O.	7.88	2.8
T. PSEUDONANA (13-1)	8.51	W.O.	7.87	1.7

a) Cells washed out of culture.

ras buffered with 20 mM of a "Good" organic er (HEPPS-pk = 8.0) along with the HCO_3^- buf- while the other set was left buffered only the HCO_3^-. Hence, in the set without the rganic buffer uptake of inorganic carbon led to destruction of the buffer system via the reac- ions:

$$H^+ + HCO_3^- \rightarrow CO_2 + H_2O \quad (1)$$

$$HCO_3^- \rightarrow CO_2 + OH^- \quad (2)$$

nd a corresponding rise in pH. Depending on how ell the particular species grew there would be a orresponding rise in pH above the medium pH alues. As seen in Table 1, only two species, haeodactylum tricornutum (a diatom) and Dunali- lla tertiolecta (a green alga), grew in the edium lacking the organic buffer, raising the pH espectively to 9.4 and 9.0. The other species ailed to maintain a steady state and washed out f the culture units. In contrast, when the cul- ure pH values were maintained at 7.9 all five pecies grew well, and both P. tricornutum and D. ertiolecta grew better than at the higher pH alues.

ecause of the low biomass present, it is highly nlikely that there was a rate bottleneck in quations 1 and 2. Thus, even if any of the spe- ies were obligate CO_2 users, this should not ave influenced the results as HCO_3^- simply could ave acted as a carbon reservoir for the supply f CO_2 to the algae. These results, hence, rovide some good circumstantial evidence that ertain marine algae are very sensitive to pH hanges. P. tricornutum typically has been the ominant species in large-scale outdoor cultures n which pH was not controlled, reaching values s high as 10.3 during mid-day [10, 12]. Hence, t may well be that the ability of P. tricornutum o tolerate very high pH values gives it a compe- itive edge over other species, thus allowing it o win out in competition as the pH rises in cul- ures buffered primarily by HCO_3^-.

CO_3^- as a carbon source: Because P. tricornutum as excellent growth characteristics over a wide H range, further experiments with this species nvolving HCO_3^- as the sole carbon source were erformed. Employing the same environmental con- itions and eliminating the gas phase as in the revious experiments, P. tricornutum was grown in ontinuous culture with varying levels of HCO_3^- s the sole limiting nutrient. Culture pH values ere controlled with up to 40 mM of the HEPPS uffer. All other nutrients were added in excess e.g., NH_4^+, PO_4^{-3}, trace metals) and HCO_3^- as added in the range 6-108 mg·ℓ^{-1} total inor- anic carbon. At the higher HCO_3^- additions (and orresponding higher biomass levels) OH^- pro- uction via reaction 2 was so great that culture H values rose to as high as 9.0 even with 40 mM f the HEPPS buffer present. Even with this arying pH, there was a linear relationship be- ween steady state biomass level and HCO_3^- addi- ion at a fixed dilution rate of 0.5 day^{-1} (Fig.). Moreover, the efficiency of dissolved inor- c carbon conversion to algal organic carbon - via photosynthesis - varied from virtually 100% at low HCO_3^- additions to over 90% at the highest levels.

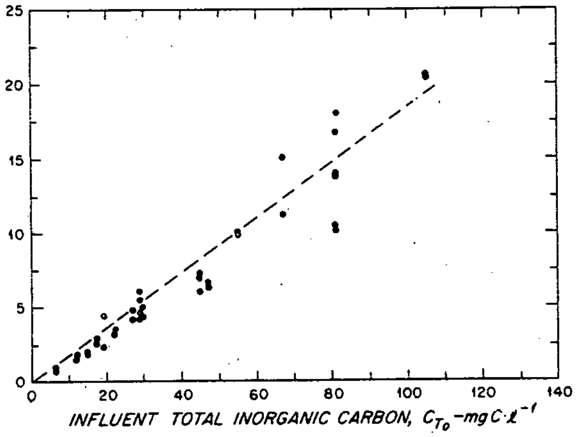

Fig. 4. Steady State Biomass of P. tricornutum as a Function of Influent Inorganic Carbon Concentration as HCO_3^- Added to Medium. Culture Buffered with 40 mM HEPPS and Dilution Rate Fixed at 0.5 day^{-1}.

Above 80 mg·ℓ^{-1} added inorganic carbon some chemical precipitation was observed, and it suspected that the solubility product of $CaCO_3$ and/or other compounds was exceeded. In some experiments at these high HCO_3^- additions the cultures would deteriorate over several days, probably due to the relative slow formation of the inorganic precipitates.

The results of this experiment conclusively show that marine species such as P. tricornutum can utilize inorganic carbon from HCO_3^- at concentrations over four times that present in natural seawater and at very high efficiencies when the pH is controlled below 9. Similar experiments without the organic buffer will be required to determine the effects of high pH on the relation ship between HCO_3^- addition and biomass production.

It is interesting to compare these results with those from previous outdoor mass culture experiments with marine microalgae [10, 11]. In those experiments P. tricornutum was one of several diatoms that dominated at times through natural selection. The bell-shaped curves of yield (gr dry wt·m^{-2}·day^{-1}) vs growth rate (= dilution rate) are summarized in Figs. 13 and 14 of Goldman [7]. Peak yields up to 25 gr dry wt·m^{-2}·day^{-1} occurred at growth rates of 0.75-1.00 day^{-1}. Considering that the depth of the culture units was 50 cm and that the cultured algae contained 45% organic carbon of the ash-free dry weight, then, according to the simple formula:

$$P = \mu d X \quad (3)$$

in which P is the yield per unit area, the growth rate, d the depth, and X the standing crop, for the above conditions a standing crop of 20-30 $mg \cdot \ell^{-1}$ organic carbon was required for maximizing yields. As seen from the results in Fig. 4, marine algae can easily convert that much HCO_3^- to organic matter. Thus the key to optimization of inorganic carbon supply may be in pH control to allow full utilization of available HCO_3^- which in seawater and many freshwaters is 1-2 mM (12-24 $mg \cdot \ell^{-1} C$).

Kinetics of inorganic carbon-limited algal growth: The kinetics of inorganic carbon limited algal growth of two freshwater green algal Scenedesmus obliguus and Chlorella vulgaris, species commonly found in intensive outdoor cultures, have been investigated. Both species were cultured in an artificial freshwater medium with inorganic carbon supplied to the continuous cultures at a limiting concentration of 10 $mg \cdot \ell^{-1} C$ as HCO_3^- [9]. Once again, no gas phase was present. Mixing was accomplished by magnetic bar stirring. Other nutrients were added in excess and the pH was maintained at 7.1-7.2 with a 10 mM phosphate buffer (equi- molar di- and monophosphate).

Steady state measurements of biomass and residual inorganic carbon were made at various growth rates spanning a range from less than 0.2 day^{-1} to cell washout. As seen in Fig. 5a, at all growth rates up to the maximum rate, there was virtually 100% utilization of available inorganic carbon for S. obliguus. As the maximum growth rate was approached there was a rapid drop-off in steady state biomass. Translated to production of organic carbon (Fig. 5b), there was a linear increase in yield with growth rate up to peak yield just before cell washout at ~1.65 day^{-1}. The results for C. vulgaris were identical except that the maximum growth rate of this species was somewhat higher (~ 2.20 day^{-1}).

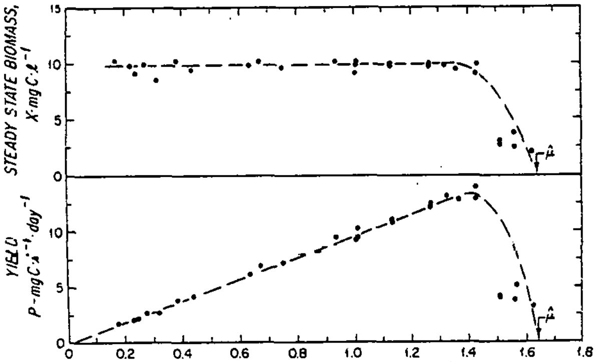

Fig. 5. Growth Kinetics of Scenedesmus obliguus in Inorganic Carbon-limited Continuous Cultures. pH Kept at 7.1-7.2. 3a - Steady State Biomass as Particulate Carbon Concentration as a Function of Growth Rate; 3b - Yield of Particulate Carbon as a Function of Growth Rate.

The carbon cell quota (carbon content per cell) was measured as a function of growth rate (Fig. 6). There was over a three-fold increase (7.5-25 pg $C \cdot cell^{-1}$) in this parameter with increasing growth rate for S. obliguus. For C. vulgaris, the trend was similar, but the increase (2.9-10.4 pg C $cell^{-1}$), although in approximately the same proportion as with S. obliguus, reflects the considerably smaller size of this species.

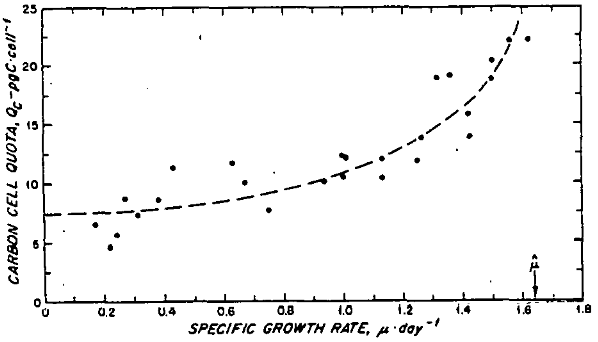

Fig. 6. Carbon Cell Quota of Scenedesmus obliguus as a Function of Growth Rate in Inorganic Carbon-limited Continuous Culture. pH Kept at 7.1-7.2.

The overall results to date suggest that for both freshwater and marine microalgae HCO_3^- is a suitable carbon source provided that the pH is controlled. The efficiency of inorganic carbon utilization under these conditions is virtually 100%.

Effect of carbon source on maximum growth rates: To establish the relative response of C. vulgaris and S. obliguus to different inorganic carbon sources, both species were grown in batch culture on enriched medium. Inorganic carbon was supplied either as HCO_3^- added directly to the medium (6 or 10 mM), or as a gas mixture of CO_2 and air bubbled into the cultures at a rate of 0.03 $m^3 \cdot hr^{-1}$ (std). The mixtures included 100% air (0.03% CO_2), 1, 5 and 100% CO_2. A phosphate buffer (25 mM) was used to maintain the pH between 6 (100% CO_2) and 7.7 (10 mM HCO_3^-). Continuous light (~0.06 $ly \cdot min^{-1}$) and 20°C temperature were established. Each culture was inoculated with the respective species obtained from an inorganic carbon-limited continuous culture similar to those described in the previous experiment.

As seen in Table 2, under batch culture conditions C. vulgaris grew at a maximum rate ($\hat{\mu}$ = 2.1-2.2 day^{-1}) regardless of whether the inorganic carbon source was HCO_3^- (6-10 mM) or gaseous CO_2 (0.03-1% CO_2). At 5% CO_2 $\hat{\mu}$ was reduced one-half and with 100% CO_2 complete inhibition occurred. S. obliguus, in contrast, was maintained at $\hat{\mu}$ (1.6-1.7 day^{-1}) when gaseous CO_2 in the range 0.03-5% was added; but with HCO_3^- a the carbon source there was increasing growth

Table 2. EFFECT OF DIFFERENT INORGANIC CARBON SOURCES ON THE MAXIMUM GROWTH RATES OF TWO FRESHWATER GREEN ALGAE GROWN IN LABORATORY CULTURES.

INORGANIC CARBON SOURCE	MAXIMUM GROWTH RATE (DAY^{-1}) CHLORELLA	SCENEDESMUS
BATCH GROWTH		
HCO_3^- (mM)		
6	2.07	1.32
10	2.07	1.16
CO_2^- AIR MIXTURE (% CO_2)		
0.03	2.22	1.54
1	2.14	1.71
5	0.97	1.57
100	0	0
CONTINUOUS CULTURE (WASHOUT)		
HCO_3^- (mM)-1	ND*	1.65

*ND - NOT DETERMINED AS OF YET

inhibition with increasing HCO_3^- in the range 5-10 mM. As with C. vulgaris, no growth occurred with 100% CO_2.

The major conclusion from these data, is that the rate of supply of inorganic carbon as gaseous CO_2, rather than the concentration of CO_2 in the gas mixture, is the critical factor controlling the availability of inorganic carbon for algal growth. With proper pH control, air alone is an adequate source of inorganic carbon as long as the supply rate, in terms of a mass transfer rate, is maintained at an optimum level. In addition, at least for some species like C. vulgaris, HCO_3^- is as good a carbon source as gaseous CO_2.

CONCLUSIONS

In their recent cost analysis of large-scale biomass systems for bioconversion of energy Ashare et al. [1] investigated the requirements for CO_2 supply to a 100 square mile freshwater algal system. They considered both atmospheric air supply (0.03% CO_2) and power plant stack gas with a CO_2 content of ∿ 10% as potential sources of inorganic carbon. They arrived at the startling conclusion that inorganic carbon supply via a network of pipes, headers, and sparging units placed intermittently in the large-scale culture units would be prohibitively expensive even with enriched power plant exhaust gases. The analysis was based on a number of assumptions, many of which were criticized by Oswald and Benemann [16].

[illegible] main point of the analyses (and the subsequent discussion of this analysis) is that very little information is available upon which to estimate rationally the carbon requirements of large-scale algal cultures and to design supply systems accordingly. For example, the analysis of Ashare et al. [1] did not include consideration of the aqueous chemistry of inorganic carbon (i.e., pH, alkalinity, and HCO_3^- concentration). Carbon requirements were treated as purely a gas transfer problem with no consideration given to the chemical reactivity of CO_2 with HCO_3^- and CO_3^{-2}, the fact that pH plays a crucial role in controlling the CO_2 gradient between the gas-liquid phase, and that algal species differ considerably in their ability to tolerate pH variations and to utilize different inorganic carbon sources. Moreover, if CO_2 is supplied via gas transfer systems then the conditions for maximizing the efficiency of CO_2 transfer and assimilation by algae are not necessarily the same as those required to maximize yields from the standpoint of forcing available sunlight to be the sole limiting growth factor.

Based on our results to date, it appears that with proper pH control, HCO_3^- is an adequate source of inorganic carbon for maximizing algal yields. The economics of supplying HCO_3^- as a salt versus a distribution system for gaseous CO_2 supply have not been investigated. A major factor overriding such an analysis is that the critical component is pH control. As seen from equations 1 and 2, without a non-carbon buffer, pH in an algal culture is controlled by the bicarbonate alkalinity, the intensity of algal growth, and the rate of gaseous CO_2 supply. Whether HCO_3^- additions alone can be used simultaneously for controlling pH and providing adequate inorganic carbon in such a system have not been determined as yet, but are included in my future research plans.

FUTURE RESEARCH

Research in the four areas described earlier is in various stages of progress. I originally hoped to complete by June, 1979 experiments in all four areas with a variety of algae (marine micro and macro and freshwater micro species). This projection now appears to be overly optimistic.

As seen by the results to date, work is progressing well in the first two areas (growth kinetics and comparisons of different carbon sources). Research in these areas will be continued during the remainder of this first year (through May, 1979) with the two freshwater and one marine species. Experiments will involve comparisons at varying pH of bubbled CO_2 in varying mixtures with air and helium (up to 100% CO_2) with HCO_3^-. In addition, it is hoped to repeat these experiments with other species. To be included are additional freshwater algae (the blue-green species Microcystis and Spirulina), and several marine algae (Dunaliella tertiolecta and Thalassiosira pseudonana - 3H). All of these species appear to tolerate a reasonably wide pH range.

Thus, a catalog of responses for representative and typically dominant microalgae from both freshwater and marine environments will be obtained.

In addition, work on the second two research areas, pH effects under non-carbon limiting conditions and mixing effects, will be initiated on both the freshwater and marine microalgae. Experimental work on macroalgae, similar to that with the microalgae, will be postponed until later in the research program. All of the laboratory research has been and will be carried out under rigid laboratory conditions in continuous cultures.

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BIOLOGICAL INVESTIGATIONS OF MARINE FARMS
(FOR BIOMASS PRODUCTION)
CONTRACT NO. DE-AC03-76ET20000, JULY 1 1975

M. Barcelona S. Lieberman
V. Gerard S. Manley
J. Kuwabara W. North
California Institute of Technology, Pasadena CA 91125

3STRACT

ur project seeks to develop a methodology for roducing plant biomass in surface oceanic envionments far from shore. Our research thus far has rimarily utilized giant kelp, *Macrocystis*. *acrocystis* grows naturally near the outer edge of he littoral zone, it can be coppiced, and has been arvested mechanically for decades in California. ur earliest field studies demonstrated that *acrocystis* grew poorly in offshore surface water ue to scarcity of nutrients such as phosphorus and itrogen. A proper balance of trace metals is also mportant. Seawater from 1000 or more feet deep is ich in nutrients and proved superior to enriched urface water in supporting kelp growth. We also nvestigated elementary requirements of *Macrocystis* nd amounts needed of the most critical microutrients. We are presently extending the scope of ur research to include assessments of biomass ields obtainable from adult plants continuously ertilized.

ESCRIPTION OF TASK

he general objective of our project is to develop methodology for producing plant biomass in the ast stretches of the world ocean, comprising about) percent of the planetary surface. We have hosen giant kelp, *Macrocystis*, as the test organsm for our basic studies. *Macrocystis* is one of he few seaweeds that can be coppiced, and it can e harvested mechanically, as shown by commercial entures in California that have successfully perated for almost 50 years.

ur first field studies examined adult *Macrocystis* ransplanted to artificial structures anchored in ffshore waters. These plants grew poorly or not t all, primarily because of scarcity of mineral utrients [1]. Fertilizing operations would learly be needed on marine farms. We concluded hat a comprehensive understanding of *Macrocystis* utrition was a priority field of research. We ave consequently devoted considerable effort owards determining the kinds and amounts of critial elements required by giant kelp. From this nformation we hope to devise an optimal fertilzing strategy. Sufficient information has now een accumulated to permit preliminary fertilizing xperimentation in the field and in large-scale [illegible]oratory studies. We are presently expanding [illegible] scope of our program to include such fertilzing investigations with a view toward gathering yield data from plants that are adequately nourished on a continual basis.

OBJECTIVES

Our current objectives are threefold:

1. Determine kinds and amounts of elements critical for nutrition of *Macrocystis* under oceanic conditions.
2. Devise an optimal fertilizing strategy for use on marine kelp farms.
3. Assess harvesting yields from adult *Macrocystis* plants that are continuously fertilized.

APPROACH

Field Studies

Most of our field work has been involved with adult *Macrocystis* transplanted to structures moored well away from the coast to provide exposure to waters characteristic of oceanic environments [1]. The earlier structures typically consisted of networks or grids of rope positioned about 40 to 60 ft beneath the sea surface by appropriate use of buoys and cables. Studies at any one time utilized about 10 to 30 transplants. Most of the early field work was conducted collaboratively with the Naval Ocean Systems Center of San Diego. During the past two years we have become associated as participants in General Electric Co.'s Marine Biomass Program, while remaining under separate DOE funding. Global Marine Development, Inc. (GMDI), a subcontractor to G.E., designed and supervised construction of a novel modular substrate for an experimental oceanic kelp farm. The novel structure consists of a central spar buoy that provides flotation for the system and houses machinery and instrumentation (Fig. 1). The spar buoy is held in a three-point mooring that attaches to a swivel near the base. Just above the swivel, six poles project outward in a symmetrical pattern. Circumferential cables join the outer sections of the poles and serve as substrate for transplanted kelp. A PVC pipe, 25" I.D. and 1500 ft long, is suspended from the base of the spar buoy, below the swivel. Three diesel-powered pumps in the spar buoy draw up about 10,000 GPM of nutrient-rich water from 1500 ft depths and disperse it near the surface to fertilize the kelp canopies. The system also requires deployment of a peripheral curtain just beyond the kelp canopies

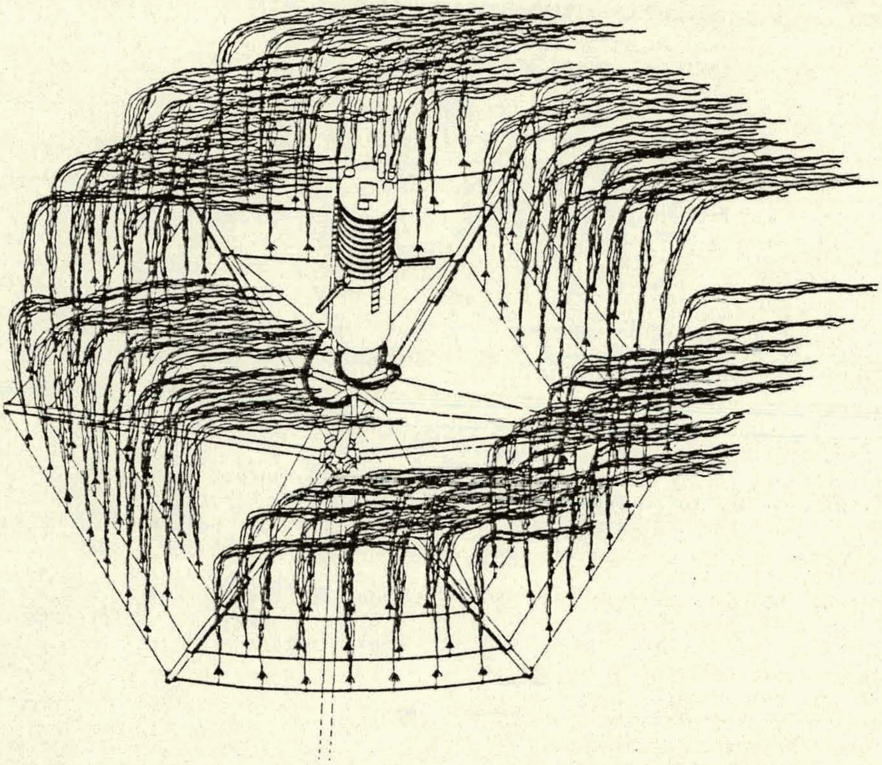

Fig. 1 Diagram of Test Farm with transplanted kelp in place. Striped cylinder in center is machinery buoy that provides flotation.

to prevent all but the strongest currents from removing the deep water before the kelp has had an opportunity to accumulate the dissolved nutrients. GMDI installed the buoy and associated components during fall 1978 near our laboratory headquarters in water 1800 ft deep. Our research group was responsible for transplanting Macrocystis to this Test Farm, monitoring its condition, and assessing biomass yields by periodically harvesting the canopy. We are also studying oceanography of the deep water in the region of the Test Farm, to provide background information relating to character and composition of this critical component of the experiment.

Laboratory Studies

Adult Macrocystis plants may be 100 or more feet long and weigh several hundred pounds. We could not accommodate such giants in the laboratory facilities available to us, so we have been constrained to work with juvenile plants that are usually one to twelve inches long. Most of our nutritional studies involve holding such small plants in about ten gallons of seawater medium, in specialized aquaria [2]. Typical experiments involve periodic determinations of wet weights of the plants as a measure of growth. Usually test plants in seawater supplemented with chemicals under study are compared to controls in unenriched seawater or in seawater enriched only with a minimum of elements (e.g. N and P) at standardized concentrations to ensure that at least some growth occurs. Most of the juvenile plants used in our experiments have come from local natural populations or from material we have raised in culture. Quite recently we have been fortunate in securing genetically uniform specimens raised for us in the laboratory of Dr. Michael Neushul at the University of California, Santa Barbara.

In another phase of our laboratory research, we have been attempting to culture Macrocystis gametophytes and small sporophytes in a completely defined artificial seawater named Aquil by the designers [3]. A major advantage of using a completely defined medium such as Aquil is that the speciation of its chemical components can be calculated using the computer program REDEQL2 [4]. Free ion concentrations of trace metals are of particular interest. We prepare and use Aquil under scrupulously clean chemical conditions to avoid possibilities of contamination.

ACCOMPLISHMENTS

Field Studies

We introduced 103 adult Macrocystis transplants the Test Farm during late November and early December 1978. A number of unanticipated ional difficulties developed with the Test quipment. For example, the protective cur- .. .as lost, so that dwell time of the deep water ound the kelp canopies was inadequate. Nitrogen ntents of canopy fronds were in the range of 2.5 3.0 percent of the dry weight while the curtain s in place (usual values are in the neighborhood 1.0 to 1.5 percent). Consequently, we believe t design of the system is suitable, since the ints were apparently able to accumulate nutrients om the dispersed deep water. Unlike juvenile ants in enriched water in the laboratory, we corded no growth stimulation among kelp fronds on e Test Farm while the system was operating satisctorily. Possibly adult tissues respond fferently to nutrient enhancement compared to veniles. The key to increased yields among adult onds may involve increasing the density of biоss in the water. Presumably higher concentraons of nutrients would support higher biomass nsities. The data, however, are still few and ch more research must be conducted before conusions can be drawn. Without the protective rtain, storms and lack of nutrients led to complete losses of all our transplants. A replacement rtain is currently under design. We anticipate rther experimentation with *Macrocystis* transants after the new curtain is in place.

welling processes occur seasonally in southern lifornia, bringing deep water toward the surface. Strong upwelling occurred locally in spring 1979. Probably this natural phenomenon stimulated microscopic stages of *Macrocystis* that developed on the Test Farm from spores liberated by our transplants several months earlier. We observed massive numbers of juvenile plants on the solid structures of the Test Farm in April and May 1979. We are presently monitoring these individuals to assess their fates. Hopefully, the artificially upwelled deep water will encourage survival and growth of these recruits.

Our oceanographic studies of deep water near the Test Farm have just begun, so that data are still sparse and do not yet describe seasonal patterns. It is clear, however, that conditions are complex, involving several water masses.

Laboratory Studies

Thus far we have screened 21 different substances for potential as stimulators of kelp growth [5]. Very recently, one of us (James Kuwabara) successfully carried cultures of microscopic stages of *Macrocystis* from settled spores completely through the gametophytic phase to embryonic sporophytes in a petri dish culture. Only nine elements were involved among the micronutrients used in formulating the successful medium (Table 1). All of the trace metals except Mn in this Aquil formulation were added at concentrations considerably above

TABLE 1

Nannomoles of nine micronutrients and of EDTA added to the completely defined artificial seawater Aquil to yield a medium that sustained development by *Macrocystis* zoospores in petri dish cultures completely through the gametophyte stage to embryonic sporophytes.

Micronutrient used	nM added	Free ion conc. nM	Major species present and (%)	
Fe^{+3}	400	$7X10^{-11}$	FeEDTA	(100)
Mn^{+2}	10	1	MnEDTA $MnCl^{+}$	(65) (23)
Co^{+2}	40	0.04	CoEDTA	(99)
Cu^{+2}	5	0.00002	CuEDTA	(99)
Zn^{+2}	250	0.1	ZnEDTA	(100)
MoO_4^{-2}	100	100	--	
$EDTA^{-2}$	6,000	0.00007	CaEDTA FeEDTA	(89) (6)
NO_3^{-1}	15,000	15,000	--	
PO_4^{-3}	2,000	0.3	HPO_4^{-2} $MgHPO_4^{-1}$	(51) (47)
I^{-1}	100	100	--	

levels reported for oceanic waters. The computed free ion concentrations, however, were always low in Aquil due to the strong complexing capacity of the chelating agent used (i.e. EDTA). More than a hundred-fold volume increase occurred from growth in some of these plants, so it seems likely that any residual elements carried in as components of the spores would be quite diluted among the sporophytes if an element was not available from the medium. The cultures were run for only 14 days. This span is quite short for completion of the gametophytic phase. It seems unlikely that any critical element was limiting development in these cultures. It appears, rather, that the micronutrient levels essential for gametophytic development were provided. These requirements may be similar for the sporophyte.

Quite interestingly, critical elements such as N, P, and possibly Zn and Cu that may be scarce in offshore surface water seem to be adequately abundant in deep water. Other elements that may occur

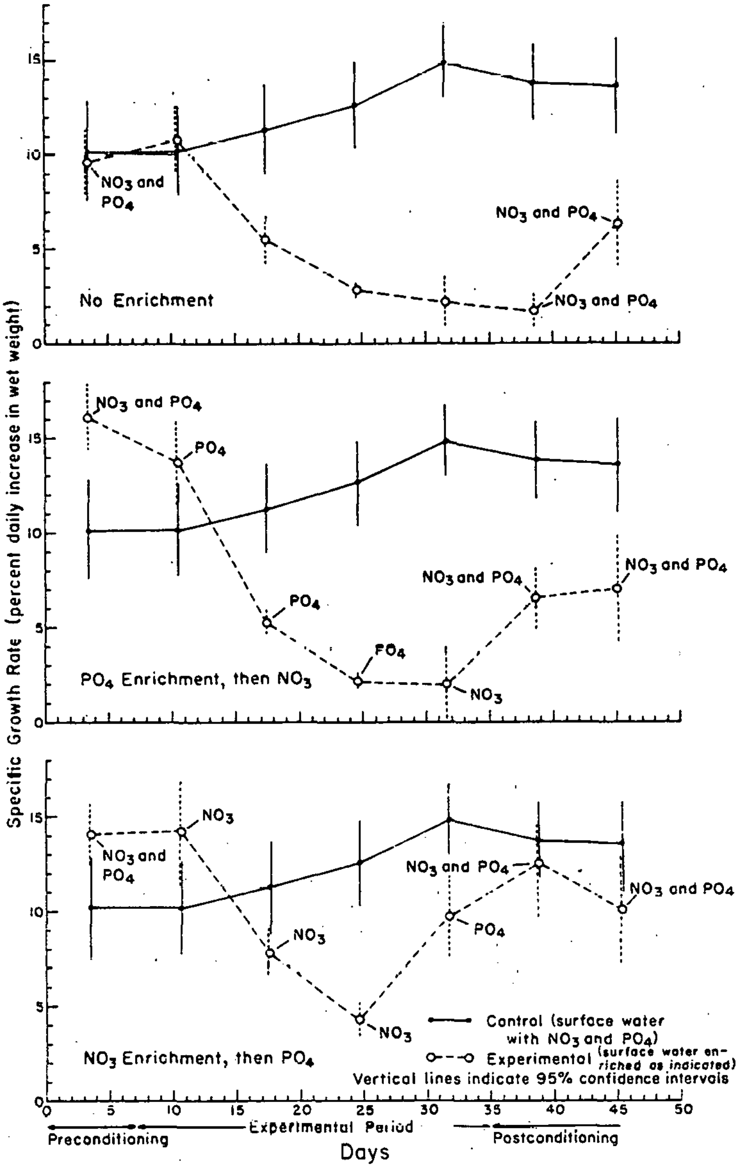

Fig. 2. Growth histories from groups of nine juvenile Macrocystis plants, comparing growths by controls to performances in unenriched, PO_4-supplemented, and NO_3-supplemented surface seawater. Concentrations used were 2 uM for PO_4 30 uM for NO_3.

w concentrations in oceanic deep water . Mn) are probably present in sufficient its in surface water. Consequently, appropriate mixtures of surface and deep water appear to constitute the best medium for growth by Macrocystis. We still need much more information, owever, on trace metal contents of deep water, articularly whether seasonal fluctuations occur nd availabilities of the various chemical species o plants such as Macrocystis. Trace metal content f surface water near our Test Farm appears adequate to support good growth after it is supplemented with N and P (Fig. 2). But the site is only our miles from shore. It seems possible that ater here may be somewhat influenced by the nearby ontinent. A different fertilizing strategy may be eeded at a site a hundred or more miles from the oast where concentrations of elements such as Zn nd Cu appear to be quite low in the surface waters 6,7].

e are presently investigating relations between -uptake, growth, and light intensity. We have ound that photosynthesis among juvenile acrocystis sporophytes from rope cultures grown in ur culture room saturates in the range 12,000 to 5,000 ergs/cm^2/sec (cool white fluorescent lighting). A number of variables affect nitrogen uptake nd metabolism in Macrocystis, and we are hoping to ssemble sufficient data so that a basis for understanding the more important relationships is chieved.

UTURE PLANS

ur plans for future work involve four major areas f research: continuing laboratory studies of utritional requirements of kelp, including experientation with completely defined media; laboratory tudies using intermediate-sized kelp plants in arge tanks to study responses to nutrient-enriched ater; continuing studies of fluctuations by microutrients in deep water and causes thereof; and urther studies utilizing the Test Farm substrate o determine harvest yields from kelp plants coninually exposed to artificially upwelled water.

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NOTES

MEMBRANE PROCESSES OF SEPARATION AND CONCENTRATION IN BIOMASS HARVESTING, PRODUCTION AND REFINING

Columbia University Harry P. Gregor, Professor

BSTRACT

ew developments in the application of embrane technologies to biomass produc- ion, harvesting and refining are summari- ed, including the use of ultrafiltration o harvest microscopic algae and single ell protein, to concentrate suspended and issolved solids in corn refining, and to emove and concentrate the constituents of eer and stillage resulting from the fer- entation of corn and molasses to ethanol. esearch activities involving the use of ew membranes for other biomass-related pplications are described and include: he use of alcohol-selective membrane per- eation systems for the removal of ethanol rom a recirculating beer; the use of com- arable water-selective membranes to en- ich distillates from 80% to 99.5% ethanol; he use of "tight" anion-selective electro- ialysis membranes to remove concentrated Cl from acid-sugar mixtures to allow for he low-cost reuse and recycle of that acid n the conversion of cellulose to sugars. ecause membrane systems consume little nergy and are highly efficient in effec- ing separation and concentration proces- es important in many phases of biomass onversion systems, it is anticipated that heir applicability will prove significant n the overall effort.

NTRODUCTION

his communication summarizes the work of ur laboratories relevant to the various issions of the Biomass Energy Systems rogram, Department of Energy. Our labora- ories have for several years emphasized esearch and development on the prepara- ion of new membranes, studies of their undamental properties and an examination f the feasibility of membrane processes or a number of different industrial appli- ations. A general review on membranes membrane technologies summarizes the ate of the art [1].

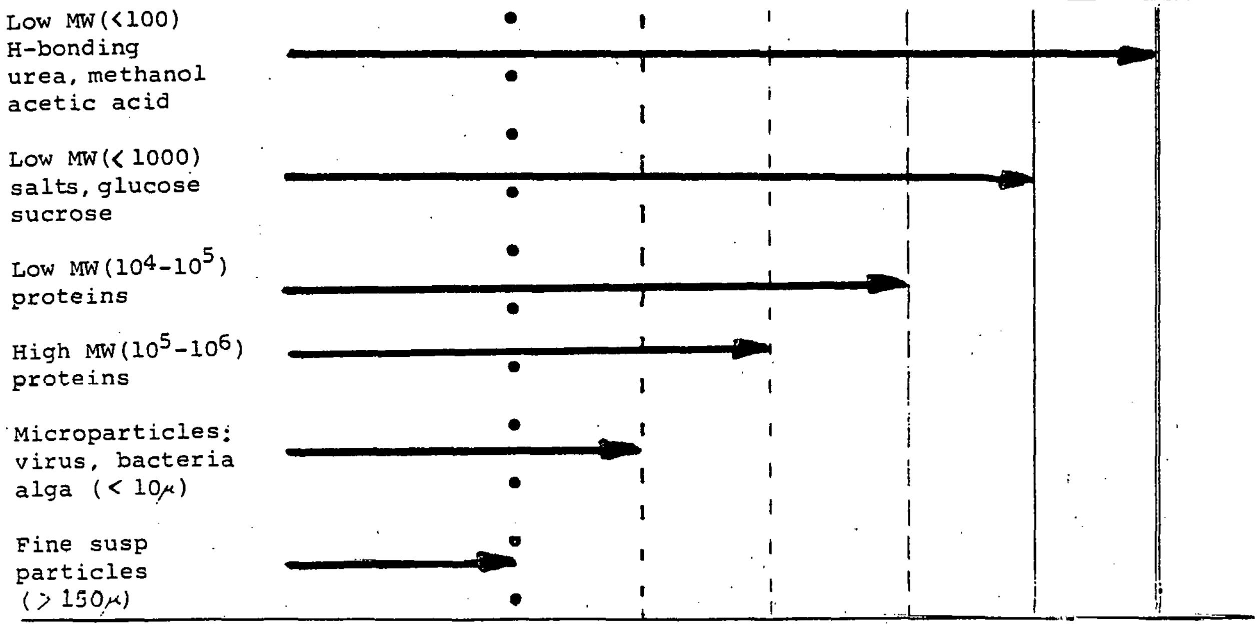
Membrane technologies have a unique place in the field of biomass production and refining because they are capable of carrying out sophisticated processes of concentration, separation and purification at minimal energy requirements. This is because membrane devices can operate at practical rates while consuming only about 3-4 times the thermodynamic, minimal energy requirement. Conventional processes of distillation, as an example, require several times that energy requirement. For example, the conventional quadruple effect evaporator (which delivers about 2.6-2.7 plates), when used to dewater 1000 gallons (KG) of fermentation stillage consumes about $12 of fuel at today's prices. The additional costs for the device itself, labor, etc., is of the order of about $0.50/ KG. In contrast, we have shown [2] that membrane processes can effect the same separation at energy costs of about $0.50/ KG, with an additional cost for the devices and their operation of about $2.00/ KG. Because of high energy costs, new stills are coming into use, ones more efficient in their use of energy but more complex and expensive. The membrane systems still appear to have an overall cost advantage over distillation processes of about 2-3 fold at the present time, and this advantage should increase as fuel costs increase.

Our programs have explored the application of membrane technologies to a wide range of food-related processes, including those of corn wet milling, sugar refining and soy processing, the operation of feedlot and related facilities, and the fermentation industries (acetic acid, ethanol), among others. Many of these applications are relevant to energy supply and production, and all of these help conserve significant amounts of process energy. For the past 15 years we have concentrated on the development of non-fouling membranes, as up to now the principal deterrent to the widespread, low cost utilization of membranes has been fouling. We have developed a new kind of polymer formulation which can be applied as a coating or cast upon a porous support as a membrane; this formulation leads to a non-fouling material, one which has shown itself to remain unaffected by process streams of natural origin. We have prepared membranes and surfaces in the following range of pore diameters: screens having openings of about 100 mesh or 0.15 mm; microporous filters having pore diameters of about 0.1-0.3μ; loose ultrafiltration (UF) membranes which separate high ($>10^6$ MW) from low ($\sim 3\times10^4$ MW) proteins; tight UF membranes which pass small molecules of MW < 1000.

Each of these systems has an important place in industrial processing. Screens which can treat suspensions of coarse particles at a very high flux and low cost will allow one to use spiral wound microfilters and UF membranes, the only low cost assembly which also allows for a suitable cross-flow. These microfilters could be used, in principle, to remove and concentrate algae, bacteria and other microorganisms, while the ultra-(or nano-filters) remove, separate out and concentrate proteins, while at the same time preventing fouling of the finer pore membranes of the electrodialysis (ED), reverse osmosis (RO) and similar types. The approximate flux of the latter systems are given in Table 1.

Table 1 FILTER, MICROFILTER, ULTRAFILTER AND REVERSE OSMOSIS SYSTEMS

	Type	Screen	Micro-filter	Loose UF	Tight UF	RO	Dense RO
	Opening	100 mesh .15 mm	.1-.3nm	3-10 nm	1-2 nm	none	none
Solute, Susp Solids	Hyd Perm /ft^2-psi		5 GPM -15	250 GPD -100	25 GPD -100	15 GPD -400	15 GPD -900



The attainment of the maximum flux for a given separation is all-important in terms of costs; in general, total costs and flux are inversely proportional, other things being equal. A reasonable flux and membrane life are <u>sine qua non</u> for practical membrane processes.

BIOMASS HARVESTING

Finely divided, suspended solids of an adherent nature have traditionally been difficult and expensive to concentrate. This is particularly true for the microscopic algae having diameters of from 2-10μ. They are the most rapid photosynthetic converters, but grow rapidly only to the point when their concentration is about 100-200 mg/l, and they are strongly adherent, rapidly blinding off the usual filter and membrane systems. It has be shown that non-fouling sulfonic acid p membranes of the ultrafiltration (UF) are capable of harvesting each of a ety of fresh water and marine organ- tested [3]. Figure 1 shows results f typical dewatering processes carried ut at the laboratory scale with UF mem- ranes of two different, high fluxes. At he present time we are engaged in making mall modules incorporating loose UF mem- ranes and microfilters for field testing 4].

embranes useful for algal concentration ust pass all dissolved solutes including roteins, but collect all suspended solids ncluding bacteria, viral bodies and algae; hey must also show high fluxes because of he stringent economic considerations hich obtain. We have estimated that the ost of a large scale algal harvesting rocess of this kind can be about 5¢/KG.

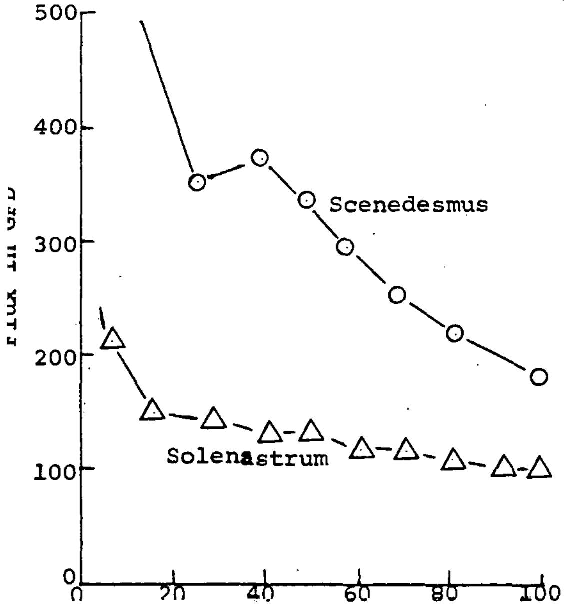


Fig. 2 UF of Scenedesmus (6x19μ) and Solenastrum (6x6μ) from ~115 mg/l to 3.9 and 0.94% TS respectively, at 50 psi in laboratory cell [3].

Processes having these kinds of capabili- ies and costs can be competitive with the oagulation-flocculation systems used in iological treatment processes, sewage reatment and the like, with widespread applications to several biomass harvesting systems. For example, they could be used to process and concentrate starch resulting from the processing of potatoes and he wet milling of corn. There are also substantial amounts of sewage sludge which could be dewatered and thus become a practical biomass source. A salient example of the potency of membrane processes as applied to biomass harvesting can be found in a recent study of Gregor and Lee [2] on the treatment of feedlot fermenter effluents. It has been shown by several investigators, particularly Hashimoto et al. [5], that when feedlot wastes are fermented for 5-9 days the process converts most of the nitrogenous material present into single cell protein, with methane as a byproduct. This single cell protein is an excellent animal feed, but ordinary centrifugation is too expensive and recovers only 20% of that available. Our studies have shown that feedlot digester effluents can be treated by membrane systems as follows: first by screens to remove gross particulates, and then by ultrafiltration to remove all dissolved constituents of $MW < 1000$ (containing ammonium salts among others), while producing a concentrate of about 17% total solids for direct refeeding. The salts in the permeate can be used directly as fertilizer, with or without concentration by RO. This study showed that all of these processes could be carried out at reasonable fluxes, and that no fouling of the RO membranes was observed once the pre-UF process (operating at about 20 GFD) had taken place.

The economic advantages of this membrane-controlled process are impressive. For a 10,000 cattle feedlot producing about 200 K GPD of effluent, the total daily process costs for screening, UF and RO was estimated at current prices for investment, labor power, etc. at about $330, the value of the methane recovered about $540 and the cost of the fermentation process itself (exclusive of the membrane process) at approximately $270. Thus, the value of the methane produced does not substantially offset the cost of the fermentation. However, the byproduct refeed credit was estimated to be about $6000; this makes this a process which could approximately double the profitability of feedlots.

These studies also showed that these digester effluents which foul conventional membranes in short order could be ultrafiltered adequately by the non-fouling sulfonic acid UF membranes, and once this had been accomplished, that the other membrane processes of RO and ED, each of which is highly susceptible to membrane fouling, could be carried out without any evidence of that phenomenon.

Comparable studies on biomass harvesting were carried out under the sponsorship of the Corn Refiners Association on the two major process streams of the corn wet milling industry, namely high solubles process water and oxidized starch filtrate. The former is a process stream of about 3% total solids containing substantial amounts of finely divided suspended matter, denatured protein and the like; the oxidized starch filtrate contains over 1% susp. solids as finely divided starch granules, notorious for membrane fouling. In every case tested the former could be treated by non-fouling UF up to 99% water recovery, the latter up to 98% water recovery with a high level of solids concentration; their permeates could be further concentrated by RO or the salts and other electrolytes removed by ED without problem [2]. An economic analysis shows that in a conventional plant producing about 280 K GPD of high solubles process water, the total membrane treatment including RO would cost about \$700 per day as contrasted with about \$3000 per day for the fuel alone for evaporative processes; the byproduct credit (as animal feed) here was about \$3000.

BIOMASS REFINING

Membranes have manifold applications in the refining of biomass. Certain of the more salient areas of application which are under study in our laboratories are described

The processes of fermentation as carried out in the conventional manner involve feeding the appropriate amount and concentration of feedstock into the fermenter and allowing the fermentation process to take place until the point of maximum conversion and concentration of the product, at which point product inhibition of the organism makes the process stop. Then the fermenter contents or beer is refined to remove the final product and concentrate the byproducts as stillage. While continuous or semi-continuous fermentation processes have been described frequently in the literature, none of these has as yet been adopted.

Our approach to the problem has been to employ membrane ultrafiltration (using membranes selectively permeable to solutes of molecular weight <1000) to remove the products of fermentation, while in the concentrate stream the suspended solids and cell are recycled or wasted. Our earlier studies were on the fermentation of glucose to methane and the organic acids [6]; more recently we have concentrated on the fermentation of glucose and other sugars to acetic acid. Figure 2 is a schematic of the general system employed, with the ionic products converted by water-splitting (WS) into their constituent acids and bases, where solvent extraction (SE) membranes remove the organic acids as their salts and RO concentrates the final products. We have achieved rates of conversion very much higher than those attained heretofore, at the same time obtaining high concentrations of products [7].

Under DOE-ANL sponsorship we are applying similar techniques to the fermentation of sewage sludge for the general purposes of its environmentally acceptable disposal and the production of organic acids, ammonia, methane and fertilizer [8].

When applied to the fermentation of sugars to ethanol by yeasts, the ultrafiltration process could reduce the time of fermentation by a factor of 2-3 by maintaining the concentration of alcohol in the product stream at not greater than 6%, a level at which little product inhibition is present. A semi-continuous fermentation could be carried out where periodically the process is terminated to minimize problems associated with foreign organisms.

We have carried out a series of laboratory studies on the membrane treatment of beer and stillage resulting from the fermentation of corn and molasses to ethanol. Both corn and molasses stillage could be treated effectively by UF, followed directly by RO in the case of corn stillage and with subsequent evaporation to raise the concentrate to 50% total solids, at which point the dissolved and suspended solids are suitable for use as animal feed. For corn stillage the total membrane process and evaporator fuel costs amounted to \$3.53/KG; when evaporation alone is used, the fuel cost for the evaporator is approximately \$12/KG. The byproduct credit for the animal feed amounts to about \$20/KG. In the case of molasses stillage which is high in ash and thus unsuitable for animal feed when subjected to conventional evaporative processes, ED was employed following the

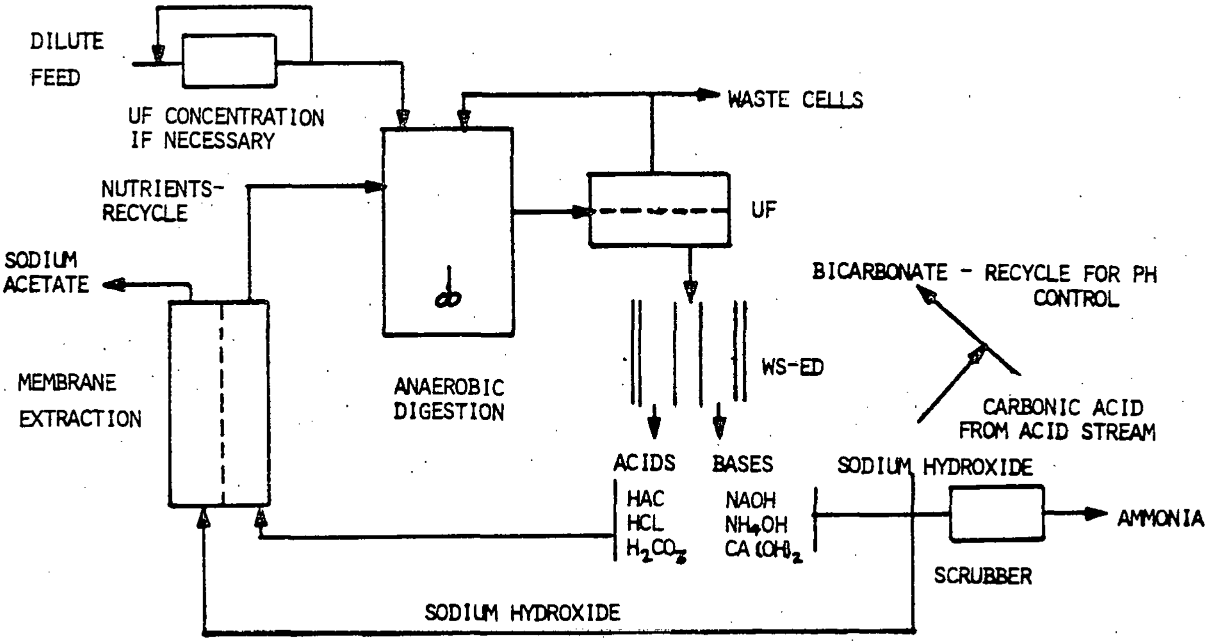


Fig. 2 Membrane-controlled fermentation process employing ultrafiltration to concentrate the feed and to remove cells and products from the fermenter, a water-splitting (WS-ED) process to convert all salts into acids and bases, and a solvent membrane extraction system to recover organic acids (as their salts) from the product stream [7].

step; no membrane fouling was encountered and a feed credit of $32/KG could be realized where none was available before; the total process cost here would be approximately $3.50/KG.

Subsequent experiments showed that fermenter beer itself could be equally well treated by UF to remove all protein and suspended solids, producing the animal feed concentrate directly; the clear permeate was passed directly into a beer still where it could be evaporated at relatively low fuel costs. The stillage in this case (from corn) was found to be largely a mixture of lactic acid and its salts at the 21,000 mg/l COD level, suitable for further concentration by ED, if so desired.

NEW REFINING PROCESSES

We have currently under development two new systems of the membrane permeation (MP) type. These contain membranes selectively permeable to one of two solvents present and are impermeable to most if not all of the other dissolved solutes present. MP processes are driven by imposing a vapor pressure gradient across the membrane, in effect distilling across it. We have concentrated upon two new kinds of membrane permeation systems, one selectively permeable to ethanol in the presence of water at low ethanol concentrations, the other selectively permeable to water in the presence of ethanol at high ethanol concentrations.

At the present time many organizations desire to enter the field of biomass conversion to make ethanol, using as feeds such diverse materials as corn, juice concentrates from cane, beet or sorghum, cull potatoes, etc. They plan to purchase conventional fermenters, beer stills and evaporators, and plan to improve and expand their plant capacities and capabilities as new technologies become available. Our program is designed to make available "add-on" membrane devices to achieve these ends. The flow diagram for the system we envision is shown in Fig. 3. First, one adds UF to the fermenter to treat the beer prior to the beer still, and also installs ED (where needed) and RO capacity to reduce the energy requirements of evaporation. The UF system will subsequently be employed to allow for the recycle of all low molecular weight solutes and products so as to increase fermenter capacity. We have already achieved some success in removing ethanol from the UF permeate at the 6% level and raising it to the 12% level in the

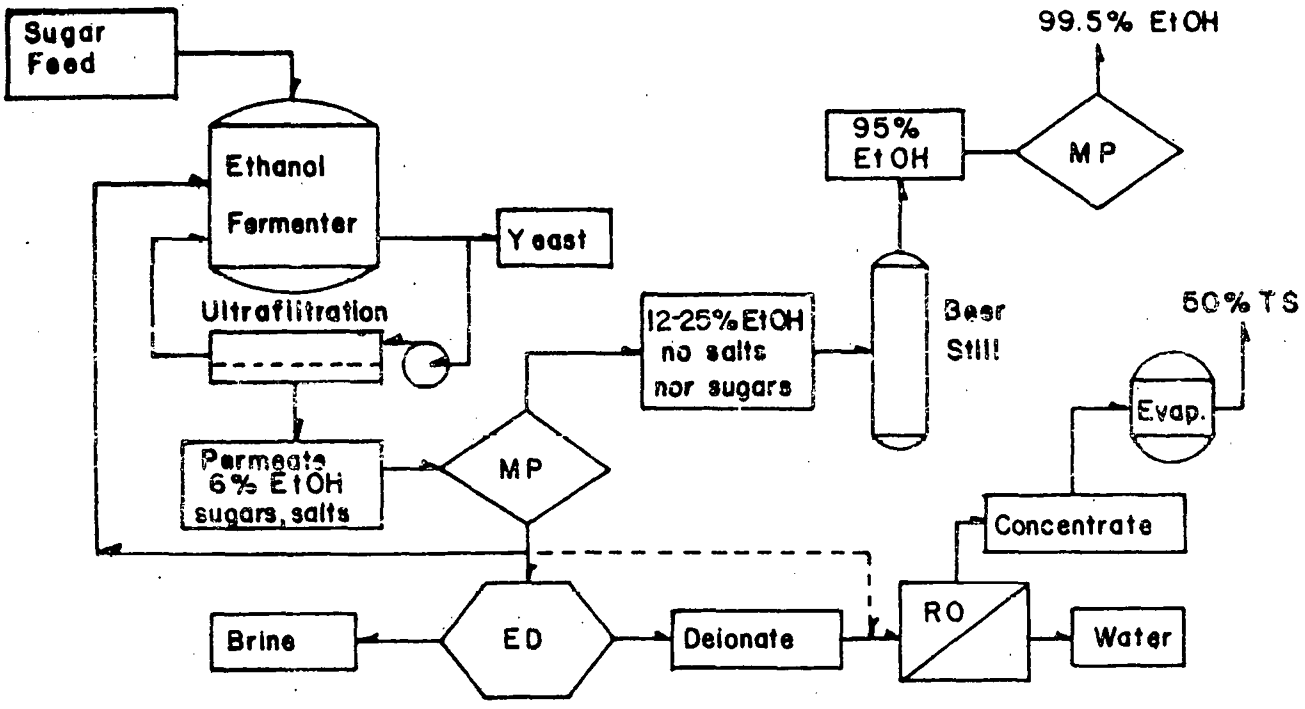


Fig. 3 Membrane-controlled fermentation of sugars to ethanol.

MP permeate, but fluxes are not yet at the desirable levels. If other organisms are employed which show ethanol inhibition at the 2% ethanol level, we expect to have MP systems to raise ethanol from that concentration to at least 6-8% in the MP permeate.

It has already been shown that distillation of the ethanol-water system from 6-8% ethanol to about the 75-80% level can be achieved at low cost both in terms of energy and equipment, but distillation to the azeotrope at 95% and further dehydration to gasohol (99.5%) are relatively expensive and energy-consuming. The conventional RO membranes, being selectively permeable to water as opposed to other solutes, can be used in the MP processes to reach and break the azeotrope, but their selectivities are relatively low as are their fluxes. We are making membranes specifically designed for the task, and expect to be able to lower both energy and investment costs for the dehydration process. The complete system shown in Fig. 3 should allow for the reduction in process costs from the present value of 60-70¢/gal [9] to about 15¢/gal.

We have also been turning our attention to the conversion of cellulose into sugars. Of particular interest in our program is the Bergius process [10] for wood hydrolysis, based upon the classical studies of Willstätter who showed that 35% hydrochloric acid dissolved hemicellulose into pentane sugars rapidly at room temperature and made the structure porous, so when followed by 41% HCl the hexose sugars (as oligomers) became soluble. The Bergius process was operated at several large plants between the two World Wars, and subsequently was abandoned because of the high cost of separating the acid and sugar components of the digestion liquors. Much of the acid could be removed by flash evaporation, but the remainder (approximately 24%) defied a convenient and inexpensive separation from the 40% sugars. This separation should be able to be accomplished now by the use of extremely "tight" ED membranes capable of removing the 24% HCl from the reaction and concentrating it to at least 35% HCl. Earlier studies on the ED concentration of sulfuric acid from sulfuric acid-water mixtures [11] showed that extremely tight ED membranes could accomplish this kind of separation for hydrochloric acid but not for sulfuric acid because of the relative sizes of the bisulfate and chloride anions compared to that of the hydrogen ion.

We are also studying the use of enzyme-coupled UF membranes [12] and affinity-

orption membranes [13] for the production
urification of the cellulase and cel-
se enzymes and their utilization in
onversion of cellulose into glucose.
s already been demonstrated that en-
yme-coupled E.coli galactosidase UF reac-
ors are highly efficient for the conver-
ion of lactose into glucose; it is antici-
ated that a comparable cellobiase reactor
ould serve for the conversion of cellobi-
se. An ultrafiltration-assisted sacchari-
ication process is under study employing
e new Rut 3C-30 strain of Trichoderma
eesei, with cellulose fed to the organism
a growth and hydrolysis reactor where
removes the cellobiose and glucose in
e permeate and cells and enzymes in the
oncentrate. The cells are filtered off
d the extracellular enzymes fed into a
accharification reactor (at 50°C) to pro-
ce glucose in the permeate [14].

e latter studies which involve the esta-
lishment of the feasibility of treating
eer and stillage by UF,RO,ED and MP sys-
ems, the dehydration of distillates by MP,
e separation of acid-sugar mixtures and
e use of membranes to facilitate cellu-
ase action are being supported, in part,
a new DOE-SERI program [15].

IOMASS PRODUCTION

mbrane technologies will play an impor-
nt role in most biomass production sys-
ems because they allow for the complete
ecycle of all dissolved salts and nutri-
ts. In alcohol production from corn or
lasses, as an example, all of the inorga-
c nutrients remain in the stillage con-
entrates at concentrations which make
eir return to the local soil convenient.
e same applies to the conversion of wood
to ethanol, by either the acid or enzy-
tic digestion process. Since most bio-
ss conversion systems are by their nature
te-specific, the local distribution of
rtilizer values is convenient.

veral studies, particularly those by
oldman [16], have shown clearly that if
rge scale biomass production is to become
reality, the amounts of fertilizer re-
ired will exceed that now produced in the
ited States by one or two orders of mag-
itude, aside from unacceptable cost consi-
erations. The availability of low-cost
rtilizer is of particular importance in
iomass production by silviculture and
m marginally productive lands. If thin soil such as is present in rain forest areas is to be utilized, closed-cycle, fertilizer-return systems are necessary.

ACKNOWLEDGMENT

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INCREASING THE BIOMASS PRODUCTION OF SHORT ROTATION COPPICE FORESTS

Klaus Steinbeck

Associate Professor, School of Forest Resources
University of Georgia, Athens, Georgia 30602

BSTRACT

n overview of the research project supported by ontract #EG-77-S-09-1015 between the U.S. epartment of Energy and the University of Georgia s presented. The objective of this project is to valuate various ways of increasing the biomass ields from short-rotation coppice forests established for energy production. To this end several ardwood species are being evaluated for their otential as biomass producers and growth responses weed control, mechanical cultivation, fertiliation and irrigation determined. Desirable provenances and individuals of candidate species are elected and tissue culture techniques for mass loning are being developed.

wo 16 hectare plantations have been established on ntrasting sites with soils that may be available r short-rotation forestry. *Platanus occidentalis*, *Liquidambar styraciflua*, *Alnus glutinosa* and *binia pseudoacacia* were planted in early 1978 and 79 either in pure stands or in mixtures of *atanus* or *Liquidambar* with one of the other two ecies, which are nitrogen fixers. The growth, ergy input-output and nutrient status of these antations is monitored. First seedling harvests e expected in 1981 and coppice rotations will en be initiated.

BJECTIVE AND APPROACH

e objective of this project is to evaluate rious ways of increasing the biomass yields from ort-rotation coppice forests established for ergy production. The major approaches to meeting is objective involve:

) *Species Screening*. Several hardwood species enable to short-rotation forestry are being anted. Species are being selected on the basis their juvenile growth rates, their capacity to oduce vigorous stump sprouts, their resistance to [illegible] and wildlife damage and their adaptedness to [illegible]ation management.

(2) *Cultural Treatments*. Growth responses to cultivation, mechanical and chemical weed control, fertilization and irrigation are being evaluted. This involves the establishment of test plantations on soils in Georgia which might be available for energy and fiber production and data collection over at least a five-year period.

(3) *Genetic Improvement*. Superior individuals and provenances of the most promising species will be selected and propagated. Selection will be based on biomass production rates, adaptability to a variety of sites, pest resistance and ease of multiplication. Tissue culture techniques for hardwoods are being developed with a view towards practical systems of mass cloning as well as rapid screening of the growth potential of clones.

(4) *Species Mixtures*. The energy input-output balance of plantations might be improved with reduced fertilizer inputs. Therefore, the biomass production of nitrogen fixing, woody species grown either in pure or mixed stands is being determined.

(5) *Ecological Assessment*. The test plantations offer an opportunity to compare the susceptibility of pure and mixed stands to pests. No formal treatments are planned in this regard and comparisons will only become possible where the opportunity arises naturally. Changes in soil fertility and organic matter status under pure and mixed stands, however, will be followed in a systematic fashion.

PROGRESS

The focal point of this project are two 16 hectare plantations in which American sycamore (*Platanus occidentalis*), sweetgum (*Liquidambar styraciflua*), European black alder (*Alnus glutinosa*) and black locust (*Robinia pseudoacacia*) were planted in pure and mixed plots (Fig. 1).

Both plantations were established on soils which might be available for short-rotation forestry. One of them is located in the Piedmont province of Georgia (Putnam County) on soils in the Cecil series with a small portion in the closely related Davidson series. Both of these soils series are

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is paper presents an overview of the ort-rotation research supported by contract G-77-S-09-1015 between the U.S. Department of ergy and the University of Georgia. This contract became effective on 16 June, 1977.

widely distributed in the Piedmont. Soil texture ranges from clay loam to sandy clay in the A and B horizons. Slopes range from 2 to 15 percent and are generally around 5 percent.

The second plantation is located in the eastern part of the Middle Coastal Plain and close to the Atlantic Coastal Flatwoods Region of Georgia (Tattnall County). The soils on this site probably are mostly in the Stilson series. They are moderately well drained, low to moderate in natural fertility and acid throughout. They are of loamy sand texture and the site is essentially flat.

Planting of seedlings commenced in January, 1978 but lack of commercially available seedlings prevented completion. Therefore, approximately half of both plantations were not planted until early 1979. There was considerable variation among species in seedling size at planting (Table 1). Sycamore and alder were taller in both years than sweetgum and locust. Planting speed of a single row, modified wildland planter drawn by a Case 850 crawler averaged about 900 seedlings per hour at a 1.3 x 2.4 m spacing. This included turn around time, seedling loading, and breakdowns. Sometimes the Coastal Plain site became so wet that machine travel was impossible. Under such adverse conditions hand planting rates averaged about 500 seedlings per man day.

TABLE 1. HEIGHTS AND ROOT COLLAR DIAMETERS OF SEEDLINGS AT PLANTING

Planting Season	Species Sycamore Ht.	 Dia.	 Sweetgum Ht.	 Dia.	 Alder Ht.	 Dia.	 Locust Ht.	 Dia.
	---------	----	-----cm	-----	----	----	----	----
1978	60	1.3	45	0.7	72	0.6	--	--
1979	86	1.5	41	0.6	58	0.7	32	0.8

Fertilization on both sites was delayed until the summer after planting so that the seedlings had a chance to develop a root-system and could therefore utilize the applied nutrients better. In July 1978, 900 kg/ha of 10-10-10 fertilizer was banded in 90 cm wide strips over the seedling rows. This was followed in September with a banded topdressing of 135 kg of elemental nitrogen per hectare in the Piedmont and 90 kg of elemental phosphorus per hectare in the Coastal Plain. This topdressing was applied to half-plots (Fig. 1 shows topdressed half-plots in Piedmont).

The two plantations posed contrasting weed control problems which were in large part due to the difference in the past land use of the two areas. The Piedmont site had supported a fully stocked stand of an upland pine-hardwood mixture in which 30- and 60-year age classes predominated. Weed competition in the new plantation was not severe and consiste mainly of sprouts from hardwood roots not removed during site preparation, vines of greenbriar (*Smilax* sp) and passion flower (*Passiflora incarnata*), scattered individuals of poke weed (*Phytolacca americana*) and beggar ticks (*Desmodium* sp). Weeds were controlled on this site by single pass, between row harrowing on two occasions during the growing season and some hand-hoeing within the rows.

The Coastal Plain site, on the other hand, was an old soybean field and an abandoned Christmas tree plantation in the past. The weed community was much more herbaceous than on the Piedmont site. Numerous species grew, the most competitive among them were Johnson grass (*Sorghum halepense*), Bermuda grass (*Cynodon dactylon*), nutgrass (*Cyperus rotundus*) and ragweed. Control attempts included mowing between the rows on two dates and rototilling, also on two occasions. Handspraying of the heaviest grass infestations with Roundup (Tradename of Monsanto Chemical Co., active ingredient an isopropylamine salt of glyphosphate) controlled the grasses well but also proved fatal to the trees, even though the spray had been directed away from them. The growth of ragweed became so rank that locating the trees from the tractor seat became impossible and control attempts were abandoned after early August.

In 1979, Princep (Trademark of CIBA - GEIGY, active ingredient simazine, 2-chloro-4,6 bis (ethylamino)-s-triazene) was broadcast sprayed over the Coastal Plain site and over the 1979 planted portion of the Piedmont site. This pre-emergence herbicide was applied as the buds of the seedlings began to swell at the rate of 4.0 kg/ha of active ingredient (a.i.) in the Coastal Plain and 3.7 kg/ha a.i. in the Piedmont. Some seedlings, especially black locust, showed herbicide damage (leaf scorch) but few (less than 1% overall) were killed. As of May, 1979 the simazine treatment is still effective and may continue to last into July or August, dependin on the weather.

A portion of the Piedmont plantation (Fig. 1) was irrigated with a self-propelled water cannon durin the droughty summer of 1978. An average of 2.5 cm water was applied on each of 10 different dates. This first year irrigation resulted in small heigh and diameter growth increases, but did not affect survival in any consistent way (Table 2, end of text). The Coastal Plain site was not irrigated because the water holding pond dried up.

Black locust grew from the smallest tree at planting (Table 1) to the largest at the end of the first season (Table 2). All species grew less in the Coastal Plain, this was in part due to the severe weed competition and also partly to site differences. Only half of the European alders survived after the first growing season in the Coastal Plain.

The energy input-output balance is of critical i portance for any biomass for energy project. I this project direct energy inputs for site preparation and cultural operations are recorded and

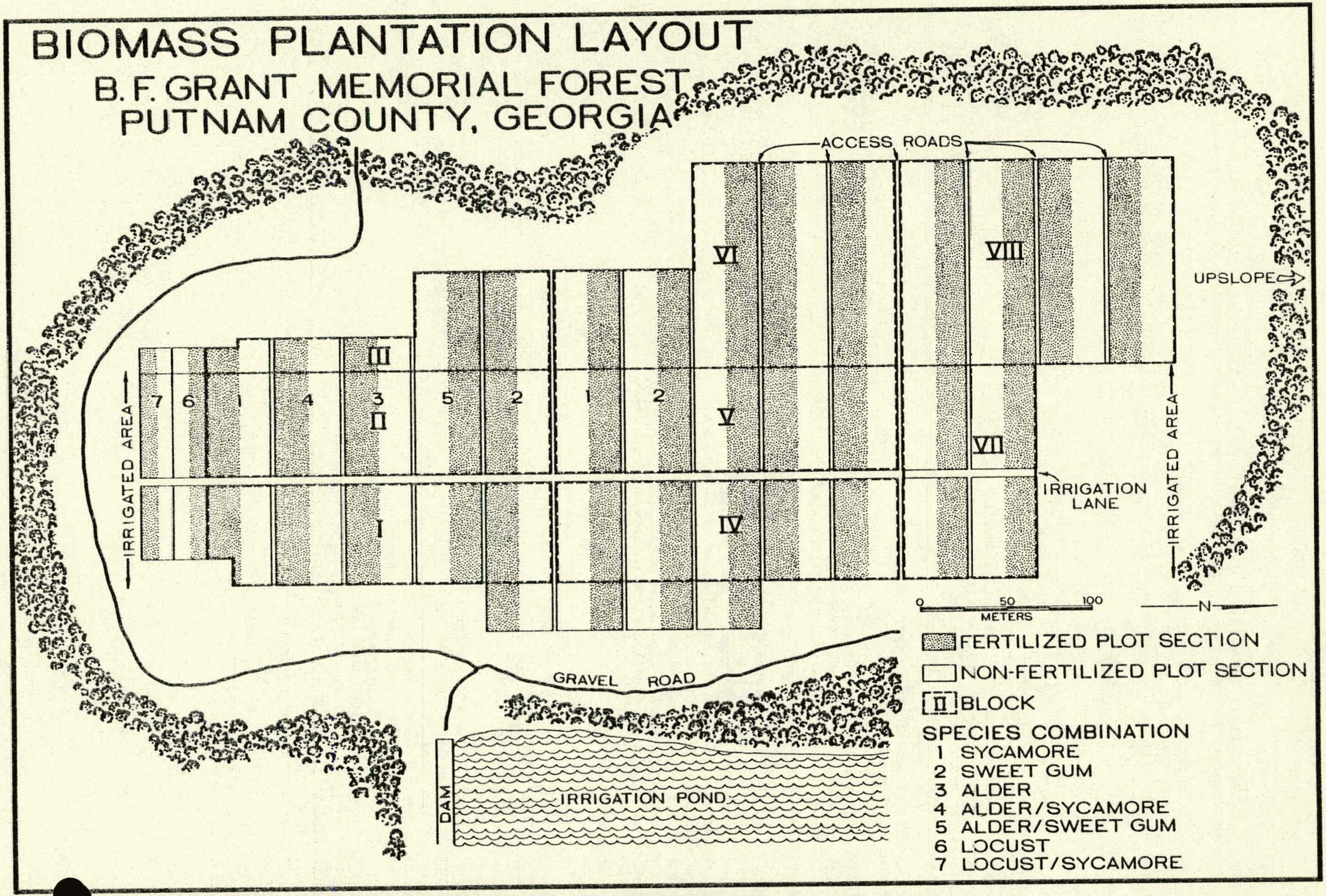
ble 3 (end of text) lists some diesel fuel in-
in our cultural operations. These figures
ould be regarded as a very rough approximation
cause they originate from two small scale opera-
ons with not necessarily optimum equipment sizes.
anting consumed the highest quantity of diesel
el and the 11 liter/ha difference between the
o sites reflects soil texture (sandy clay vs.
amy sand) and slope (2 to 15% vs. flat) differ-
ces. The same equipment and crew planted both
tes. If one assumed a first year cultural
hedule of planting, one fertilization, one her-
cide spray and two single pass harrowings as
asonable, a direct #2 diesel fuel input to drive
e machinery during these operations of 50
ters/ha on heavy soils and 40 liters/ha on light
xtured soils may serve as a first approximation.
rigation expenditures should be less dependent
soil texture, but were high at nearly 33
ters/ha used to apply 25 cm of water. These
ergy inputs will decrease to near zero in the
llowing seasons until harvest.

the genetic improvement program, seedlots of
erican sycamore were collected from various
ands in Georgia in order to identify provenances
th desirable characteristics.

ing tissue culture techniques, root and bud
fferentiation of several hardwood species was
tained. Plantlets of sweetgum and black locust
ve been grown. However, the use of solid media
single culture tubes restricted the number of
plants that could be handled at one time. There-
re the feasibility of mass cloning in liquid
lture to bring about controlled differentiation
numerous intact plants in single containers is
ing investigated.

TURE PLANS

restry research involves long lived organisms
d therefore requires time. In 1981 the first
rvest will take place and the coppice rotations
ll be initiated. The yields from this harvest
ll not be as high as those that can be expected
om subsequent coppice. The growth responses and
trient status of the plantations will continue
be monitored and additional species screening
ials will be established on various sites. The
arch for superior genetic material in promising
ndidate species will be extended and their mass
oning techniques developed. In 1981, we hope to
tablish an outplanting with plantlets produced in
ssue culture. The energy input-output of short
tation forests, the effects of harvesting at
rious seasons of the year on subsequent sprout
owth and the storage properties of harvested
terial all are of interest.



BIOMASS PLANTATION LAYOUT
B. F. GRANT MEMORIAL FOREST
PUTNAM COUNTY, GEORGIA

TABLE 2. AVERAGE HEIGHTS AND DIAMETERS OF THE TREES AT THE END OF THE FIRST GROWING SEASON

Site	Species											
	Sycamore			Sweetgum			Alder			Locust		
	Sur.	Ht.	Dia.[1]	Sur.	Ht.	Dia.	Sur.	Ht.	Dia.	Sur.	Ht.	Dia.
Piedmont												
Non-irrigated	83	130	1.8	87	52	1.1	81	73	1.1	86	186	2.0
Irrigated	80	133	2.0	77	59	0.9	84	76	1.2	84	221	2.7
Coastal Plain												
Non-irrigated	70	71	0.8	69	49	0.5	50	100	1.3	--	--	--

[1] Survival in percent, heights and diameters (15 cm above groundline) in cm.

TABLE 3. DIRECT FUEL EXPENDITURES FOR CULTURAL OPERATIONS DURING THE INITIAL GROWING SEASON (1978)

Operation (Equipment Used)	Number of times/yr	Site	
		Piedmont	Coastal Plain
		Total liters of #2 diesel/ha	
Planting (Case 850 crawler, single row planter)	1	37.5	26.5
Fertilizer spreading (70 SAE hp tractor, modified spinner)	2	7.3	7.1
Mowing (70 SAE hp tractor, 6' rotary cutter)	2	-	5.5
Rototilling (70 SAE hp tractor, 6' wide tiller)	2	-	13.5
Harrowing (40 SAE hp tractor, light harrow)	2	6.3	-
Irrigation - 2.5 cm H_2O/application (Pump driven by 70 SAE hp motor)	10	32.8	-
Herbicide spray[1] (70 SAE hp tractor, 30' boom)	1	2.6	2.5 (est.)

[1] Herbicide was sprayed in second season (1979).

NOTES

FUELS FROM MICROALGAE

John R. Benemann
Sanitary Engineering Research Laboratory
University of California, Berkeley
Berkeley, California 94720

BSTRACT

wo microalgal systems for the production of fuels
ere studied: methane fermentations of sewage-
rown microalgal biomass and biophotolysis,
sing blue-green algae. Although microalgal
onds have been used in wastewater treatment for
everal decades, the high cost of harvesting micro-
lgae is limiting the applications of sewage oxi-
ation ponds. A novel, low-cost process for micro-
lgal harvesting, based on bioflocculation in a
atch settling pond, has been demonstrated over the
ast year at the pilot quarter-acre (0.1 hectare)
cale using a shallow, paddle wheel-mixed pond. It
ppears that such a process can meet both the re-
uirements of wastewater treatment, that is, an
verage 85% removal of suspended algal solids and
f low-cost microalgal biomass production. The
ey technical problems in designing large-scale
icroalgal biomass systems for fuels-chemicals pro-
uction are discussed. The production of hydrogen
rom water by nitrogen-fixing heterocystous blue-
reen algae has been demonstrated in both the labora-
ory and with an outdoor converter; the relative
dvantages of alternative biophotosynthesis systems
re considered.

NTRODUCTION

e first report that blue-green algae produce
ydrogen dates to the end of the last Century (1).
icroalgal cultivation for the production of methane
uel was first suggested over twenty-five years
go (2). Thus, like most ideas in the fuels from
omass area, the use of microalgae for solar
nergy conversion is not a new one. Indeed, an
xperimental demonstration and conceptual analysis
an algal-methane system was carried out twenty
ears ago by Oswald and Golueke (3,4) in the first
etailed presentation of an "energy farming"
roposal. Work on the production of fuels from
croalgae was resumed in the mid-1970's (5-8),
nder N.S.F., E.R.D.A, and D.O.E. sponsorship, by
rofessor Oswald and the author and assisted by a
umber of graduate students. Last year's pre-
ntation at this meeting reviews the progress up
then (9). The key objective of this research
s to develop microalgal systems for fuel and
rtilizer production, integrated with wastewater
eatment. Such a system would have near-term
plications. Longer-term applications of micro-
gae in photosynthetic hydrogen production--
iophotolysis" (10,11) and large-scale production
fuels and chemicals (12)--were also pursued by
is research group. A brief update of recent
ess is presented, based on data reported during the past year (13,14,15). The key technical problems in microalgal biomass-fuel production are discussed, based on a recent assessment by the author (16).

HARVESTING SEWAGE-GROWN MICROALGAL BIOMASS

The objective of the research in microalgal biomass production was to demonstrate maximum, sustainable productivities and low-cost harvesting technology with pilot quarter-acre scale (0.1 hectare) ponds. Prior research (5,7) used small, shallow, paddle-wheel-mixed, sewage-fed algal ponds to test microstrainers (fine mesh, rotating screens with backwash) as a harvesting technology for the microalgae. The objective had been to establish which pond operating parameters would result in the establishment of large colonial microalgal populations that were readily harvested by 26 μm opening screens, selected as the most effective mesh size. The result of this research demonstrated that the pond operations--primarily long detention times--which resulted in microstrainable algal populations also resulted in low biomass productivities. Also, the small-scale ponds used in most experiments, 3 to 10 m^2 in size, proved to be highly susceptible to zooplankton predation and, thus, resulted in unstable cultures. Such instabilities appeared to be less frequent in the large-scale two-thirds acre (0.25 hectare) pond operated at the same time. A detailed cost analysis of microstrainer harvesting revealed that, even if the problems of culture instability and low productivity could be overcome, it would still be too expensive for low-cost microalgal biomass production in absence of wastewater treatment credits.

For these reseasons the research effort was redirected to developing a simple bioflocculation-settling process using a 2-day detention time, batch-isolation pond, which was demonstrably of low enough cost to allow inexpensive microalgal biomass production (12). Prior work on a longer (2 week) isolation pond process for microalgal removal from conventional (deep, unmixed) sewage oxidation ponds had indicated that effluents from "high-rate" (shallow, mixed) ponds would flocculate and settle very rapidly (17). Experiments with small (10 m^2) ponds demonstrated a rapid, spontaneous flocculation and settling of microalgae, with removal efficiencies above 95% (on a chlorophyll basis) after one day in a one-liter Imhoff cone. Removal efficiencies were affected by detention times, with short detention times correlating with poor settling (14). Nevertheless, the detention times required for obtaining good removal by settling still gave fairly high productivities. Indeed, in the summer and fall achievement of good flocculation-settling did not appear to require a significant compromise with productivity.

To avoid the above-discussed problems of culture instability in small-scale ponds, subsequent to August 1978, all experiments were carried out with two 0.1 hectare paddle wheel-mixed ponds. These have been operated since that time under varying conditions of detention time, depth, and mixing speed to establish the optimal conditions to achieve both a readily flocculating culture and high biomass productivity. Results obtained with one of the 0.1-hectare ponds during September-October 1978 are shown in Table 1. A very high degree of algal removal was observed, even though a low detention time was used and a relatively high productivity was achieved (13). The observations with a 24-hr Imhoff cone settling correlated quite well with those made with a 32,000-liter 48-hr pilot settling pond. These data indicate the basic feasibility of this process. It must be pointed out, however, that during the winter of 1978-79 the settling process was not continuously effective, with some periods of upset experienced (18). The proper pond operating conditions for year-round effective performance of such a microalgal removal process are not yet established, although existing experience would indicate that it is feasible. This should be established by the ongoing experimental program (18). Although reliable and effective (approximately 85% removal), year-round performance of the microalgal separation process is critical in a wastewater treatment system, otherwise effluent standards (30 mg/ℓ suspended solids) will not be met; it is of lesser importance in a biomass-fuel production system where wintertime operations would be of relatively little interest and where effluent recycle would be required to conserve water and nutrient resources.

The microalgal sludge accumulated in the 32,000-liter settling pond contained 1-1.5% volatile suspended solids. This is a very low concentration for normal anaerobic digestion operations. Nevertheless, laboratory digestion studies were carried out to determine the suitability of freshly harvested microalgae as a feedstock for methane production (19). The results indicate that microalgae digest at about the same rate and extent as primary sewage sludge. To overcome the drawback of the low sludge concentration, either an anaerobic filter-type process or a low-cost, covered-lined, unmixed earthwork digester must be designed and tested. In sewage treatment applications it may be cost-effective to further concentrate the microalgal sludge, but this may not be possible for large-scale biomass-fuel systems.

The wastewater treatment process studied with the above system is only the first, but currently the only required, step in the cleanup of wastewaters. Although effective biological oxygen demand (BOD) and suspended solids removal could be achieved by such a system, the remaining nutrients in the wastewaters would still cause a significant amount of algal growth and eutrophication in any open body of water receiving such effluents. Pond systems are generally effective in nitrogen removal

TABLE 1

SUMMARY OF SEPTEMBER-OCTOBER 1978 OPERATIONS OF THE WEST 0.1 HA POND

Parameter / Date	9/3-9/9	9/10-9/16	9/17-9/23	9/24-9/30	10/1-10/7	10/8-10/14	10/15-10/21	10/22-10/28
Detention Time (days)	2.8	2.6	2.7	2.7	--	2.7	2.4	2.4
Pond VSS, mg/ℓ	224	191	177	217	211	188	204	182
% Chl. a, (of VSS)	1.93	1.42	1.75	2.04	1.55	1.33	1.76	1.67
Total VSS Production g/sq m/day of VSS	16.6	14.8	13.3	16.2	--	13.9	17.2	14.9
24-hr Settling Cone								
% VSS removal	89	92	88	95	98	92	85	80
% chl. a removal	92	94	96	97	99	80	91	98
32,000-ℓ Settling Pond								
% VSS removal	--	92	--	92	--	92	70	78
% chl. a removal	--	93	--	97	--	93	71	75

Pond operations were carried out with a 100 m^2 high-rate pond fed with settled domestic sewage during the daytime and mixed with paddle wheels (about 15 cm/sec). Operating depth was 20 cm. Data are average of two data points per week. The 32,000-liter settling pond was operated on a batch basis with a 2-day detention time.

outgassing of ammonia and uptake of nitrogen gal cell mass. However, phosphate removal ids is not complete, and this is the limiting ent for the nitrogen-fixing blue-green algae ich are the largest problem in eutrophication. is project has investigated the controlled culure of nitrogen-fixing blue-green algae for phosate removal from wastewaters (6, 7, 19). The sults indicated that such a "tertiary" treatment ocess is feasible (19). However, expected proctivity of the nitrogen-fixing microalgae would ly be about half of that achievable with green lgae (14). Temperature is a critical parameter the cultivation of this type of microalgae.

RGE-SCALE MICROALGAL BIOMASS SYSTEMS

recent assessment by the author (16) of the U.S. tential of energy from municipal-industrial stewater aquaculture systems concluded that total ng-term energy impacts from tertiary treatment stems would be limited to approximately 1 x 10^{14} U/yr (current U.S. energy consumption is almost x 10^{16} BTU/yr). This relatively low impact due, in part, to the assumptions underlying is estimate: geographic restriction of such stems to half of the U.S. population, maximum rket penetration of 50% due to alternative mpeting waste treatment technologies, a subantial decrease in per capita phosphate detergent nsumption, and only a minor contribution from dustrial wastewater flows. In particular, agriltural, confined animal, and food processing stes were excluded from this assessment, as they re estimated to have greater applications to oductions of animal feed than fuels. Neverthess, a 1 x 10^{14} BTU/yr energy contribution from stewater aquaculture systems is probably an per bound for any realistic long-term implentation plan. Despite this limitation, wasteter aquaculture systems are the most immediate plications for aquatic plant biomass-energy stems. More importantly, the potential economic d environmental benefits of such systems far tweigh their energy contributions (16). However, ch limited energy impacts do not justify a major phasis on wastewater aquaculture systems by D.O.E. d an examination of the potential of microalgal omass systems designed specifically for energy oduction is of importance.

e objective for the development of any biomassergy system should be an optimization of yield rsus cost. With microalgae, the high capital sts of even the simplest algal production pond d harvesting system (12) result in high fixed sts and, thus, make yield the most important ctor. A detailed review of the theory of perience with microalgal biomass production has en published (20) and need not be covered here. at review demonstrates an almost total lack of ta for microalgal productivity under defined contions in sufficiently long-term experiments and rge-scale systems. The best data is with sewage idation ponds in which conditions of nutrient mitations and other limiting factors (turbidity) t be excluded. Data from the large-scale se microalgal production systems is second hand. The theoretical analysis is limited by the paucity of light-saturation data for microalgae. However, yields between 25 and 50 tons/acre/yr (55 and 110 metric tons/hectare/yr) can be forecast for favorable U.S. locations, based on practical experience and theoretical considerations. These are very high productivities when compared to land-based biomass production systems. They have not yet been achieved in low-cost production systems and, in practice, only the lower end of this range may be actually obtainable.

The cost of microalgal biomass production was estimated in some detail a year ago by the author (12), an estimate which formed the basis for the land-based portion of a D.O.E.-sponsored cost analysis of aquatic biomass systems (21). The key assumption made in that exercise was a microalgal harvesting system, based on a two-day batch flocculation-settling process; a process which was at that time completely speculative but has since been demonstrated in practice, as discussed above. Thus, it may appear that the result of that cost analysis, which estimated microalgal biomass production costs of as low as $50/ton (dry weight), could be applicable at present. However, there were a large number of other favorable assumptions, from very wide channels to very low mixing power inputs, whose combined effect was to make that estimate rather unrealistic. A greater effort at engineering design and cost analysis may be justified to allow arriving at a more realistic figure for the cost of microalgal biomass production. However, first, several specific uncertainties must be resolved.

One of the most important issues in microalgal biomass production concerns nutrient supply. The principal nutrient is carbon. Unlike other plant systems, microalgae in mass culture cannot take up sufficient carbon dioxide from air due to the limitations of mass transfer across a water interface. Power requirements for carbon dioxide supply to an algal culture by sparging air through it would be prohibitive (21). Thus, an enriched source of carbon dioxide is required, a fact which will limit microalgal biomass production to situations where such is available at little or no cost. Although stationary flue gas (about 10% CO_2) sources are plentiful, the matching of these, in both location or size, with the requirements of microalgal biomass systems may be difficult. This problem led to the consideration of the potential of the "Photosynthetic Energy Factory," in which combustion of a land-based energy crop provides the carbon dioxide required by an algal pond (8).

A key question is how to introduce the carbon dioxide into a pond system by the most cost-effective route. Active or passive transfer may both be considered, in one case a power consumption penalty, in the other increased capital costs are paid. One suggestion was to cover the paddle wheel combining the two approaches (12). The transfer coefficient for the system cannot be determined a priori, as it cannot even be for oxygen transfer in conventional fermentation systems--a much better studied situation. However, it does not appear that carbon dioxide transfer is a significant technical or economic problem. This issue should not be confused with the question of carbon dioxide supply to the photosynthetic apparatus of the algae. Intracellularly algae possess carbonic anhydrase--the enzyme with the largest turnover number known--which is able to maintain the equilibrium carbon dioxide levels dictated by intracellular pH and alkalinity. Extracellularly, pond alkalinity and pH dictate the bicarbonate and carbon dioxide concentrations. By properly regulating the rate of carbon dioxide supply, pH control can be achieved and thus optimal growth conditions maintained for the algae under cultivation, within the limitations of the growth medium used. Thus, carbon supply is not a question of algal physiology but strictly an engineering problem; one which can be readily solved wherever a carbon dioxide supply is available. It should be noted that pH control is also desirable to prevent undue ammonia volatilization or phosphate precipitation. Indeed, the high concentrations of nitrogen and phosphorus in microalgae is likely to make their supply, rather than that of carbon, economically limiting in microalgal biomass production.

The above are only some of the systems engineering considerations in designing a large-scale algal biomass system. Until they, and many others, are addressed, a more realistic cost analysis will not be possible. At present, a minimum production cost of about $200-$300/ton dry weight is likely (16), the potential for reducing this cost by future scientific/engineering advances is undertain. However, even within this cost range it may be practical to produce microalgal biomass for its energy content, as long as a significant fraction of the biomass produced is in the form of relatively valuable petrochemical substitutes. Indeed, the ability of microalgae to produce a variety of commercially interesting, energy-intensive products is one of the primary reasons that microalgal biomass systems have significant near- to mid-term potential. Among the algal products of interest are fats and oils, polyols (including glycerol), and polysaccharides. A greater effort is required to develop chemicals production from microalgae cultivated on brackish or saline waters which are not suitable for conventional agriculture. The ability of microalgae to tolerate high and varying salt concentrations is one of their important attributes. In both these respects, salt tolerance and chemicals production, microalgae have advantages over other aquatic plants (e.g. water hyacinths, marsh plants) whose chemical composition is not as variable and salinity tolerances not as broad. It should also be recognized that the ability of such emergent aquatic plants to extract carbon dioxide from the air is coupled to a high rate of water evaporation, which may be a limiting factor in their production.

If the key technical/economic issues surrounding microalgal biomass production were resolved, their potential contribution to U.S. energy supplies would still be uncertain, as the applicable water and land resources have not been estimated. In California alone about 100,000 acres of microalgal ponds could be operated using agricultural drainage waters, whose use and disposal would accrue a significant credit (16). Throughout the southwest, where climate is suitable and land available, significant amounts of brackish water exist, which may be used in microalgal cultivation. However, the potenti resource base is uncertain, and microalgae must considered, on the whole, a minor, though not negligible, future energy source.

HYDROGEN PRODUCTION WITH MICROALGAE

The possibility of producing hydrogen fuel from water and sunlight using a biological catalyst ("biophotolysis") has been studied for about six years. The only system taken, thus far, to a sta where a truly biophotolytic reaction has been demonstrated is based on nitrogen-starved, heterocystous, blue-green algae (10,11). The basis of this system is the oxygen protected nitrogenase reaction localized in the heterocysts of these al As is well known, nitrogenase will produce hydrog in the absence of molecular nitrogen; thus, it is relatively simple to show hydrogren production (b heterocysts) and simultaneous oxygen production (by vegetative cells) when a culture of such alga is sparged with argon. By maintaining such a culture for prolonged periods of time under argon, hydrogen and oxygen are produced continuously, and catalytically, for several weeks, both in the laboratory (22) and outdoors (23). A small amoun of nitrogen (N_2) helps maintain culture stability Achieved photosynthetic efficiencies have been up to 3% in the laboratory and 0.2% outdoors (11, 15 The development of this system will require a comprehensive understanding of biochemical pathways and physiological adaptations of these organisms. Studies carried out by this project involved the purification of nitrogenase and hydrogenase, nitrogenase regulation and localization, and the mechanisms of oxygen protection of nitrogenase. Eventually, species selection (particularly of thermophilic strains) (24) and genet improvements will be necessary. The achievement a 3% conversion efficiency outdoors should be achievable, and is the minimum that can be considered for practical applications.

A vertical array of glass tubes has been proposed as a possible biophotolysis converter (15). The algae would remain suspended by gas mixing, with the inert gas recirculated until enough hydrogen has built up in the gas phase and recycled after the hydrogen (and oxygen) have been used or separated. The design of such a system is technically fairly straight forward. However, it economics will be, at best, marginal. Although the materials required are of low cost (the glass tubes, themselves, cost less than $\$2/m^2$) the valu of the hydrogen produced is also low (less than $\$1/m^2/yr$), resulting in only very minimal allowab assembly, installation, and operational costs for the converters. For this reason alternative systems of biophotolysis are of interest, systems capable of producing a higher value fuel--that is pure hydrogen not mixed with oxygen--or more hydrogen per unit area. A number of such systems have been proposed (25) and their relative merits recently discussed (26). The use of photosynthet bacteria for the breakdown and "photofermentatio of low nitrogen wastes has the nearest term potential. Rates of hydrogen production from a organic substrate would be about ten times higher per unit area, then hydrogen production from water. ddition, some waste treatment credits would ue to such systems. Two approaches are pos- e: selection for or genetic construction of otosynthetic bacterium that is capable of breaking down and metabolizing the organic wastes or the establishment of a mixed microbial culture that can accomplish the same in a symbiotic association. Some initial work in this area has been carried out (15). Other concepts include separating the oxygen and hydrogen reactions spatially or temporally. The advantage of such systems is that a pure hydrogen stream would be generated, which has a higher use and value than a mixed H_2/O_2 gas phase. Biophotolysis is a long-term research option.

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Session II B

FUELS FROM FERMENTATION OF BIOMASS

Henry R. Bungay
Rensselaer Polytechnic Institute
Troy, New York 12181

st of the people working on fermentation of bio-ss feel that other technologies are better for rge central power stations but biomass is the in hope for the future for alcohol fuels and for trochemical substitutes. In other words, coal photovoltaic devices or windmills or something se may turn out better on a dollar per BTU basis, ile biomass will be the winner for high-value ansportation fuels and chemicals.

me biomass components ferment easily and others st first be hydrolyzed. Starchy plants such as rn, manioc, and potato are readily extracted for arch which can be fermented by microorganisms. garcane, sugar beets, and sweet sorghum have gary juices which are excellent fermentation sub-rates. The cost of starches and of fermentable ices is high because of their use in foods or in nventional fermentations, thus lignocellulosic terials have been given prime consideration as edstocks for biomass fuels. However, the pro-sal by Lipinsky to ferment sweet sorghum juices d to take byproduct credits for its excellent ber is leading to more emphasis on sugary plants. e Purdue group has an analogous idea for extrac-ng pentose sugars from corn stover and using the sidue for fuel or feed.

gnocellulosic materials may come from solid stes or any crop stalks and branches. Readily ailable sources are trees and corn stover. A ugh analysis of a typical material is shown in g. 1. Using only the cellulose in biomass is onomically unattractive because raw material sts are too high for producing ethanol at a com-titive price and expensive waste treatment would required for hemicellulose and lignin. If good es can be found for the other components, ligno-llulosic biomass will be a highly attractive edstock. There is so much lignocellulose from many diverse sources that price stability is sured.

llulose is difficult to hydrolyze because it is otected by lignin and has a highly order crystal-ne structure. Pretreatment is essential to get od hydrolysis yields. The main types of pre-eatment are in Fig. 2. Acid hydrolysis was ough to be limited to yields of about 55 per cent cause of competing reactions for decomposition of e sugar. Enzymatic hydrolysis can give almost antitative yields of glucose. However, pretreat-nt helps acid hydrolysis almost as well as it does enzymatic hydrolysis. For initial ventures for biomass fuels, it may be better to employ acid hydrolysis and later on to use enzymes alone or in conjunction with acid. Until the recent advances by D.O.E. contractors in achieving high titers of cellulase enzymes, enzyme cost was a main item in converting biomass to alcohol fuels. There are now many runs in shake flasks and in stirred fermenters with enzyme titers of 10 times those of two years ago, and cost of enzyme is becoming quite reasonable.

Perhaps the best way of providing an overview of the DOE fermentation program is to show what is ready for commercialization, what is being tested, and what research should be initiated. A factory built now to be competitive with petroleum-based ethanol should not use lignocellulosic biomass because there are no reliable markets for hemicellulose and lignin in the quantities that would be produced. The factory could use corn grain, molasses, or any easily fermentable carbohydrate, but the fuel alcohol will cost over $1.50 per gallon. Looking through the D.O.E. contractor reports, there are not many ideas that are ready right now to lower this cost. Groups at Battelle and at Purdue have pointed out ways to get cheap sugars from sweet sorghum or from other biomass by taking the easily extractable material and finding uses for the insoluble residues. Since raw materials make up at least 2/3 of the cost of manufacturing ethanol, these suggestions could impact rapidly on the economics of fuel alcohol.

Turning attention to what is almost ready, the efforts of D.O.E. contractors seem remarkable. The order of discussion will try to approximate the probable chronological sequence of commercialization. A very new and pregnant observation at Purdue has great promise and overriding importance. They have an alcohol recovery scheme that may cut distillation energy requirements to one-tenth of that of today's factories. This is based on using a short column to distill to an intermediate ethanol concentration and removing the rest of the water with drying agents. This will dispell the criticism that fuel alcohol processing takes about as much fossil fuel energy as the energy content of the product. Earlier schemes depended on burning residues to generate process heat, but a breakthrough in recovery of ethanol will allow us to find better uses than burning for side fractions.

Another chance for early commercialization is fermentation of the sugars from hemicellulose. Several contractors have developed processes for fermentation of pentose sugars to organic chemicals such as ethanol, acetone, butanol, and 2,3-butanediol. Common strains of yeast do not ferment pentoses, but other organisms utilize these sugars well. The acetone/butanol fermentation does not give high yields because butanol is toxic; recovery costs are too high because the solutions are dilute. Ethanol from pentoses is beginning to reach good yields, and a commercial process may be near.

Hydrolysis of cellulose to glucose for fermentation works quite well and needs scale up research mainly on the pretreatment step. A decision on which pretreatment is best may be possible within a year. Byproduct credits are crucial to alcohol fuels from biomass, thus several things must fall into place before a factory should be built.

Really outstanding progress has been made in improving cellulase enzyme titers. However, acid hydrolysis has also been improved so that acid is a clear choice for early commercialization. For the longer range, enzymatic hydrolysis seems quite probable.

Taking the long range to be 5 years, the most exciting development is direct fermentation of cellulose. Already, ethanol concentration is approaching 2 per cent in experiments at MIT. This could render separate hydrolysis obsolete. Other projects are dealing with reactor design, new methods for recovery of products, and processes for other petrochemical substitutes. MIT also has leads on using lignin as a fermentation substrate.

There are many ways of spending more money on fermentation research and development, but the most cost-effective projects must be given highest priority. The program has an excellent spectrum of projects on hydrolysis and pretreatment; only a slight expansion of this work can be justified. However, there is not enough study of innovation and cost reduction of obtaining cheap fermentable sugars and converting them to alcohol fuels. There should be a fall-back position if the Purdue idea for saving energy in alcohol recovery does not work and analogous energy efficient schemes are needed for recovery of other products.

A very impressive program of R and D on alcohol fuels by fermentation should unfold at this meeting. Suggestions would be highly welcome. The rapid rate of progress makes it likely that production of alcohol fuels and petrochemical substitutes from biomass can soon become a very large scale operation in the United States.

Component	Approximate Percentage	Use
Cellulose	35	Hydrolyze to glucose
Hemicellulose	30	Hydrolyze to mixed sugars
Lignin	25	Burning
Protein, organics	5	None but may help fermentation
Ash	5	Recycle to soil

Fig. 1. Composition and Uses of Biomass

1. Weak H_2SO_4 at 180°C
2. Alcoholic removal of lignin
3. Grinding to a fine powder
4. Solvents for cellulose
5. Explosive decompression

Fig. 2. Pretreatment Options

DIRECT MICROBIOLOGICAL CONVERSION OF CELLULOSIC BIOMASS TO ETHANOL

aniel I. C. Wang, Ivka Biocic, Hung-Yuan Fang and Sy-Dar Wang, Department of Nutrition and Food Science,
.I.T., Cambridge, MA 02139.

BSTRACT

ne of the major objectives of this project is to chieve the direct microbiological conversion of ellulosic biomass to a liquid fuel, ethanol. ithin the scope of this objective, it is also ne intent to maximize the conversion efficiency f ethanol production from biomass. This can be chieved through the effective utilization of both ne cellulosic (6-carbon sugar) and hemicellulosic 5-carbon sugar) in biomass. The degradation of ellulosic biomass is achieved through the use of thermophilic and anaerobic bacterium, *Clostrid- um thermocellum*. This microorganism is quite nique in that it is able to hydrolyse both the ellulosic and hemicellulosic fractions of biomass ut, unfortunately, it is not able to metabolize ne pentoses. Therefore, to achieve total utili- ation of biomass, a second thermophilic and naerobic microorganism, *Clostridium thermosac- harolyticum*, has been under study due to its bility to convert pentoses to ethanol. Mutation, election and adaptation programs have yielded thanol tolerant strains of both organisms. astly, mixed culture fermentations using these wo organisms show environmental and biological ompatibilities to convert cellulosic biomass to thanol.

TRODUCTION

llulosic biomass represents a renewable resource tentially convertable to liquid fuel such as hanol. Different techniques are being consid- ed to achieve this conversion. These include id or enzymatic hydrolysis of the biomass fol- wed by fermentation using soluble sugars to pro- ce ethanol. This approach generally uses the ast, *Saccharomyces cerevisiae*, due to its ferm- tative ability for the production of ethanol. fortunately, this yeast is unable to metabolize ve carbon sugars such as xylose. It has been tablished that the major production of ethanol rough biological conversion processes in the st for the feedstock [1]. The inability to ilize the hemicellulosic fraction in biomass de- acts to a certain extent such a technique for omass conversion.

searching for novel bioconversion processes for hanol production, the M.I.T. group has been ex- ing a different concept from those stated above. It was rationalized that a single step conversion of cellulosic biomass to ethanol could offer potential economical advantages over those using multiple processes. Within this single step conversion scheme, it was also the intent to utilize both the cellulosic (6-carbon sugars) and hemicellulosic (5-carbon sugars) fractions in biomass to produce ethanol. To achieve these objectives, we have focused our attention on an anaerobic and thermophilic (Temp = 60°C) bacterium, *Clostridium thermocellum*, which is able to hydrolyse cellulose and hemicellulose in biomass. This microorganism is also able to catabolize the six carbon sugars to produce ethanol and other organic acids. However, it is not able to metabolize five carbon sugars such as xylose. Therefore, a second anaerobic and thermophilic (Temp = 60°C) bacterium, *Clostridium thermosaccharolyticum*, is being examined for the conversion of the hemicellulosic fraction to ethanol. The overall goal is to use a mixed culture fermentation for the direct conversion of biomass to ethanol.

MATERIALS AND METHODS

The microorganisms used in the studies were *Clostridium thermocellum* ATCC 27405 (American Type Culture Collection, Rockville, MD) and *Clostridium thermosaccharolyticum* isolated and identified in our laboratory. Medium composition, equipment and the associated analytical procedures have been presented previously [2,3] and will not be repeated. We would like to, however, present our adaptation and isolation procedure for the selection of ethanol tolerant strains of *C. thermocellum* and *C. thermosaccharolyticum*. The organisms were grown using a soluble carbon source such as cellobiose (for *C. thermocellum*) and xylose (for *C. thermosaccharolyticum*) containing initially 10 g/l of ethanol. Serial transfers of the culture were then performed and the ethanol concentration was increased step-wise at 2 g/l each time. At each concentration of ethanol, the cellulolytic activity of *C. thermocellum* was also tested by examining its ability to grow on cellulose. The total number of transfers with progressively higher ethanol concentrations varied between 40 to 65. After these numbers of transfers, an isolate which possesses cellulolytic activity as well as being able to grow in the presence of high ethanol concentration was selected for subsequent studies.

RESULTS AND DISCUSSION

The growth of the parent strain Clostridium thermocellum ATCC 27405 on Solka floc is shown in Figure 1. Since high concentration of cellulose cannot be easily agitated in a fermentor, Solka floc had to be fed in a batch-wise fashion. This can be seen in Fig. 1 as the "saw-tooth" noted as residual cellulose. Cell growth accompanies the utilization of cellulose up to an optical density of 500 Klett unit (330 Klett = 1 g/l cell concentration). A decline in the cell concentration was noted and this is due to the adsorption of cells onto the cellulose. Reducing sugars are produced as a result of cellulose degradation up to about 8 g/l. Previous quantitative and qualitative analyses through high performance liquid chromatography have shown the reducing sugars are mainly glucose, cellobiose and xylose [2]. Catabolic products consisting of ethanol and acetic acid were accumulated at equal concentrations of 4 g/l. These results show that C. thermocellum is able to degrade cellulose and produce soluble sugars, ethanol and acetic acid. Analysis of the biochemical pathway in the catabolism of cellulose by C. thermocellum indicates the natural products are equimolar formation of ethanol and acetic acid [3]. Therefore, if a liquid fuel, ethanol, is the primary product of interest, the accumulation of acetic acid must be eliminated. Furthermore, the concentration of the end products must also be increased.

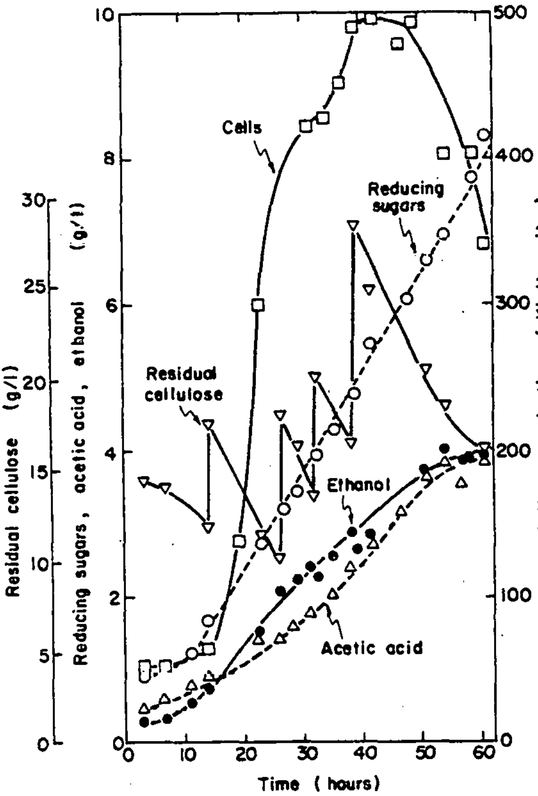


Fig. 1. Growth and Product Formation by C. thermocellum, ATCC 27405 (Parent) Using Batch Feeding of Cellulose (Solka floc)

In order to increase the ethanol concentration by C. thermocellum, its tolerance to this product was examined. Adaptation and selection programs were also performed to increase C. thermocellum's growth tolerance to ethanol. The results from these studies are shown in Fig. 2. The growth of parental strain of C. thermocellum is highly inhibited by low concentrations of ethanol. For example, a 1% (V/V) of ethanol, growth inhibition to 40% of maximum resulted. From our selection procedure, three isolates, designated as S-4, S-6 and S-B, were obtained. All of these isolates can be seen from Fig. 2 and have higher tolerance to ethanol than the parent (ATCC 27405). The properties of these isolates will now be presented.

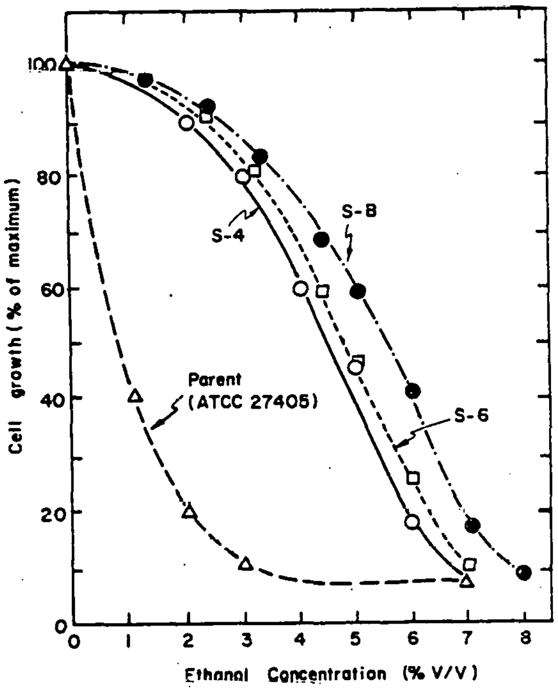
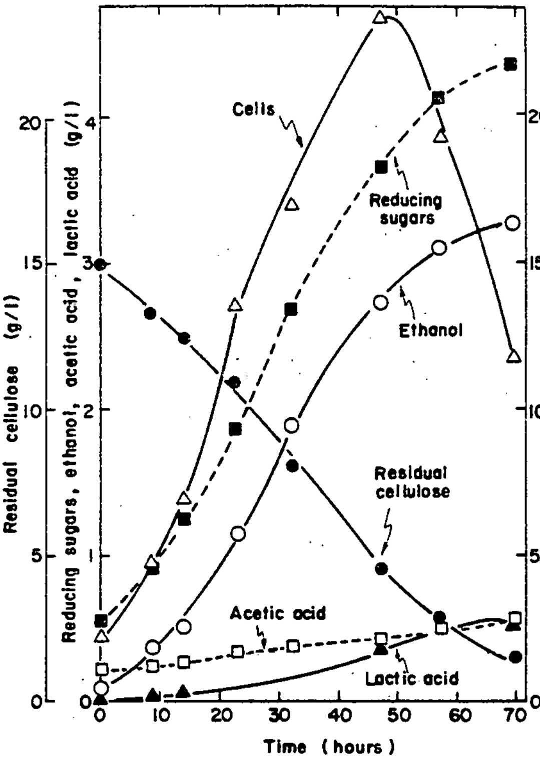


Fig. 2. Comparison of Ethanol Tolerance for Different Strains of C. thermocellum (ATCC-27405 → S-4 → S-6 → S-B)

The fermentation behaviors of the new ethanol tolerant strain, S-4, are shown in Fig. 3. Here again, the cellulose used was Solka floc. Fro initial cellulose concentration of 15 g/l, nea



ig. 3. Fermentation Profile of C. thermocellum (S-4) on Solka floc

otal degradation occurred. Reducing sugars were ccumulated to a final concentration of 4.5 g/l. he most interesting and amazing behavior of this ew strain is the high ratio of ethanol formation n comparison to acetic acid. From Fig. 3, it can e seen that about 3.2 g/l of ethanol was formed ut less than 0.5 g/l of acetic acid accumulated. t is hypothesized that this new strain, S-4, not nly is it able to tolerate higher concentrations f ethanol, but its metabolism has also been al- ered. This is possible if one considers the cata- olic pathway of this organism with respect to thanol and acetate formation. For example, the onventional catabolic reactions leading to eth- nol and acetic acid can be represented as [4]:

$$_6H_2O_6 + H_2O + 3ADP + 3P_i \rightarrow CH_3CH_2OH + CH_3COOH + 2CO_2 + 2H_2 + 3ATP \qquad (1)$$

n this case, one would expect equimolar ratios of thanol and acetic acid as well as H_2 and CO_2. On he other hand, when carbon and oxidation-reduction alances were performed using the results in Fig. , the reaction shown in Equation (1) was shown to e no longer valid. It is postulated that the new train, S-4, produces very little hydrogen gas and early all of the reducing power generated was sed to produce NADH and thus increasing the ratio [illegible]thanol to acetic acid formation.

Further adaptation and selection of Strain S-4 to increase its ethanol tolerance resulted in the isolation of strain S-6. The fermentation characteristics of this strain are shown in Fig. 4. In

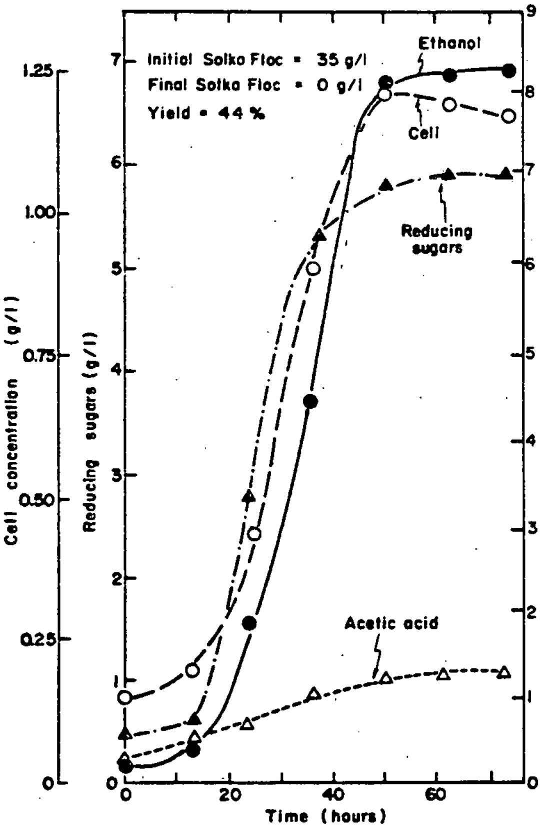


Fig. 4. Fermentation Profile of C. thermocellum (S- 6) on Solka floc

this experiment, 35 g/l of cellulose (Solka floc) was used. At the end of the fermentation, no residual cellulose was found thus showing 100% cellulose degradation. From an initial 35 g/l of cellulose, the fermentation products were: cells = 1.2 g/l, ethanol = 8 g/l, acetic acid = 1.0 g/l and reducing sugars = 6 g/l. These results again show that this new strain (S-6) still maintains the favorably high ethanol to acetic acid production ratio, as well as having a high efficiency of cellulose degradation. The conversion yield of products from cellulose is also excellent. For example, the overall yield including all catabolic and degradative products was calculated to be 0.44 grams of product per gram of cellulose. Since the conversion efficiency from feedstock is of primary importance, these results show an excellent potential in the use of C. thermocellum for biomass conversion.

The ability of Clostridium thermocellum to degrade natural biomass has also been examined.

Dried corn stover (kindly furnished by A.E. Staley Co., Decatur, IL) was first hammer milled (U.S. Army Natick Labs, Natick, MA) to an average diameter of about 2mm. This material was then tested using the new isolate, strain S-6, of C. thermocellum. The initial corn stover concentration was 35 g/l. The results from these studies are shown in Fig. 5. The final corn stover after degradation was about 10 g/l, representing a degradation efficiency of about 70%. The degradation products were 5.2 g/l of reducing sugar, 4 g/l of ethanol and 1.0 g/l of acetic acid. These results show that C. thermocellum is able to effectively degrade a natural biomass, corn stover, and still maintain a favorable ethanol to acetic acid production ratio.

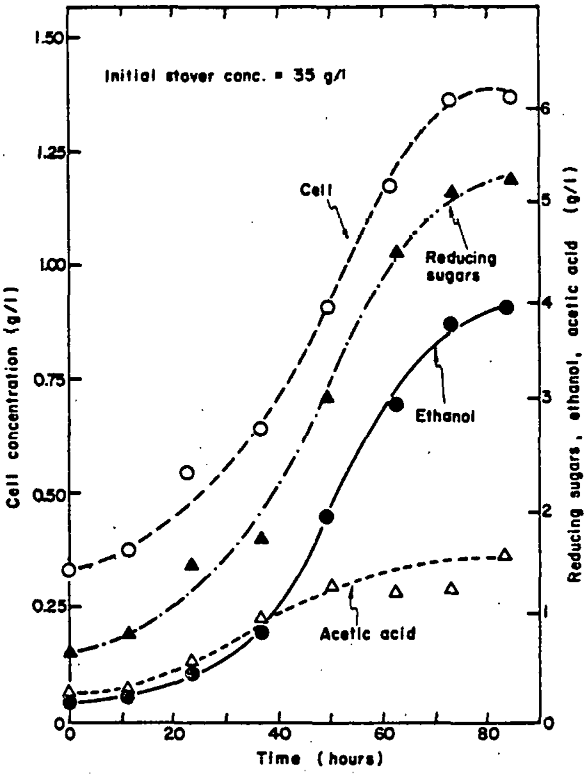


Fig. 5. Fermentation Profile of C. thermocellum (S-6) on Corn Stover

The ability to convert the cellulosic (6-carbon sugars) fraction of biomass to useful products represent only one-half of the overall problem. This is because most natural biomass such as corn stover contains about an equal amount of hemicellulose (5-carbon sugars) and cellulose. If the hemicellulose cannot be effectively utilized, the overall conversion efficiency would be quite unfavorable with respect to process economics. Therefore, an equally important segment of our project has been on the conversion of xylose to ethanol. We have shown in our laboratory that the cellulase enzyme complex from Clostridium thermocellum is equally effective in the hydrolysis of hemicellulose in biomass to xylose as compared to cellulose hydrolysis. Unfortunately, C. thermocellum cannot metabolize pentoses. Therefore, to effectively utilize the five carbon sugars, we have been examining the capability of another anaerobic thermophile. This organism was isolated and identified in our laboratory to be Clostridium thermosaccharolyticum. The uniquenesses of this organism are its xylose catabolism as well as being environmentally and biologically compatible as a mixed culture with C. thermocellum. The behaviors of C. thermosaccharolyticum in pure and mixed culture will be presented.

Our immediate efforts with C. thermosaccharolyticum were to characterize its growth tolerance as well as selection to increase its tolerance towards ethanol. Using the adaptation technique, a higher tolerant strain of this organism was selected and its behavior compared with the parent in shown in Fig. 6. The parental strain (HG-2) is more resistant to ethanol than Clostridium thermocellum. For example, for the parent (HG-2), 50% growth inhibition occurred at an ethanol concentration of about 3% (V/V). The new strain, HG-3, was found to be more resistant to ethanol as can be seen in Fig. 6. For example 50% growth inhibition for strain HG-3 has now been increased to an ethanol concentration 4% (V/V). These results are encouraging since they demonstrate the selection procedure established

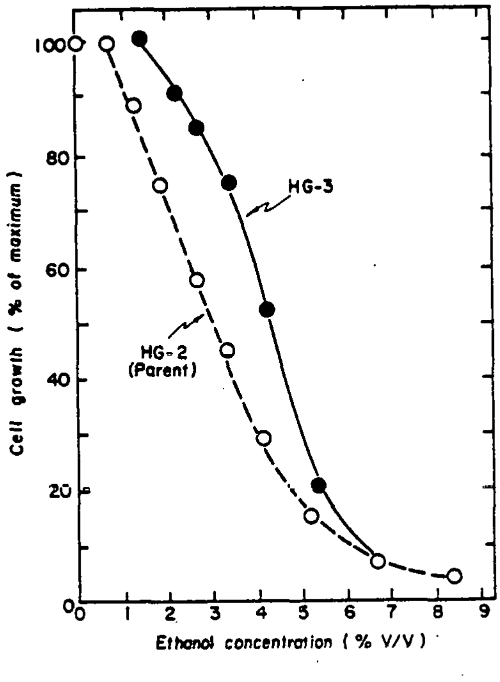
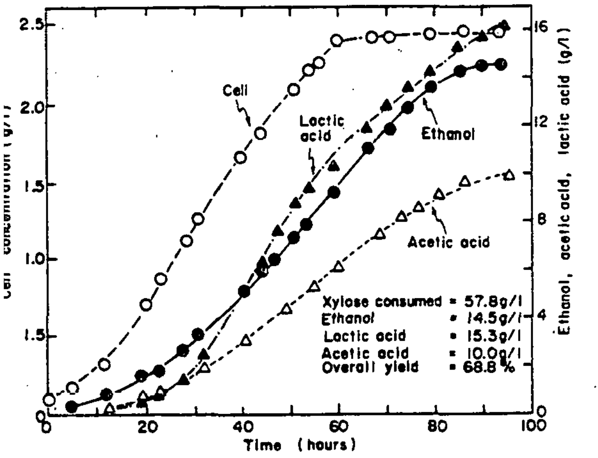


Fig. 6. Comparison of Ethanol Tolerance for Different Strains of C. thermosaccharolyticum (Parent: HG-2 → HG-3)

thermocellum can be readily adapted to microorganisms. We are confident that fur- ier increased tolerance to ethanol is still :tainable through this adaptation and serial :ansfer technique.

ie behavior of Clostridium thermosaccharolyticum, :rain HG-3, as a pure culture using xylose as the ırbon source is shown in Fig. 7. Cell growth is



.g. 7. Fermentation Profile of C. thermosaccharolyticum (HG-3) on Xylose

ıite excellent attaining a maximum value of 2.4 '1. Xylose was fed in a batch-wise fashion since ır earlier studies have shown high concentrations : xylose is not desireable. In this fed-batch ırmentation, xylose was maintained between 5 to) g/l. The catabolic products were ethanol ı.5 g/l), lactic acid (16 g/l) and acetic acid .0 g/l). It should be mentioned that the ratio : ethanol to acetic acid of the new ethanol tol- ·ant strain (HG-3) is increased similarly to that ıserved for C. thermocellum. For example, the ırental strain (HG-2) produced equal concentra- .ons of ethanol and acetate whereas the new :rain (HG-3) has an ethanol to acetate ratio of 2. A total of 57.8 g of xylose was consumed in ıe fermentation as shown in Fig. 7. This repre- nts a total product (ethanol, acetic, lactic) nversion efficiency of 68.8%! This is quite en- uraging since it shows that C. thermosaccharolyticum is already quite efficient with respect to s catabolism from xylose.

has been assumed in our overall program scope at a mixed culture of C. thermocellum and C. ermosaccharolyticum can be compatibly established that will lead to simultaneous cellulose gradation and product formation. To test the lidity of this assumption, a mixed culture fermtation was performed using Solka floc (30 g/l) the carbon source. The results from this exriment are shown in Fig. 8. Cell growth was [illegible] rapid, although we were not able to differen-[illegible] the relative proportion of the two microbial populations. The catabolic products were ethanol (9 g/l), acetic (3.5 g/l) and lactic (2 g/l) acid. No detectable reducing sugars were formed during the fermentation. These results are extremely encouraging since they attest to our hypothesis that these two microorganisms can grow compatibly in mixed culture to degrade cellulose and produce useful products.

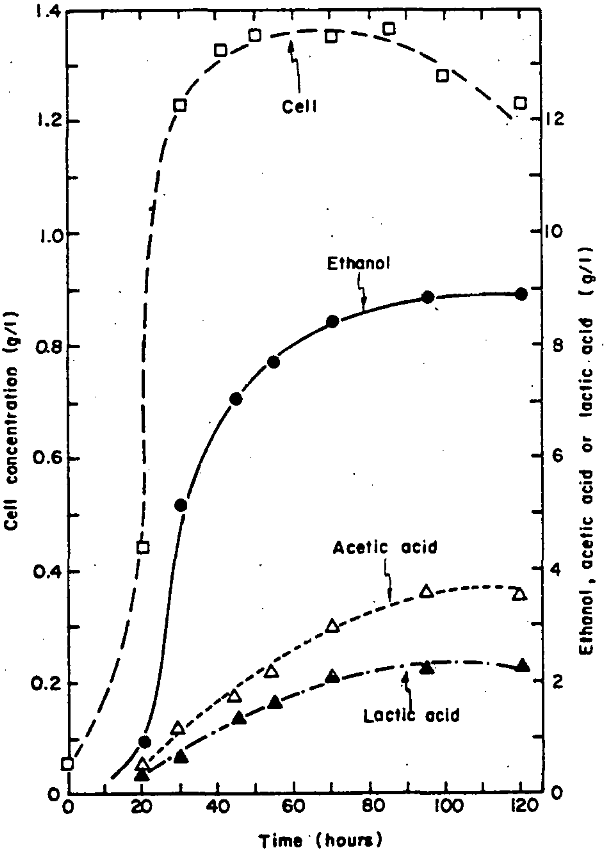


Fig. 8. Mixed-culture of C. thermocellum (S-4) and C. thermosaccharolyticum (HG-3) on Solka floc

As a further demonstration of using the concept of mixed cultures, experiments were also performed using the ethanol tolerant strain S-B which was isolated and reported earlier (see Fig. 2) in this paper. Strain S-B was isolated as a derivative of Strain S-6 of C. thermocellum during our adaptation program to increase ethanol tolerance. There were a number of interesting properties with Strain S-B in addition to its increased tolerance to ethanol. These include its ability to grow on cellulose, glucose, cellobiose and xylose. In particular, strain S-B was transferred eight times sequentially using xylose as the carbon source and subsequently transferred to grow on Solka floc and still exhibited its cellulolytic capabilities. Careful biochemical and morphological examinations have convinced us that Strain S-B is a stable mixed culture of C. thermocellum and C. thermosaccharolyticum. A mixed culture of these two organisms is easily conceivable to have been obtained since both are being studied in the same laboratory. It is not fortuitious, however, to have hypothesized that such a stable mixed culture can be so easily derived.

The fermentation profile of this mixed isolate S-B when grown on an initial concentration of 35 g/l Solka floc is shown in Fig. 9. As one would

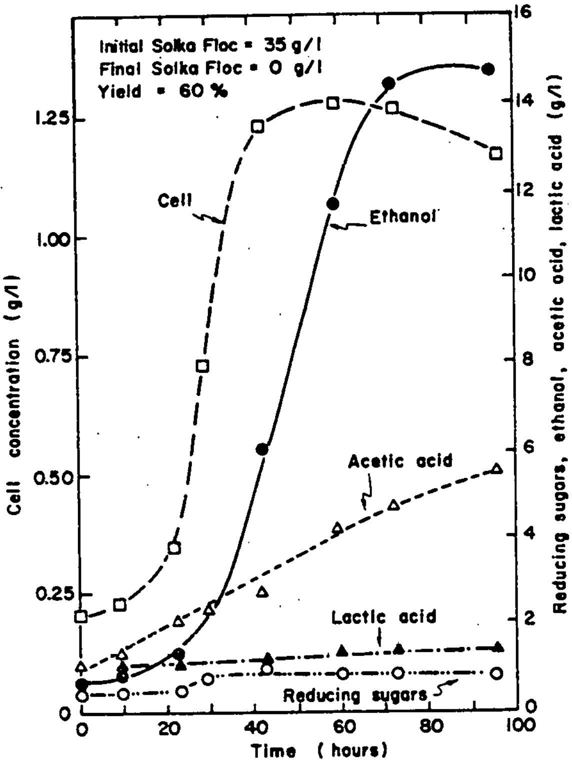


Fig. 9. Mixed-Culture (S-B) of C. thermocellum and C. thermosaccharolyticum on Solka floc

expect, no reducing sugars accumulated since C. thermosaccharolyticum can readily metabolize these sugars from cellulose. From an initial 35 g/l of cellulose, 100% degradation resulted. The catabolic products were ethanol (15 g/l) and acetic acid (5 g/l). This high concentration of ethanol (15 g/l = 1.9% (V/V)) began to show promise of our approach for liquid fuel production from cellulose. We believe the cessation of product formation was a result of cellulose depletion. Unfortunately, to obtain high initial concentration of cellulose in fermentors is extremely difficult due to its rheological properties. The conversion efficiency from this fermentation was calculated to be 0.57 gram of product per gram of cellulose. These results continue to give us encouragement in the realization of our overall concept.

The last experiment to be presented in this pape is the behavior of this stable mixed culture, S-B, when grown on corn stover. The results are shown in Fig. 10. From an initial corn stover concentration of 35 g/l, 71% degradation resulted. Fror the data in Fig. 10, it can be seen that only small amount (less than 1 g/l) of reducing sugars accumulated. Here again, the main catabolic products were ethanol (4 g/l) and acetic acid (2 g/l).

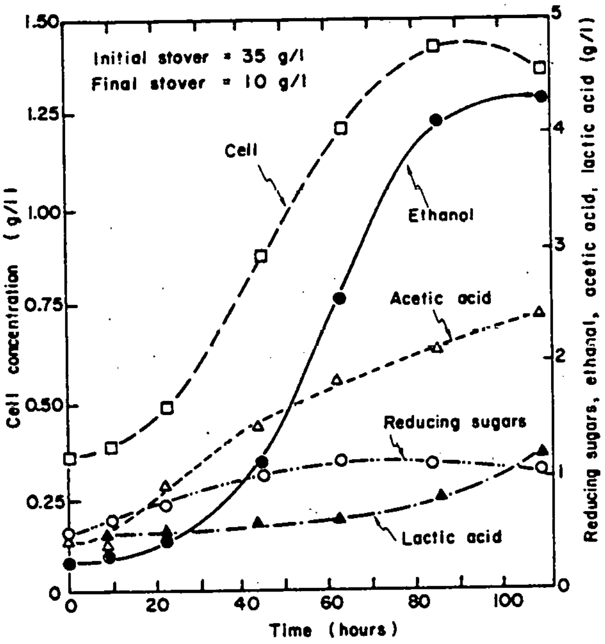


Fig. 10. Mixed-Culture (S-B) of C. thermocellum and C. thermosaccharolyticum on Corn Stover

CONCLUSIONS

The results from these studies show that the concept for the direct microbial conversion from cellulosic biomass to ethanol is sound. The use of an anaerobic and thermophilic bacterium, Clostridium thermocellum, to degrade cellulose and to produce ethanol can be achieved. Mutation and selection program to increase ethanol tolerance has been possible. An isolate which has higher tolerance towards ethanol as well as being able to produce ethanol in favor of acetic acid has been obtained. To effectively utilize the degradation products from hemicellulose, a second anaerobic and thermophilic bacterium, Clostridium thermosaccharolyticum, has been isolated. This organis can grow compatibly with C. thermocellum which ca simultaneously degrade cellulose and produce useful products.

WLEDGEMENT

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ram under Contract Number EG-77-S-02-4198.

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NOTES

PRODUCTION OF LIQUID FUELS FROM CELLULOSIC BIOMASS

U.S. DOE Contract # EY-76-S-02-4070
(Original Starting Date September 15, 1976)

E. Kendall Pye
Department of Biochemistry & Biophysics
University of Pennsylvania
Philadelphia, PA 19104.

Arthur E. Humphrey
School of Engineering & Applied Science
University of Pennsylvania
Philadelphia, PA 19104.

STRACT

novel, elevated temperature process for the total nversion of cellulosic biomass to liquid fuels s been devised and shown to be technically feas- le by our research group. It has several unique atures which allow reduced process energy costs the simultaneous production of two liquid fuels- ethanol for use as a gasoline extender and octane hancer, and a butanol/lignin slurry for use as a mpable furnace or diesel fuel. The process relies a hot aqueous butanol pretreatment of the bio- ss to yield an enzyme-degradable cellulose frac- on, a high quality polymer-grade lignin fraction, partially degraded hemicellulose fraction for rmentation to butanol, and a butanol/lignin urry for use as a fuel. The cellulose is fer- nted to ethanol either by an elevated temperature 60°C) simultaneous saccharification/vacuum rmentation step using cellulase from <u>Thermoactin- yces</u>, and the anaerobic bacterium <u>C. thermocellum</u>; by hydrolyzing it to a high glucose syrup (>20%) th <u>Thermoactinomyces</u> cellulase for use in a andard yeast fermentation. The process is cur- ntly being tested and optimized with wood chips om fast growing poplar trees, but it also appears tractive for conversion of municipal solid waste d agricultural residues. Production costs for hanol appear to be about 70-80¢/gallon, depending biomass costs, yields and byproduct credits.

SCRIPTION OF TASK

e experiences of Spring 1979 with supply restric- ons and rapidly rising prices for gasoline, esel fuel and heating oil have served to emphas- e the dangers to the U.S. economy of a continued pendence on foreign sources for a major fraction its oil supply. Further political instability the major oil exporting countries could lead at y time to additional reductions in the availa- lity of oil on the world market, resulting almost rtainly in dramatic price increases and possibly vastating effects on the U.S., Western European d Japanese economies. To defend itself against is continuing threat the U.S. must rapidly reduce s dependence on foreign oil by substituting mestic resources to power its current transporta- on, heating and much of its electricity generat- g systems. Of the resources available, increased ic oil production is now possible because the current prices for oil make some forms of oduction, such as low producing wells and tertiary recovery, economically feasible. But it must be recognized that this is at best only a temporary situation which ultimately will decline in the not too distant future together with our overall reserves of crude oil. In order to allow continued use of our liquid fuel-dependent systems we need to generate compatible substitutes from our more abundant fossil energy reserves, such as coal, or renewable resources, such as biomass. Biomass as an energy source has many appealing features including the fact that it is renewable over the short term, it is capable of being stored, it is environmentally relatively inoffensive, it is free of horrendous hazard potential, it is available in one form or another throughout the U.S., it is tolerably priced, it is produced by currently available technology, it can be developed rapidly with the manpower and material resources already available, and it is high in thermodynamic availability [1]. In addition, the U.S., because of its large land mass and great agricultural capacity has the potential to make biomass a major contributor to the nations energy resources, providing at least 10% [2], and possibly much greater, of our total energy needs.

The problem, however, is how to efficiently convert any or all of the available forms of biomass into oil-sparing liquid fuels or chemical feedstocks. Technology already exists for converting many agricultural products, such as whole grain, sugar, molasses and starch into organic solvents and chemical feedstocks. Also, ethanol for use as a motor fuel in the form of Gasohol, is already being produced from these materials at a rapidly expanding rate, significantly stimulated by recent government financial incentives, the lessened economic competition from higher-priced gasoline and gasoline supply restrictions. However, by far the largest biomass resources are cellulosic in nature, as are many readily available waste materials such as forestry residues, municipal solid waste, cattle feed-lot residues and agricultural residues such as corn stover and bagasse. In addition, a major new industry could develop around silvaculture plantations for rapid production of woody biomass.

The advantages of these woody biomass resources are their enormous present and potential volumes, relatively cheap cost, non-competition with food uses, relatively cheap cultural requirements, and, in many cases, year-round availability. The problem lies in the lack of an available technology for their efficient conversion to the most desirable liquid fuels and chemical feedstocks at a price which is competitive with todays hydrocarbon fuels.

OBJECTIVES/COST AND PERFORMANCE TARGETS

In light of the situation described above, the objectives of this particular project are to design a process for the efficient conversion of woody biomass to oil-sparing chemical feed-stocks and liquid fuels. The process should have the following features.

- It should produce the most-needed products including substitutes or extenders of gasoline, diesel fuel and heating oil, as well as chemical feed-stocks.
- It should provide for total biomass utilization in order to be cost efficient since biomass cost is the major component in the cost of the final products.
- It should produce no noxious wastes or by-products, in order to be environmentally compatible and ecologically acceptable.
- It should use as little process energy as possible, as one way of reducing costs.
- It should be capable of using a variety of the most available woody biomass materials.
- It should be relatively simple and as low in capital and operating costs as possible.

Such a process should be capable of generating products competitive in price with those derived from oil within 3 - 5 years, the time required to develop and test such a process. In this regard, it is difficult to identify the exact cost targets because of the recent rapid rise in crude oil prices and projections for such price rises in the future. However, because of the immediate use of ethanol in Gasohol, where the value of the ethanol, because of its octane enhancing properties, is much greater than its BTU content alone, it appears that ethanol produced for approximately $1/gallon would be highly competitive at this time. The process to be described appears to be capable of meeting this cost target.

Performance targets being set are those currently achieved in well developed fermentation processes using starch or sugar as feed-stocks, which approach process efficiencies of 90 - 95%.

APPROACH

The approach we have taken is to design a process which converts woody biomass mainly into two liquid fuels -- 1) ethanol for use as a gasoline extender and octane enhancer, and 2) a butanol/lignin slur for use as a pumpable furnace fuel, or even as a diesel or turbine fuel. In addition the process can generate high value by-products such as polymer grade lignin, acetone, cattle feed and high purity xylose. However, the process economics in no way rely on the value of these by-products, although the early introduction of the process could be enhanced by their value. The major reason for designing the process for total production of oil-sparing fuels and feedstocks is to eliminate the economic problems associated with the value and sale of by-products once large number of such facilities come on stream.

The Penn/G.E. Process for Total Biomass Utilization

The process, shown in Fig. 1. appears to meet most, if not all, of the criteria set out above, through the use of several novel or unique process steps. As will be described later, the front end of the process is also capable of being used to generate concentrated syrups (>20% glucose) for immediate application in conventional fermentation systems using yeasts.

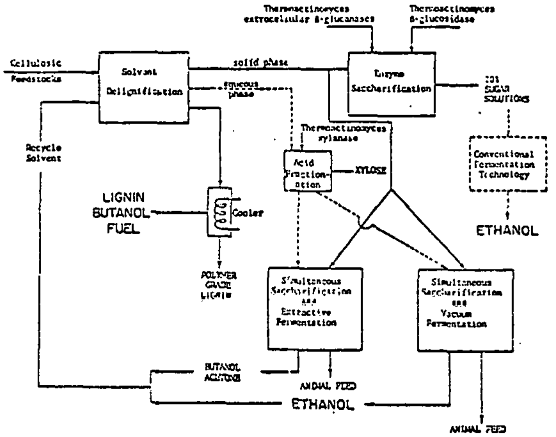


Fig. 1: The Penn/G.E. Process for Total Biomass Utilization.

The unique features of the process include the use of hot aqueous butanol pretreatment to delignify the biomass and to yield 1) a solid phase of cellulose, which is highly degradable by cellulase enzymes, 2) an aqueous phase of partially degraded hemicellulose from which can be recovered a relatively pure xylan for xylose production, the remaining material being fermented by C. acetobutylicum to yield butanol and some acetone, and 3) a butanol phase from which can be recovered a high-quality polymer grade lignin for use in thermoplastics and other applications. The remaining lignin precipitates from the butanol as it cools to yield a butanol/lignin slurry which can be u[illegible] as a pumpable furnace fuel, diesel fuel or turb[illegible] fuel. Since lignin has a high BTU content this will have greater heating value than butanol

nother unique feature of this process is its use f elevated temperature fermentations. The thermohylic cellulolytic organism *Thermoactinomyces* roduces a cellulase having extracellular endond exo-glucanases of excellent thermal stability 3]. In addition the organism, when cultured on ellulose at 60° - 65°C, will yield its maximum ellulase activity in less than 24 hrs of growth. hus its productivity is very high, even though it oes not yet produce the high cellulase activities btained with *T. reesei* [4]. The β-glucosidase of his organism is found intracellularly, but it is ot particularly stable at higher temperatures.

he high thermal stability of the *Thermoactinomyces* xo- and endo-glucanases allow these enzymes to be sed in conjunction with the anaerobic bacterium *. thermocellum* in a single, combined, high temperture saccharification/fermentation step from hich ethanol is removed continuously by a mild acuum (∿ 170 mm mercury). The advantages to this re numerous.

- *C. thermocellum* metabolizes the disaccharide cellobiose, the principal product of the action of the *Thermoactinomyces* extracellular cellulases on cellulose. The continuous removal of cellobiose by the *C. thermocellum* therefore does not allow cellobiose, a potent inhibitor of cellulases, to build up to a sufficient concentration to significantly reduce the rate of further cellulose breakdown.
- At 60°C, with a concentration of ∿3% ethanol in the fermentor and a pressure of 170 mm, the heat removed from the fermentor approximately equals the heat generated during the fermentation. Consequently the fermentor operates in an adiabatic mode. Also the composition of the condensed vapor phase is approximately 27 - 30% ethanol. Therefore the need for a beer still is eliminated. This, plus the elimination of the need for cooling the fermentor, significantly reduces energy use and costs in this step.
- The vacuum would not be pulled on the fermentor as a whole, but rather on a side arm. This allows CO_2 and H_2 to be removed at atmospheric pressure and allows, through a 50% reduction in vapor handling, a major reduction in the size, cost and capacity of the compressors. The CO_2/H_2 gas stream is sufficiently rich in H_2 to allow it to be burned and used as a source of process heat in the facility.
- The use of elevated temperatures and thermophylic organisms greatly reduces the possibilities for contamination of the culture.

utanol/acetone production is achieved in a 30°C ntation employing simultaneous saccharifica- and fermentation of the hemicellulose fraction. Again enzymes from *Thermoactinomyces* are employed but the fermentative organism is *C. acetobutylicum*. Because the costs of recovering butanol by simple distillation are so high, the concept of extractive fermentation is employed. In this concept a liquid extractant, which is immiscible with water, is contacted with the fermentation broth, either directly or separated by a membrane. The extractant absorbs the butanol (or ethanol) and is transferred to a flash heater where the butanol is flashed from the extractant, which, upon cooling, returns to the broth. A suitable extractant must have the following properties.

- It must be non-toxic to the fermentation organism.
- It must be cheap.
- It must have a very high boiling point and low vapor pressure.
- It should have a high distribution coefficient for the compound to be extracted.
- It should be selective for the required compound.

KEY RESULTS/ACCOMPLISHMENTS

The process described above has been tested for technical and economical feasibility. Each of the key process steps is now undergoing optimization studies to improve yields, reduce energy input and reduce costs.

Feed-stock

Currently the process is undergoing development using wood chips from fast growing poplar trees (hybrid poplar, clone 388). This biomass source is of interest for several reasons.

- It can grow in practically every state.
- It can grow well even on marginal land.
- It can be harvested every 4 years, chipped on site and transported in chipped form.
- It does not require replanting.
- It is projected that it will yield up to 8 - 10 ODT/acre/year.
- Its composition is in the range of glucan 50 - 57%, xylan 16 - 25%, lignin 16 - 26%, other organics 6-10% and ash 0.2 - 0.4%.
- It can be harvested year round.

Pretreatment

Preliminary trials of the hot aqueous butanol pretreatment have indicated that the best digestion conditions are in the range of 150 - 175°C for 15-30 min. Lower temperature conditions remove less than optimal amounts of lignin. The solid cellulose fraction obtained from this pretreatment step is highly degradable by the *Thermoactinomyces* cellulases, as shown in Fig. 2.

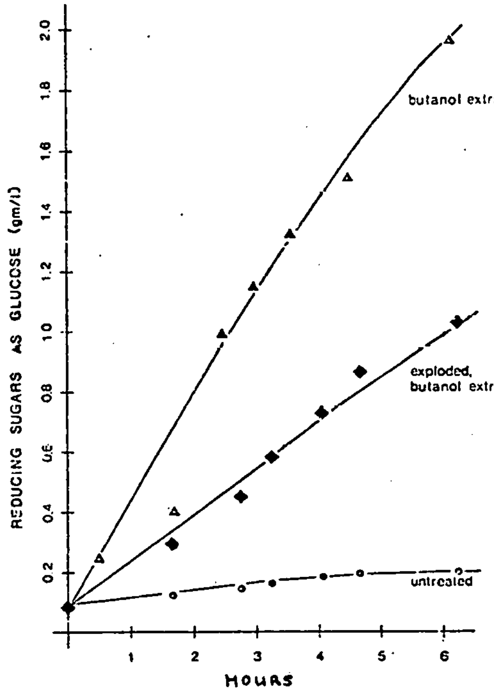


Fig. 2. Hydrolysis of Poplar Chips by Extracellular cellulase of *Thermoactinomyces* YX. 25 dry grams chips/liter. Chips wet-milled in blendor before hydrolysis.

In order to thoroughly test the numerous variables for optimization of the butanol pretreatment process, a highly versatile apparatus has been built which will allow collection of all gaseous and liquid phases. This apparatus, shown in Fig. 3, is currently undergoing trials.

Production of High Concentration Glucose Syrups (> 20%).

The major product of cellulose hydrolysis by *Thermoactinomyces* extracellular cellulases is the disaccharide cellobiose [5]. This is because the β-glucosidase in this organism is entirely intracellular [3]. As with other cellulase systems cellobiose is a potent inhibitor of further cellulase action. This is illustrated in Fig. 4. It can be seen that at concentrations of cellobiose of 5% the *Thermoactinomyces* extracellular cellulase is more than 70% inhibited. However, glucose is far less inhibitory. Therefore conversion to glucose by the combined cellulase (including intact cells of *Thermoactinomyces*) gives reasonable rates of production of glucose from cellulose. This is shown in Fig. 5. in which a

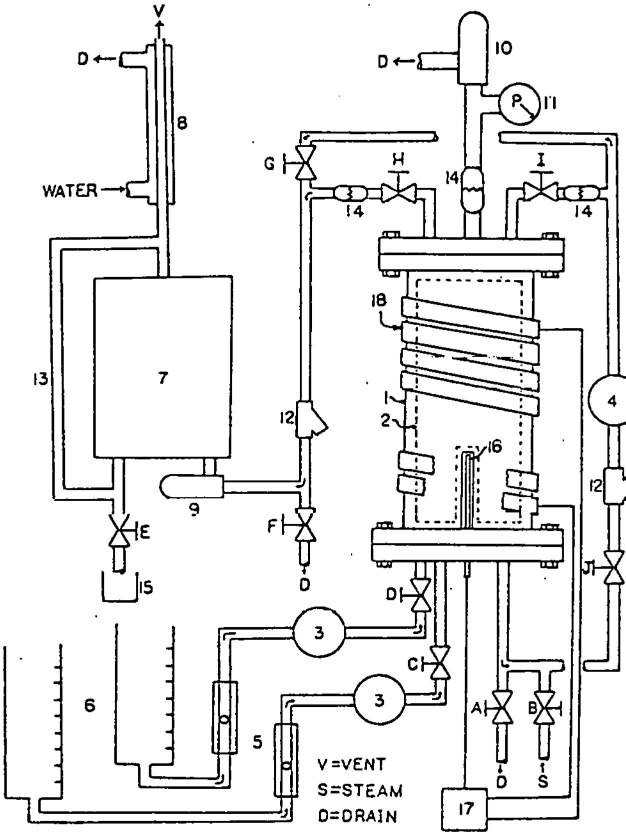
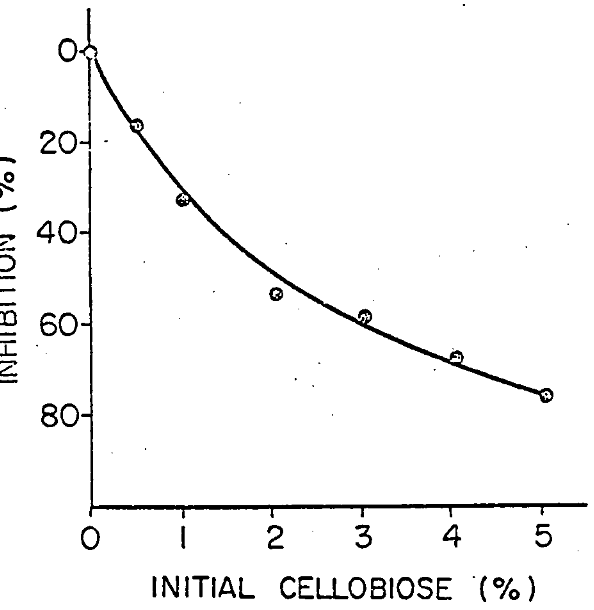


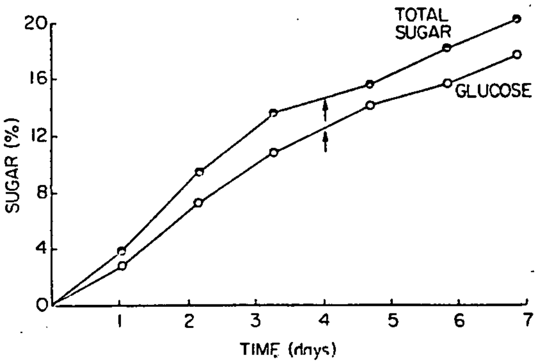
Fig. 3. Experimental Delignification Apparatus. Key: 1. Reactor; 2. Basket; 3. Meter Pump; 4. Recycle Pump; 5. Rotameter; 6. Volumetric Flask Reservoir; 7. Flash Chamber; 8. Condenser; 9. Back Pressure Regulator; 10. Safety Relief Valve; 11. Pressure Guage; 12. Y-Type Strainer; 13. Sight Glass; 14. Quick Disconnect; 15. Sample Bottle; 16. Platinum Resistanc Temperature Probe; 17. Proportional Temperature Controller; 18. Heating Tape.

syrup containing close to 20% glucose is produced from swollen cellulose by the action of the combined cellulases of *Thermoactinomyces*. The practical purpose of this demonstration is that if concentrated glucose syrups can be produced from cellulose, then they can be employed in conventional fermentations using yeasts, thus making a cheaper sugar source available for fermentations now. This would allow cheap cellulose to become a fermentation resource even before the rest of the technology is ready for commercialization.

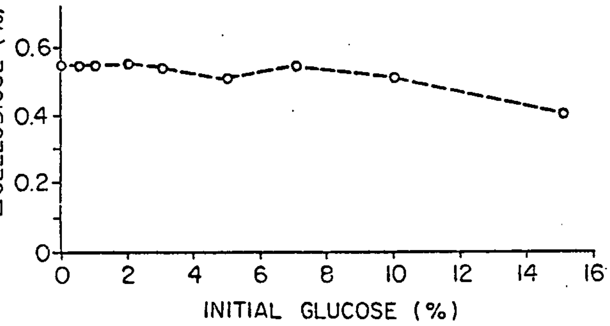
Interestingly enough the *Thermoactinomyces* β-glucosidase shows remarkably little inhibition by the product of its action, glucose. This feature of the enzyme makes it ideally suited for produ[illegible]tion of high glucose syrups. Fig. 6, which sho



g. 4. Cellobiose Inhibition of Thermoactinomyces cellulase acting on 2% acid-swollen cellulose.



g. 5. Glucose Syrup Production from Cellulose. Incubation temperature 50°C; fresh enzyme added and cellulose added every 24 hours.



6. Inhibitory Effect of Glucose on Cellobiose Hydrolysis by Thermoactinomyces Cell-Solids

the breakdown of cellobiose in the presence of various concentrations of glucose, by Thermoactinomyces β-glucosidase, demonstrates how little this enzyme activity is effected by glucose concentration.

Unfortunately, the β-glucosidase is relatively unstable at higher temperatures as shown by Fig. 7. Consequently, the production of concentrated glucose syrups would have to be performed at lower temperatures or with an enzyme possibly stabilized by immobilization. Immobilization of Thermoactinomyces β-glucosidase has been successfully accomplished, using $TiCl_4$ as a coupling agent, to controlled pore glass beads. Although >30% of the activity was retained on immobilization a more stable enzyme did not result.

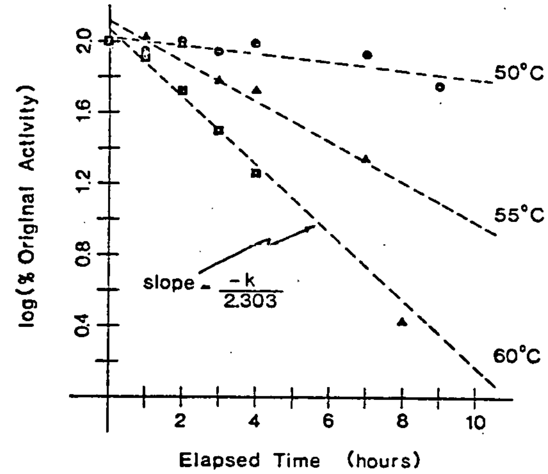


Fig. 7. Crude β-Glucosidase, Temperature Stability. (pNpβG activity).

Ethanol Production by Combined Saccharification/ Fermentation

A completely synthetic medium for the active growth of several interesting strains of C. thermocellum has been devised. The use of this medium and simplifications of it will significantly reduce the cost of the saccharification/fermentation step.

Initially it was found that extracellular enzyme preparations from some strains of Thermoactinomyces were inhibitory to the growth of C. thermocellum. Some strains of Thermoactinomyces were found which did not produce this inhibitor, which was found to be a heat labile protein. Now this problem has been resolved with the finding that small amounts of activated charcoal will selectively remove the inhibitor from the cellulase enzyme preparation from Thermoactinomyces.

Extractive Fermentation

A large number of potentially interesting extractents have been tested for the continuous recovery of butanol from C. acetobutylicum fermentation broths. One of the most interesting has proved to be dibutyl-phthalate (DBP). It costs 40¢/lb, has a boiling point of 340°C, does not inhibit the growth of the organism and has a distribution coefficient of 1.8 for butanol in water. Also, no emulsion forms in the fermentor liquor. Fig. 8 shows the vapor pressure of DBP as a function of temperature. The boiling points of acetone, ethanol and butanol at 1 atmosphere are given for comparison. It can be seen that DBP fulfils the requirements of a suitable extractant.

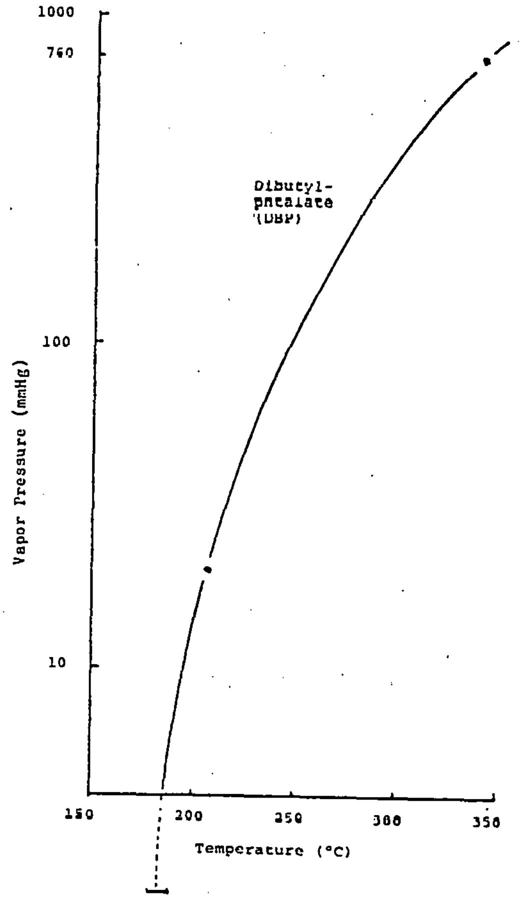


Fig. 8. Vapor Pressure of Solvent and Extractant (DBP). Boiling Point at 1 atm: Acetone 56.2°C; Ethanol 78.5°C; Butanol 117.7°C.

Economic Evaluation

An economic evaluation of the Penn/G.E. process indicates that the manufacturing price of ethanol made by this process is highly dependent on process yield and by-product credits. Fig. 9 shows the dependence of ethanol manufacturing cost on these parameters. It can be seen that with a yield of 95% of theoretical (for both cellulose and hemicellulose) and full by-product credits, manufacturing costs can be as low as 65 - 70¢/gallon 95% ethanol. This is based on a 25 million gallon/year facility with biomass costs of $11.25 ODT. The assessment includes capital related costs, ra materials and chemicals, utilities, labor and maintenance.

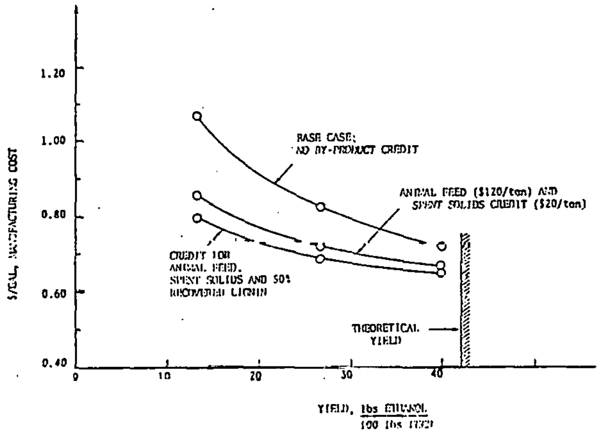


Fig. 9. Manufacturing Cost of Ethanol by Enzyme Hydrolysis. Yearly Production: 25 Million Gallons (95%).

FUTURE PLANS

Every step in the process is being investigated in order to optimize it from a yield, energy and economical standpoint. This process will continue for some time, but within one year we hope to demonstrate the integrated process on a bench scale. Major emphasis will also be placed on optimizing the production of concentrated glucose syrups, on increasing the yield of enzyme from Thermoactinomyces by mutational and cultural modification, and on testing other biomass sources especially waste cellulose, with the process.

ACKNOWLEDGMENTS

We wish to acknowledge the outstanding work of our subcontractors at General Electric Co., Re-entry and Environmental Systems Div., Philadelphia (Dr. J. Forro and E. Nolan) and at Hahnemann Medical School, Philadelphia (Dr. J. Alexander). Full credit should also be given to the many graduate students, research technicians and research specialists whose work has been summarized in this report. We also wish to acknowledge the work and advice of Dr. R. Mateles, Visiting Professor at the University of Pennsylvania.

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NOTES

SELECTIVE SOLVENT EXTRACTION IN UTILIZATION OF STORED SOLAR ENERGY IN CELLULOSIC BIOMASS

George T. Tsao
Laboratory of Renewable Resources Engineering
Purdue University
West Lafayette, Indiana 47907

DOE Contract No. ET-78-S-02-4658
Starting Date: January 1, 1978

SCRIPTION OF TASK

llulosic materials such as cornstalks used in con- rsion processes are usually made of three major mponents: cellulose, hemicellulose and lignin. utilization of such crude mixtures, fractionation the raw materials into pure components usually ll result in upgrading the value. Cellulose, in ct, is difficult to hydrolyze due to two main stacles, one of which is the presence and the terference of lignin that cements cellulose fibers gether. The other factor that makes cellulose sistant to hydrolysis is the highly ordered ystalline structure of cellulose. In order to ercome the two obstacles, Purdue researchers have ployed solvents to dissolve cellulose. Once it dissolved into a liquid, there is no longer ystalline structure nor the interference by gnin. Consequently, the treated cellulose becomes sily hydrolyzable by either acids and enzymes to ucose in high yields at a fast reaction rate.

other important feature of the Purdue Program is s heavy emphasis on the utilization of hemi- llulose hydrolysate. In cornstalks, there is tually more hemicellulose than cellulose. Unless can get a reasonable by-product credit from hemi- llulose, cellulose alone cannot yield alcohol at low cost. We recognized this fact very early in r program, when everyone else was paying tention only to cellulose hydrolysis and fermen- tion of glucose. The Purdue Program actually was arted as a study of fermentation of pentoses from micellulose. We are now investigating the oduction of butanediol, diacetyl, methyl ethyl tone, ethanol, lactic acid and pyruvic acid from micellulose hydrolysate.

JECTIVE /COST AND PERFORMANCE TARGETS

e long term objective of the Purdue Program is to oduce liquid fuel and petroleum, substituting emicals from cellulosic materials, and thus reduce e United States dependence on imported oil. To hieve such a long term objective, we need to have wide spectrum of research activities. Purdue's fort is to use solvents to help fractionate the omass into its three major components and help to ing fast conversion of cellulose into glucose in high yield. In addition, we want to make sure at we will successfully develop a number of alter- e processes for utilizing hemicellulose hydr- e.

The Purdue Project supported by the DOE Fuel From Biomass Program was started in January 1, 1978 with a 12-month funding of $210,000. On January 1, 1979, the contract was renewed for another year with support of $230,000. Our target is to complete our process research by the end of 1982. The overall process will include six sub-areas;

(1) Hydrolysis of cellulose
(2) Fermentation of cellulose hydrolysate
(3) Hydrolysis of hemicellulose
(4) Fermentation of hemicellulose hydrolysate
(5) Recovery of alcohol and other products
(6) Treatment of waste liquors in methane generation

While the laboratory research is going on, we are trying to carry out pilot plant scale process research. With funds from the Indiana State General Assembly, we have been able to purchase and install some pilot plant equipment. The target date is the later part of 1980, by then we hope to have a basic process ready for designing an industrial scale factory. The simultaneous laboratory process research, meanwhile, will help to optimize the process conditions. By 1982, the construction of the industrial scale factory is expected to be completed and by then, we should also have accumulated a sufficient amount of laboratory data to assist in the start-up and operation of the large factory.

APPROACH

(a) Use of various solvents to destroy the cellulosic crystalline structure and fractionate the three major components of cellulosic materials.
(b) Improvement of techniques of recycling the solvents.
(c) Fermentation of pentoses to yield alcohols and other useful products to assure a good by-product credit to help pay for the cost of cellulose conversion into ethanol.
(d) Improvement of ethanol fermentation from glucose by developing continuous, mold, packed bed reactors.
(e) New method of recovery of alcohols and other products with low energy cost and high energy efficiency.
(f) Methane generation from alcohol stillage wastes to create an additional energy source from the

overall process. We do not believe that drying the wastes to produce stillage grain is energy efficient enough to be feasible.

KEY RESULTS

(a) After the solvent treatment, cellulose has been easily hydrolyzed to glucose in 90+% yields.
(b) Production of butanediol from pentoses by fermentation has been successfully done in 90+% of theoretical yield in concentrations of about 100 grams per liter.
(c) Between (a) and (b), we effectively increased the liquid alcohol yield by nearly 300%.
(d) Developed a bacterial culture that can produce ethanol or butanediol or both from pentoses.

FUTURE PLANS

(a) Continue the current work on the use of solvents and pentose fermentation.
(b) Expand substantially into alcohol recovery and dehydration research.
(c) Develop a plant design based upon the overall biomass utilization including cellulose, hemicellulose and lignin.
(d) Develop a plant design based upon utilization of hemicellulose only. Hemicellulose is so easy to hydrolyze that the pentose cost can be very low.
(e) Modify fermentation processes to yield products that are easily recoverable. Integrate fermentation and product recovery in process design and research.
(f) Fermentation research and culture adaptation for production of ethanol from pentoses using bacterial and/or mold cultures.

PROCESS DEVELOPMENT STUDIES ON THE BIOCONVERSION OF CELLULOSE AND PRODUCTION OF ETHANOL

(Contract W-7405-Eng-48) (Aug. 1, 1978)

Charles R. Wilke
Harvey W. Blanch
Lawrence Berkeley Laboratory
and
Department of Chemical Engineering
University of California
Berkeley, California

Aldo F. Sciamanna
Steven L. Rosenberg
S. Kishen Tangnu
Ray P. Freitas
Lawrence Berkeley Laboratory
Berkeley, California 94720

ABSTRACT

is research program has investigated the hy- olysis of cellulose and hemicellulose to sugars l their subsequent fermentation to ethanol for e as a liquid fuel. Agricultural residues, rest product residues and whole tree biomass have en analyzed and assessed for yields of sugars and anol under various processing conditions. Pro- ssing concepts which have been investigated clude both chemical pretreatment and acid and zymatic hydrolyses of cellulose and hemicellulose glucose and xylose, respectively. These sugars e then converted to ethanol using high-rate ntinuous processes in which ethanol is removed om the fermentation broth.

SCRIPTION OF TASK

is research program is directed toward the con- rsion of the carbohydrate content of cellulosic terials to produce ethanol and other energy aring products. As the product of photosynthesis, rbohydrates may be viewed as a form of stored lar energy. Vast quantities of cellulose in the rm of whole trees, agricultural and domestic stes are available, and additional supplies could specifically grown to provide biomass for con- rsion. The Department of Energy program has ncentrated on the hydrolysis of cellulose to gars, and the subsequent fermentation of these gars to ethanol, a potential liquid transpor- tion fuel.

ocess development studies and pilot plant studies ve been concerned with various processing schemes r the optimal economic production of ethanol, and ese studies are continuing to refine the overall rkeley processing concept as fundamental infor- tion becomes available from research supported by e Basic Energy Sciences Division of the Dept. of ergy.

e present research program has concentrated on e hydrolysis of cellulose and hemicellulose to gars and their subsequent fermentation to ethanol r use as a liquid fuel. A range of candidate llulosic materials, mainly in the form of agri- ltural residues, forest product residues and ole tree biomass, have been analyzed and assessed r yields of sugars and ethanol under various ssing conditions. Processing concepts under study include both chemical (acid) pretreatment of these cellulosic materials to remove lignin and hemicellulose, enzymatic and chemical hydrolysis of α-cellulose to glucose, and conversion of pentose and hexose sugars to ethanol by fermentation. Fundamental problems related to the development of optimal processing conditions have been studied under the auspices of the Basic Energy Sciences while more applied aspects and pilot plant studies are sponsored by the Division of Solar Energy.

Important basic problems considered include the kinetics of enzymatic hydrolysis of cellulose, the physical and chemical nature of the raw materials, production and recovery of the enzymes involved, and the development of new organisms for direct conversion of cellulose to ethanol.

Major developmental studies under the auspices of the Solar Energy Division include development of chemical treatment processes, studies on optimal cellulase production, enzymatic hydrolysis of hemicellulose and subsequent fermentation of the pentose sugars to ethanol and optimization of a continuous removal process for the ethanol fermentation using low pressure distillation. In all these studies the economic impact of each process change is evaluated.

OBJECTIVES

The basic processing steps involved in the hydrolysis of cellulose containing materials to produce sugars, and the subsequent fermentation of these sugars to ethanol that have been considered are:

1. pretreatment of the raw material to render it more accessible to acid or enzymatic hydrolysis.
2. acid or enzymatic hydrolysis. One or two sugar solutions may result at this stage, these being either xylose and glucose or a mixed pentose and hexose solution. The acid or enzyme then needs to be recovered.
3. concentration of the resulting sugar solution. It may also be necessary to remove any material which may be inhibitory to the subsequent alcohol fermentation.
4. ethanol fermentation, and separation of ethanol from the fermentation broth.

In each of these stages our objectives have been to develop cost efficient techniques. Process economics indicate the cost sensitive areas in these steps are the recycle of acid used in either pretreatment or later hydrolysis, the production and recovery of enzymes in enzymatic hydrolyses, the utilization of pentose sugars for ethanol formation, ethanol separation from the fermentation broth and the cost of supplemental media constituents in the ethanol fermentation. A summary of the major costs involved in ethanol production from cellulose is given in Table 1, from Wilke, et al.[1]

Table 1

PROCESSING COST DISTRIBUTION FOR ETHANOL PRODUCTION

	¢/gal 95% ETOH	% Total Costs
Glucose	135.9	75.7
Glucose Concentration	5.2	2.9
Fermentation	7.6	4.2
Medium Chemicals	21.4	12.0
Distillation	3.0	1.7
Methane Generation	6.3	3.5
	179.4	100.0

Of the costs involved in glucose production, the most significant are those relating to enzyme production and the actual raw material costs. This arises as there is no effective means at present of recovering and recycling the cellulase enzymes.

Based on these considerations, the objectives of the Berkeley program can be summarized as:

1. development of effective acid hydrolysis schemes. Various types of pretreatments have been investigated which render the cellulosic material more susceptible to hydrolysis. The cost of the acid involved usually dictates that recovery and recycle of the acid is necessary.
2. optimization of conditions for the production of enzymes for hydrolysis of the raw material. Both cellulase and xylanase enzymes, from Trichoderma viride and Streptomyces xylophagus, are being used. Continuous production schemes, involving two stage continuous culture with cell recycle, are being optimized for both organisms. As new mutants of both organisms are developed at Berkeley and by other Department of Energy contractors, they are evaluated for high rates of enzyme production. Various combinations of both xylanase and cellulase enzymes from different organism sources are evaluated to develop an optimal enzyme mixture.
3. determination of the kinetics of enzyme hydrolysis of cellulose to allow a rational design of the hydrolysis reactor. This involves the effects of the nature of the substrate and product inhibition by both cellobiose and glucose. Analogous information on xylan hydrolysis is being obtained.
4. investigation of the effect of other metabolites, such as pigments manufactured by the cellulase producing organisms, on ethanol formation by yeast.
5. recovery of the cellulase enzymes from the hydrolysis reactor and recycle of these enzymes. Various techniques, including introduction of materials which prevent the strong binding of enzyme to cellulose, are being evaluated. Enzyme recycle will substantially reduce hydrolysis costs.
6. determination of optimal reactor configurations for high-rate ethanol production. High-cell density systems involving cell recycle are being evaluated, and continuous ethanol removal via a vacuum flash pot is used to remove ethanol inhibition and provide low cost ethanol-water separation. Nutrient requirements for yeast growth under these conditions are being optimized.
7. development of alternate low energy separation systems to recover ethanol from the fermentation broth. Processing concepts included here are membrane separation and solvent extraction.
8. evaluation of each of the above objectives i terms of overall process economics. This is done to pinpoint key cost sensitive areas.

APPROACH AND RESULTS

Chemical Treatment Processes

Various forms of chemical pretreatments have been examined with the objective of increasing sugar yields from subsequent enzymatic or acid hydrolyse The most promising in terms of yields and economic are sulphuric and hydrochloric acid treatments. Other treatments which have been evaluated include nitric oxide [2], ferric sulphate [3] and sulphur trioxide [3]. Concentrated sulphuric acid (85 wt% when used together with mechanical shear can yield over 70% sugar conversion, with an initial ratio o 0.6 lb of 100% H_2SO_4 per pound of wood hydrolyzed. Complete recovery and recycle of the sulphuric aci in this process has not yet been demonstrated. Using hydrogen chloride under pressure, however, offers advantages as HCl is more reactive in the gas phase, and lower reaction temperatures and shorter reaction times are possible as compared to other acids. Due to its volatility HCl is readily recoverable and can thus be recycled. Current processing involves use of 50-80 wt% HCl at 20-43 wt% solids loadings. After reaction a secondary hydrolysis in dilute acid is performed to obtain sugar monomers. Up to 85% conversion of glucose and 78% conversion of total sugars has been obtained. HCl can be recovered by a single stage vacuum distillation at 40°C.

Enzyme Hydrolysis

Cellulase Production

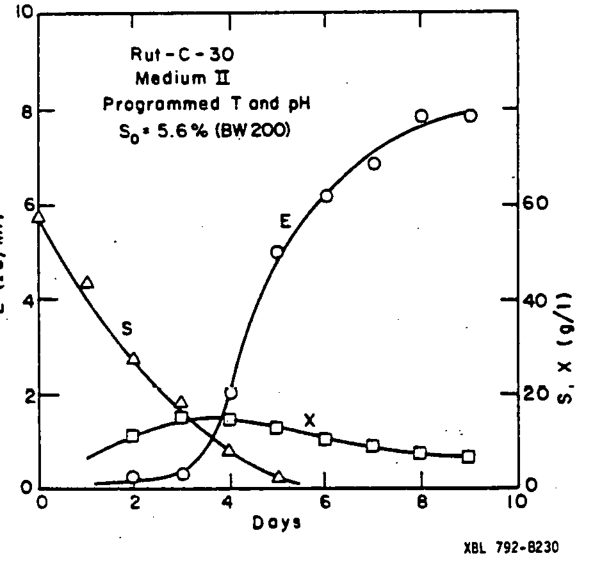
Efforts have been directed toward optimization of pH and temperature in two-stage continuous culture systems for cellulase production. This effort has been directed toward determining the effects of temperature, pH, dissolved oxygen, carbon and nitrogen sources, C/N ratios and media additiv in both batch and continuous systems. Enzyme ies are measured as filter paper activity, and β-glucosidase activity. The fungus erma viride has been shown to be the most l cellulase producer so far. Strain QM9414 eloped at the Natick Laboratory and Rutgers 4 and C-30 hypercellulase producers have been luated. Table 2 summaries the optimal environ-tal conditions for strain QM9414 and levels of yme production.

Table 2

IMAL PRODUCTION CONDITIONS FOR CELLULASE FROM VIRIDE (QM9414) IN TWO-STAGE CONTINUOUS CULTURE

riables	1st stage	2nd stage
	3.75-4.25	5.0
mperature (°C)	28°	28°
lution rate (h^{-1})	0.02	0.027
let substrate concentration (wt%)	1.75	-
A (IU/ml)	4.4	5.6
oductivity (IU/ℓ.hr)	0.088	0.11

ure 1 shows typical batch growth results of the -C-30 strain, with temperature and pH gramming.



. 1. Batch Fermentation of Rut-C-30 (T. viride) h temperature and pH programming.

individual enzyme activities of Rut-C-30 grown 5.6% Solka Floc (BW200) as shown in Fig. 2. It apparent that all components of the cellulase produced in large amounts.

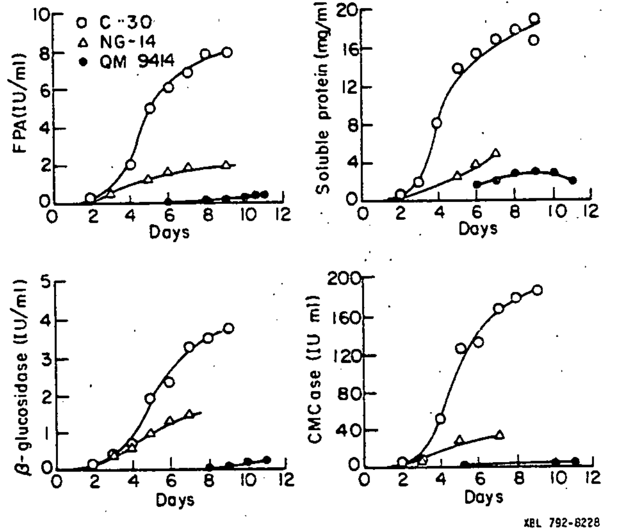


Fig. 2. Comparison of individual enzyme activities of the three T. viride strains.

Hydrolysis Kinetic Studies

The successful design of an optimal enzymatic hydrolysis reactor is predicated on a reliable model of the enzyme kinetics. A first approach has been to consider that only a fraction α of the substrate is accessible, and that this factor decreases as more product is formed, i.e.

$$\alpha = \alpha_o \, e^{-K_1 P}$$

Combining this with the usual kinetics for product inhibition yields

$$\frac{dp}{dt} = \frac{V_M \cdot \alpha_o \cdot \exp^{(-K_1 P)} \cdot (S_o-P)}{(K_M/\alpha_o)(1+P/K_2) + (S_o-P)\exp(K_1 P)}$$

Predictions based on this model for various substrate concentrations are shown in Fig. 3.

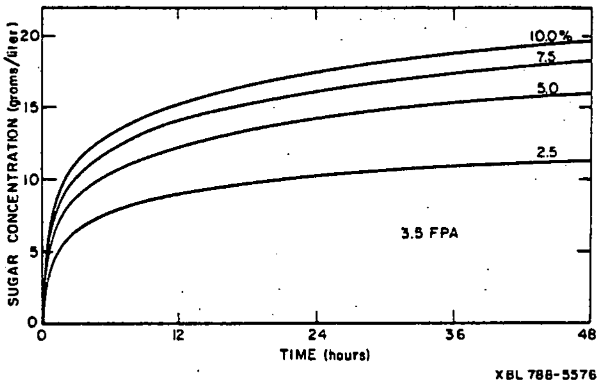


Fig. 3. Predictions for batch hydrolysis with 3.5 FPA enzyme concentration and various substrate levels.

Typical experimental data on batch hydrolysis are shown on Fig. 4. As can be seen, there is a high degree of correspondence between model and experiment, although no attempt has been made to match enzyme constants for this particular data set.

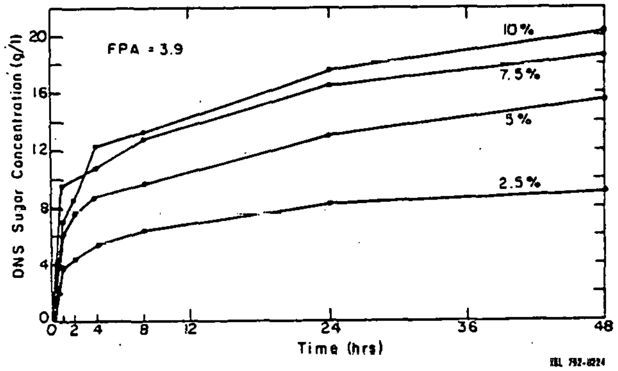


Fig. 4. Hydrolysis of Various Concentrations of 2MM Wiley Milled Corn Stover by T. viride cellulase (FPA = 3.9).

Using cellulases from different organism sources has been effective in increasing hydrolysis rates [4], as the individual enzyme components are present in varying amounts. Systems with increased levels of β-glucosidase activity have been investigated.

Enzyme Recovery

As a significant cost is associated with cellulase production, various treatments of the residual solids from enzyme hydrolysis have been examined with the objective of releasing adsorbed cellulase from the unhydrolyzed cellulose. Of those investigated to date, urea was found to be the most effective desorbing 45% of the adsorbed enzyme. Results of recombining wash solutions and hydrolysate suggest that 70% of the original enzyme can be recovered in this manner. Among various amounts of urea introduced to the hydrolysis reactor, 0.9M concentration has the optimal effect of retaining 67% of the original filter paper activity in the liquid phase. This is demonstrated in Fig. 5.

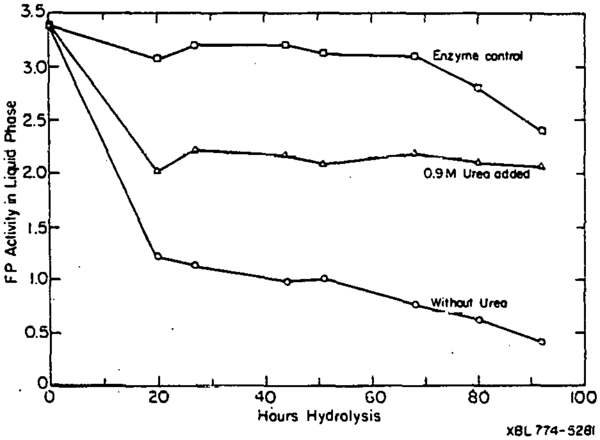


Fig. 5. Enzyme Activity Remaining in Solution during hydrolysis in the Presence of 0.9M urea

Ethanol Fermentation Studies

Nutritional Requirements

The hexose sugar solution produced by either acid or enzymatic hydrolysis is fermented to produce ethanol by Saccharomyces cerevisiae. With some strains of T. viride it has been found that metabolites which are also produced (primarily pigments) may be inhibitory to subsequent ethanol fermentation. These may be removed by adsorption onto activated carbon. Strain Rut-C-30 does not appear to have this inhibitory effect on the yeast as is seen on Fig. 6.

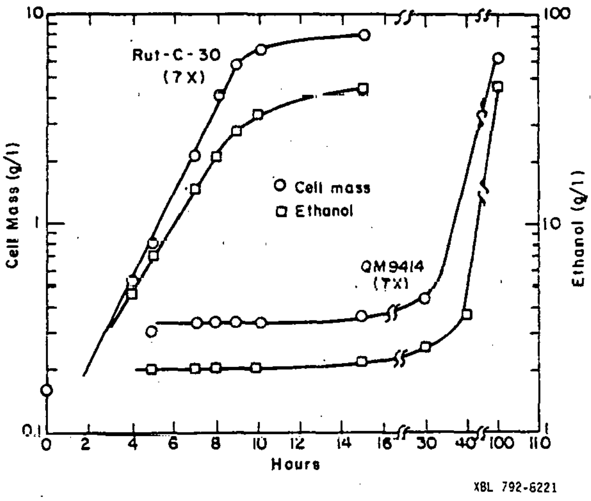
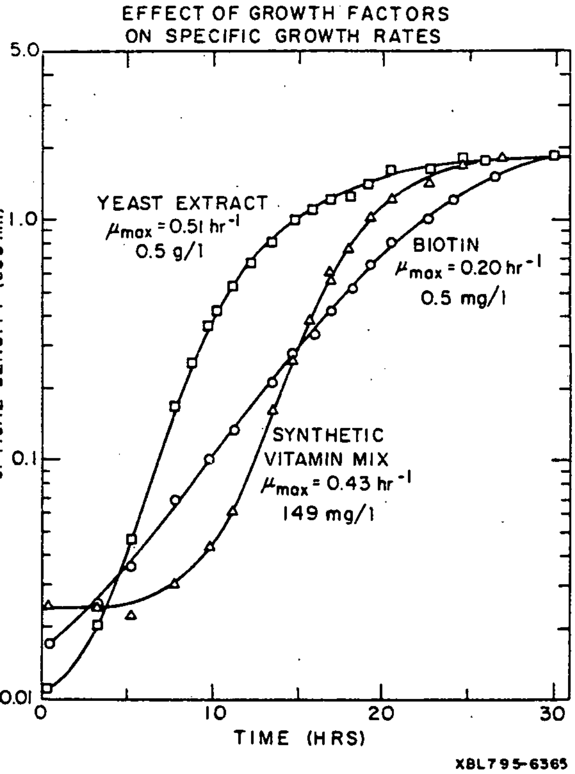


Fig. 6. Production of Ethanol and Cell Mass fr[illegible] Seven-Fold Concentrated Extracts Derived from T viride Strains.

ia chemicals, in addition to the sugar source, se a large cost in ethanol fermentation, s have examined the growth requirements of for vitamins and dissolved oxygen levels. sults to date indicate biotin to be critical in hieving rapid ethanol production. Final cell and hanol yields are independent of vitamin source d even absence of vitamins, but rates are fected, as can be seen in Fig. 7.



g. 7. Effect of Growth Factors on Cell Yield and ecific Growth Rates of S. cerevisiae in batch stems.

Process Development Studies

r current fermentation design involves a high ll density atmospheric fermentation in which end oduct inhibition is removed by cycling fermenter er to a vacuum flash pot where an equilibrium hanol-water vapor mixture (ca 22 wt% ethanol) boiled away. Using a 50 wt% sugar solution as ed will just balance the water carried over in e flash pot vapor product stream. This allows e advantages of vacuum fermentation to be ntained at substantially reduced costs by use of vapor recompression heated auxiliary vacuum flash t. Flash pot operating costs are approximately /hr for a 12 million gallon/year plant; the ocessing is shown schematically in Fig. 8.

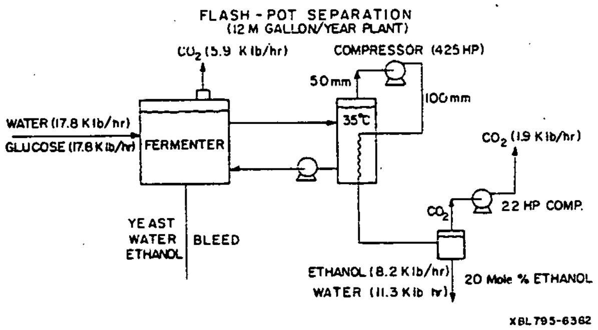


Fig. 8. Vacuum Flash Pot Separation of Ethanol from fermentation beer. Feed Streams are based on a 12 million gallon (95%) Ethanol/Year Operation.

The reduction in water load on the system represents considerable energy savings by reducing the water which must be heated to the bottoms temperature in the final distillation to produce pure ethanol. Further reduction in energy for ethanol separation can be realized by vacuum distillation. At sufficiently low pressures the azeotrope is removed, but the "pinch point" at high ethanol concentrations is so severe as to make complete separation by distillation unfeasible. At 95.6 wt% the atmospheric azeotrope) the pinch is greatly reduced in vacuum distillation and substantially reduced. Energy requirements are thus reduced an additional 22% from 20,400 BTU/gal (for flash pot and atmospheric column) to 16,000 BTU/gal. Column sizing is increased by two-fold in diameter and a small compressor must be added to compress CO_2 dissolved in the column feed up to atmospheric pressure. The process is shown schematically in Fig. 9.

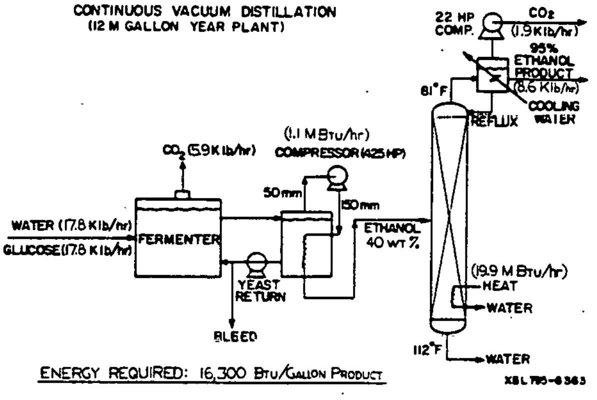


Fig. 9. Continuous High-Cell Density Ethanol Production, Using Vacuum Flash Pot and Vacuum Distillation.

Utilization of Hemicellulose Sugars

As up to one-third of the total carbon available in biomass used as a raw material may be in the form of pentosans, effort has been directed toward utilization of these pentose sugars for ethanol production. These pentoses may be released by either preliminary dilute acid hydrolysis of the raw material, or by the action of xylanase enzymes. Xylanase production by Strep. xylophagus in continuous culture has been optimized using wheat bran as a substrate. At the optimal dilution rate 0.027 hr^{-1}, the enzyme activity is 7.25 mg/ml.

Formation of ethanol from xylose has been investigated using Bacillus mascerans, which has been selected on the basis of its high ethanol productivity. It appears that the organism is inhibited by both substrate and product. In batch cultures, an initial xylose concentration of 2.3 wt% appears to yield the highest conversions. μ_{max} for this organism is 0.15 hr^{-1}, with a 27% conversion of xylose to ethanol. Typical batch data, including specific activities is shown in Fig. 10.

BACILLUS MASCERANS GROWTH ON XYLOSE

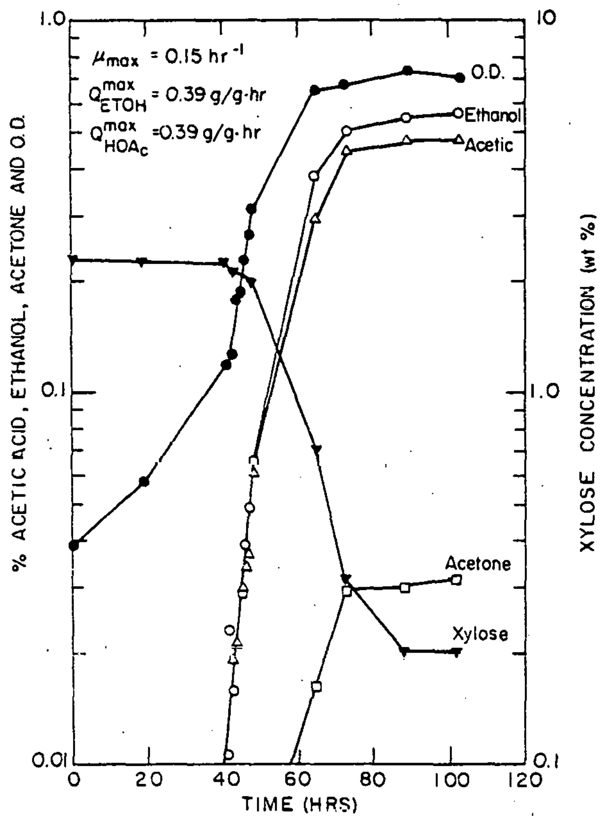


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Fig. 10. Growth of B. mascerans on Xylose. The Major Products Formed Are Ethanol, Acetic Acid and Smaller Amounts of Acetone.

FUTURE PLANS

As has been described in the preceding sections, effort will be directed toward those cost-sensitive areas in the overall processing schemes. These include:

1. development of high pressure HCl hydrolysis with recycle of the acid.
2. optimization of cellulase production and hydrolysis reactor development.
3. optimization of the ethanol fermentation and low cost ethanol-water separation schemes.
4. utilization of the hemicellulose fraction of the raw material to produce ethanol.

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IMPROVEMENTS IN CELLULASE PRODUCTION BY *TRICHODERMA* THROUGH MUTATION AND SELECTED BREEDING

Bland S. Montenecourt, Denis H. J. Schamhart,
Steve M. Cuskey and Douglas E. Eveleigh
Department of Biochemistry and Microbiology, Cook College
Rutgers - The State University of New Jersey
New Brunswick, New Jersey 08903

BSTRACT

ficient conversion of cellulose to glucose syrups s a prerequisite if biomass is to be counted as a iable starting material for fuels. Biochemical onversion employing cellulase enzymes offer everal advantages over the traditional acid ydrolysis as enzymes are non-polluting, reusable, equire low energy input while catalyzing high onversion efficiencies into a pure glucose pro- uct. Enzymatic conversion has not been achieved ommercially due to the high costs of the cellu- ases from the available microbial strains. The ajor focus of this program is to lower the cost f these enzymes through the isolation of mutant trains which synthesize the cellulase in high ields and catalyze more efficient conversion of he cellulose to glucose. A number of selective creening methods have been devised that allow etection of mutants of *Trichoderma reesei* which re capable of overcoming the biochemical and enetic controls operative during cellulase ynthesis and activity. Selected breeding is eing developed employing the parasexual cycle and rotoplast fusion in order to sequester desired haracteristics into a single strain. Current mphasis in the program has been to characterize hese mutants especially in relation to their ndustrial usefulness.

NTRODUCTION

he hydrolytic conversion of cellulose to glucose n the overall scheme to produce alcohol from oody biomass is an economic bottleneck. This ydrolysis of cellulose can be accomplished ither by acid or by use of enzymes (cellulases). le favor the enzyme approach in that high con- ersion efficiency can be obtained without the use f extreme temperature or pressure. The overall ost of enzymatic hydrolysis can be reduced sub- tantially through the selection of microbial trains whose cellulase complex is produced in igh yield, with optimal proportions of each nzyme component and whose physicochemical haracteristics allow their large scale applica- ion under optimal fermentor conditions. Methods ave been developed that allow selective isolation f such desirable cellulolytic strains. *richoderma reesei* has been the initial micro- rganism of choice with regard to its relatively igh cellulolytic activity towards crystalline llulose. This report describes the screening odologies and illustrates the characteristics of the most useful mutants isolated to date.

RESULTS

Enzymes in the cellulase complex and their controls

The sequential cooperative activity of at least three distinct enzymes is required for the efficient hydrolysis of crystalline cellulose: endo-β-glucanase, cellobiohydrolase and β-glucosidase. The current hypothesis of how these enzymes interact is summarized in Figure 1. Formulation of this model has resulted through contributions of a number of independent workers [1, 2, 3, 4, 5, 6]. The endoglucanase is thought to initially attack the cellulose randomly. The action of the endoglucanase is followed by the action of cellobiohydrolase releasing cellobiose from the non-reducing end of the cellulose chains. The continued cooperative action of the endoglucanase and the cellobiohydrolase result in conversion of the cellulose to cellobiose and small oligosaccharides. The latter are then acted upon by β-glucosidase to yield glucose.

The synthesis and activity of the cellulase complex are subject to a number of rigid biochemical and genetic controls. The most prominent regulatory mechanisms are induction and catabolite repression. Microorganisms do not synthesize many enzymes unless the substrate for the enzyme is present in the environment. Thus, the presence of insoluble cellulose or other inducers are required for cellulase synthesis. In addition, if other readily metabolizable material such as glucose is present, in addition to cellulose, the organism will preferentially utilize the glucose and the synthesis of cellulase will be repressed. The combination of induction and repression of the cellulases is responsible for the low yields obtainable during enzyme production in the fermentor. In addition to the genetic controls, the activity of each of the cellulases is subject to end-product inhibition. In this case, the products of the reaction, glucose, cellobiose and perhaps small oligosaccharides bind to the enzyme and inhibit further activity. This end-product inhibition can be responsible for the low efficiency of enzymatic conversion of cellulose to glucose. End-product inhibition resistant enzymes could allow 100% hydrolysis of cellulose to glucose and greatly reduce the cost of

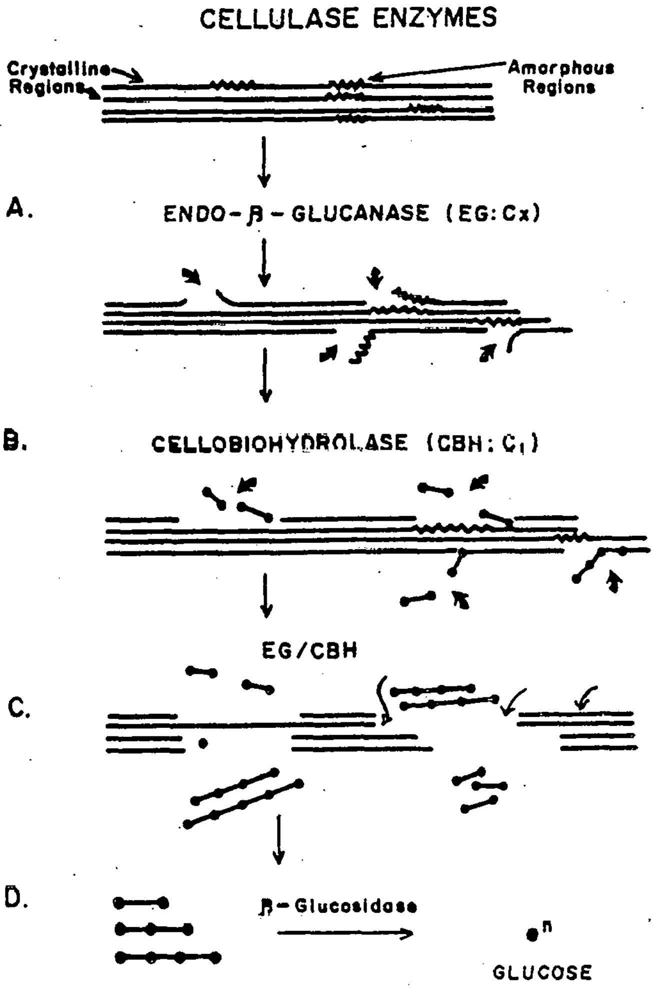

Figure 1. Schematic representation of sequential stages in cellulolysis

biochemical conversion.

The objectives of the Rutgers program are to select genetically altered strains of *T. reesei* which are no longer subject to these regulatory mechanisms. Two approaches have been employed, direct selection of mutant strains resistant to the controls and selective breeding of desirable strains through heterokaryon formation and protoplast fusion.

Selective screening for mutant strains.

Table 1 summarizes the selective screening methodologies developed to date which allow visual detection of the desired mutant strain in the presence of large numbers of non-mutant organisms [7, 8]. As *T. reesei* is a fast growing soil fungus, colony inhibitors Oxgall and Phosfor D are routinely added to the agar plates allowing formation of compact well defined colonies. The substrate for the desired enzymatic activity is also incorporated into the medium e.g. esculin fo the selection of β-glucosidase mutants. If catabolite repression resistant mutants are sough high concentrations of the repressing substance a also incorporated, e.g. 5% glucose or 5% glycero If constitutive mutants are desired (mutants whic no longer require an inducer) any non-inducing substrate is incorporated into the base agar and the detecting substrate is incorporated in an agar overlay. Mutants with enzymes that are end-product inhibition resistant are selected on the basis of their ability to catalyze enzymatic conversion in the presence of high concentratio [illegible] the specific end product inhibitor. Visuali

TABLE 1. SCREENING METHODOLOGIES FOR ISOLATION OF HIGH YIELDING CELLULOLYTIC REGULATORY MUTANTS OF *TRICHODERMA REESEI*.

Enzyme	Colony Inhibitor	Substrate	Catabolite Repressor (5%)	Visualization
Total Cellulase	Oxgall (1.5%) + Phosfon D (500 µg/ml)	Walseth cellulose	Glycerol	Clearing - incubation 50°C
Endoglucanase		CMC	Glycerol or Glucose	Clearing - quaternary ammonium compounds
β-Glucosidase		1. Esculin	Glucose	Black rings - ferric ammonium citrate
		2. Cellobiose (1%)	2-Deoxyglucose (0.2%)	Large vs. point colonies
		3. CH_3·Umbelliferyl Glucoside	Glucose	Fluorescence
		4. Any Glucoside Substrate	Glycerol	Glucostat + Cellobiose 50°C, 30 min: Red
Constitutive		A Non-Inducer	None	Selective Overlay
E.P.I. Res^{n}.		Any Cellulosic Substrate	Presence of End Product	Visual Selection

f the enzymatic activity is the key to the selec-ion system. For example, esculin is a substrate or β-glucosidase. Under conditions where β-lucosidase is synthesized and active, it will ydrolyze the esculin into glucose and esculetin. he latter compound esculetin reacts with the erric salts present in the agar to form a black recipitate. Mutants will therefore have black ings surrounding the colonies where as the wild-ype will be colorless.

utgers Family of *T. reesei* mutants.

tilizing these selective screening methodologies pproximately 100 mutants have been isolated and oughly 800 thousand potential mutants screened. genealogy is outlined in Figure 2. The NG eries represent high yielding mutants derived in wo steps from the wild type *T. reesei* QM6a. UT-NG14 is the best characterized mutant of this roup. Fermentor studies at Natick have shown hat this mutant is capable of synthesizing 2% xtracellular protein and 45 units/l/hr of the ellulase enzyme complex. These enzymes show in-reased thermostability when compared with QM9414 s demonstrated in Figure 3. This latter char-eristic may allow hydrolysis to be carried out mperatures greater than 50°C which may enhance the rate and efficiency of saccharification.

Two catabolite repression resistant lines have been isolated from RUT-NG14, the "C"-series and the "L"-series (Fig. 2). Of the C-series, RUT-C30 has been best characterized. Parallel cooperative studies at Natick and at the University of California, Berkeley have verified that RUT-C30 shows more rapid growth and cellulase synthesis resulting in productivities of 100 units/l/hr. under controlled fermentor conditions. The Berkeley group has also demonstrated that the sugar syrups produced by the action of RUT-C30 cellulase are lacking an inhibitor of fermentation routinely found in syrups produced by QM9414. Thus the glucose syrups, produced by the action of RUT-C30 cellulase on cellulose can be fermented directly by yeast eliminating the previously employed charcoal filtration step.

Mounting evidence suggests that the β-glucosidase of *T. reesei* cellulase is the rate limiting step in the enzymatic conversion of cellulose to glucose [9]. Not only is this enzyme proportionally low in the cellulase complex, but β-glucosidase is severely inhibited by the end-product, glucose. Major effort in our program has

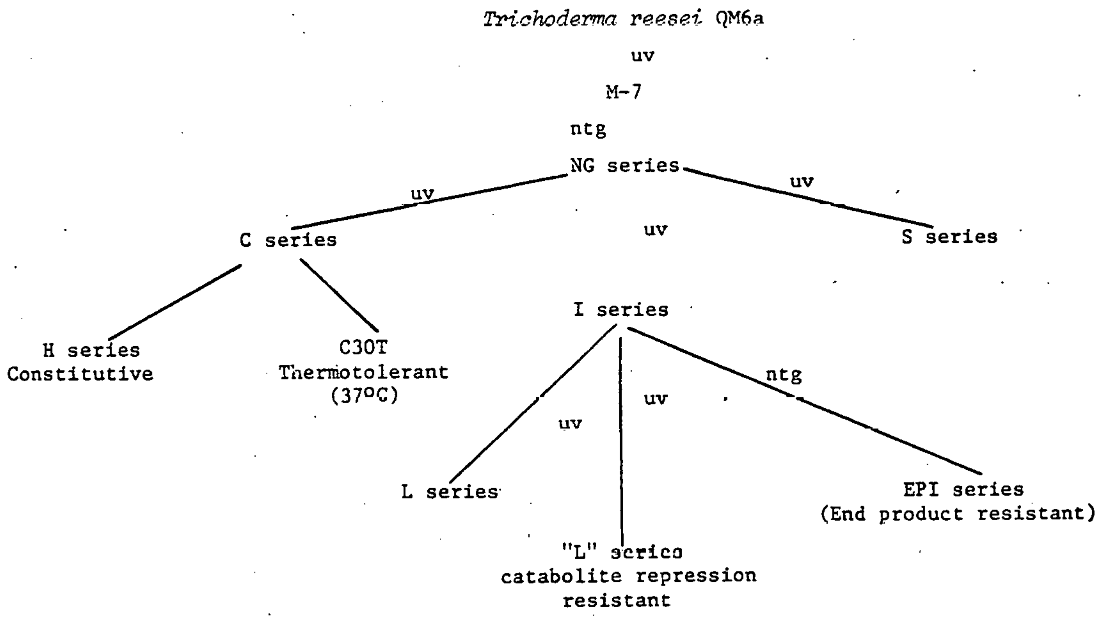

Figure 2. Rutgers family of *T. reesei* mutants.

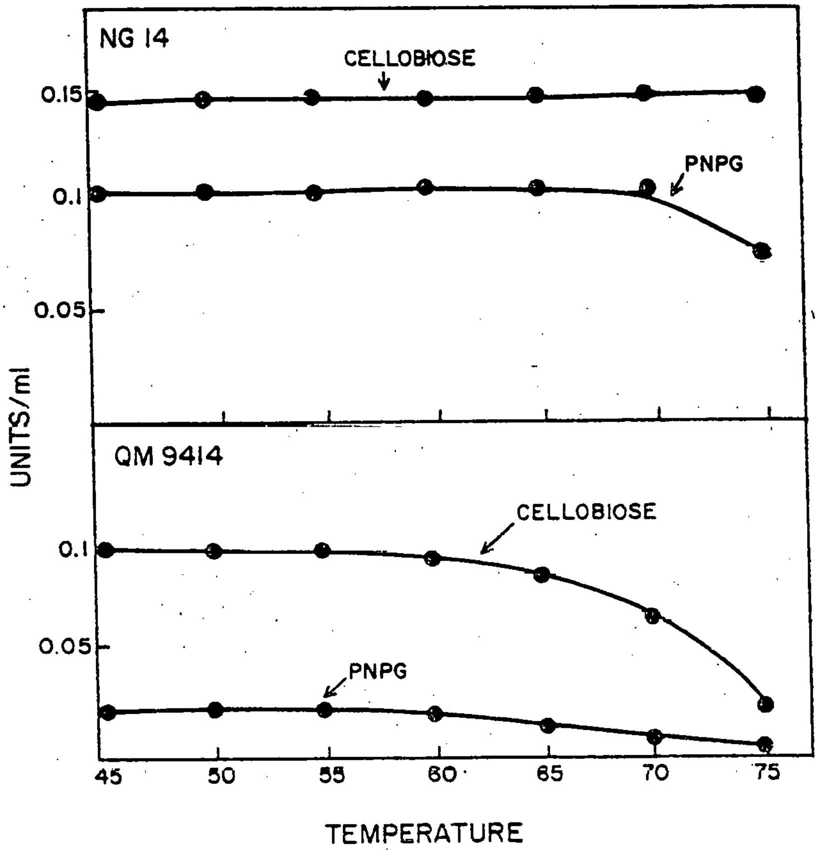

Figure 3. Thermostability of the β-glucosidase of RUT-NG14 as compared to QM9414; p-nitrophenyl β-glucoside (PNPG) or cellobiose as substrates.

given to increasing the yield and decreasing end-product inhibition of β-glucosidase. A er of end-product inhibition resistant mutants been isolated, EPI series. Although no completely resistant strains have been found to date, several mutants synthesize a β-glucosidase with altered inhibition kinetics with respect to glucose.

Selective Breeding of *T. reesei*.

Until recently, improvement of industrial micro-organisms has relied mainly on programs of mutation and selection. This situation is unsatisfactory in that it is purely empirical and such mutagenic treatments coincidently induce serious structural damage to the genome unrelated to the desired improvement. In an effort to allow recombination of important characteristics from different mutants into a single strain, alternative genetic techniques are being employed which circumvent continued mutation. A detailed knowledge of the genetics and the regulatory mechanisms of the cellulases of *T. reesei* is essential for further improvement of high yielding cellulolytic strains. Unfortunately *Trichoderma* possesses a poorly defined sexual stage (*Hypocrea*), limiting natural recombination to the parasexual cycle. Marked strains carrying abnormal spore color, antibiotic resistance and auxotrophic nutritional requirements have been isolated from the different Rutgers mutant lines to facilitate such genetic analysis. Although heterkaryons (hyphal cells containing nuclei derived from two strains) have been readily formed by allowing cultures to grow together and fuse with one another, initial attempts at isolating heterozygous diploid spores from the heterokaryons have been unsuccessful. Thus we are in the process of developing the relatively new approach of protoplast fusion to achieve somatic hybridization of the strains. This technique involves removal of the cell walls of the parent strains with a commercial lytic enzyme preparation. Methods to obtain stable protoplasts have been devised and regeneration of stable cells (with walls) from the osmotically sensitive protoplasts has been achieved. The major objective now is to fuse the protoplasts in the presence of polyethylene glycol and isolate stable hybrid strains containing marker complementation which can regenerate their cell walls and grow.

SUMMARY

Genetic techniques of mutation and selection, parasexual breeding and protoplast fusion are being applied to *Trichoderma reesei* in order to improve the yield and the biochemical nature of the cellulase enzymes. Approximately 100 distinct mutants have been isolated which show resistance to catabolite repression, enhanced yield of cellulase and partial resistance to end-product inhibition. Mutants of this type which have lost the natural genetic and biochemical mechanisms governing the synthesis and activity of cellulase can be used to remove the economic bottleneck in enzymatic conversion of cellulosic biomass to glucose and ultimately fuels and petrochemical substitutes.

ACKNOWLEDGEMENTS

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NOTES

ACID HYDROLYSIS OF CELLULOSIC BIOMASS

Alvin O. Converse and Hans E. Grethlein
Thayer School of Engineering
Dartmouth College
Hanover, N.H. 03755

STRACT

continuous plug flow reactor was built for acid drolysis of cellulosic materials. A single pass drolysis at 240°C with 1% acid gives 50 to 57% ucose yield for a slurry concentration of 5 to .5% solids. Glucose and xylose yield maps are veloped that indicate the trade off of product, -product, and unreacted product. For high xylose covery, a two-stage hydrolysis is needed, the rst stage at low temperature (180 - 200°C) and e second at high temperature (230 - 240°C). Fur-ral recovery from xylose may lead to important -product credits.

e flow reactor has also been used to give acid etreatments to cellulosic materials prior to zymatic hydrolysis. The yield increase by enzy-ic hydrolysis is significant for oak (21% to), corn stover (58% to 100%), and newsprint (60% 93%) as a result of the pretreatment. Thus, a bination of acid hydrolysis pretreatment with zymatic hydrolysis can achieve nearly quantita-e yields of glucose. Estimates of the cost of anol from newsprint and wood are given for one-ss acid hydrolysis, enzymatic hydrolysis, and the bination of the two.

BACKGROUND

sed on batch kinetics measured by Fagan and others Grethlein [2] proposed a detailed design for a ocess using a continuous plug flow reactor to drolyze cellulosic materials to glucose and other oducts. He assumed that the reactor would be othermal, operating at 230°C (446°F), at a pres-e of 500 psi, with an acid concentration of 1% weight sulfuric acid, and a residence time of .9 minute. Two designs were developed, one for a nt that would feed a concentrated slurry, 30% by ght solids, and a second plant that would feed slurry. He concluded that the processing costs, clusive of substrate, would range between 1.75¢ 2.45¢ per pound of sugar produced from newsprint, sprint having a glucan content of 69%. Since se costs looked attractive, it was decided to elop a plug flow reactor in order to gain more erience with such a process and to measure the etics under the conditions assumed in the design. effect of slurry concentration on yield was of ticular interest since the increase from 10% to solids reduced the processing costs by 30%.

2. RESULTS FROM INITIAL FLOW REACTOR

A process flow sheet for a continuous flow reactor is presented in Figure 1, and is discussed in detail by Thompson. [3,4] It is important to note that the feed in the form of a slurry is preheated to the reaction temperatures, and then the reaction is initiated by the injection of the acid. The reaction is quenched by flashing the product into a capillary tube where it is cooled. In this manner, a short reaction time is obtained at a high temperature, up to 240°C. As indicated in Fig. 2 Thompson found that the yields were independent of slurry concentration up to 13% by weight slurry for 100 mesh particles. The maximum yields, furthermore, were found to be only slightly below those predicted by the batch kinetics -- 50 to 52% rather than 55%.

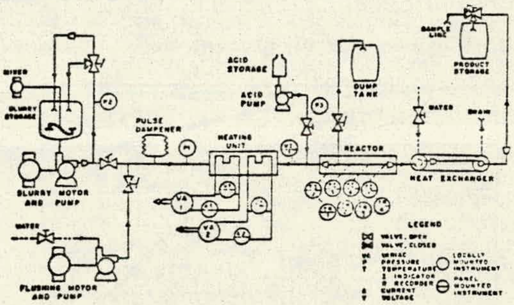


Fig. 1. Process Flow Sheet for a Continuous Flow Reactor

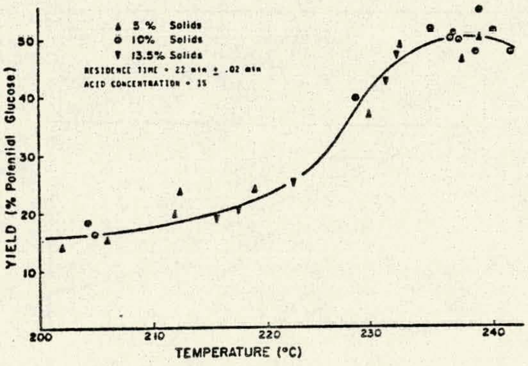


Fig. 2. Plot of Glucose Yield (% Potential) vs. Temperature

The yield maps for glucose and xylose are presented in Figures 3 and 4. The yield map for glucose, Fig. 3, indicates that a good deal of the glucose is destroyed when the reactor is operated to maximize the yield for a single pass, at temperatures in the range of 230°C to 240°C. If it is economic to recycle the solids from the reactor, then one should operate at a lower temperature and, as indicated on Figure 3, reduce the glucose yield per pass in an effort to reduce the amount of glucose decomposition and, in so doing, increase the overall yield of glucose from the recycle reactor. O'Neil et al [5] have recently concluded that an 80% glucose yield is economically feasible in a plug flow reactor with recycle of delinified cellulose.

The xylose map, Fig. 4, indicates that if one wishes to make xylose, rather than a mixture of xylose and furfural, it is necessary to run the acid hydrolysis at a lower temperature. In fact, it may be desirable to run the acid hydrolysis in two stages, with the first stage at low temperature to produce xylose, and the second stage at high temperature to produce glucose. However, if furfural is a valuable product, then there are two options: 1) to convert the xylose from a first stage to furfural in a separate process, or 2) make glucose and furfural simultaneously in one stage, high-temperature hydrolysis. More work is needed to give optimum yields of glucose and xylose or glucose and furfural.

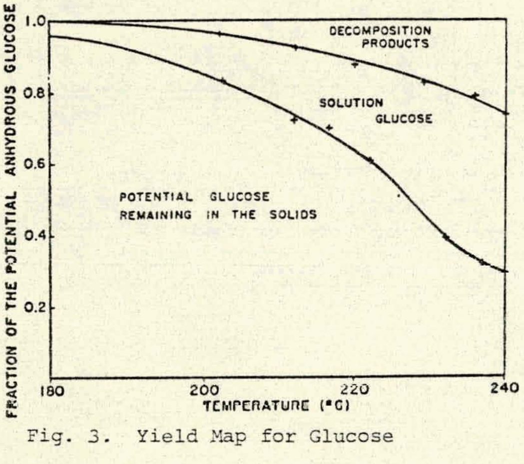


Fig. 3. Yield Map for Glucose

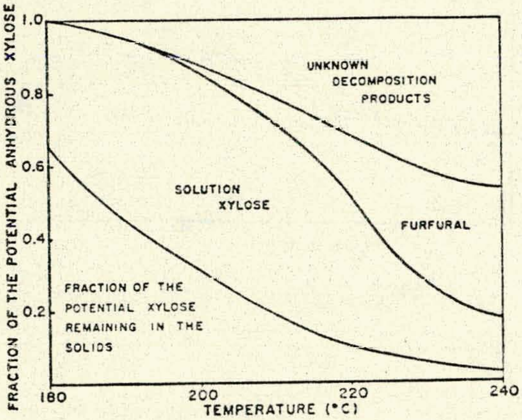


Fig. 4. Yield Map for Xylose

In short, Thompson's work further encouraged the idea of using a flow reactor for the acid hydrolysis of cellulosic materials. The experience wi that reactor, however, did point out two importa areas of difficulty. As indicated in Fig. 5, there was considerable corrosion of the Carpenter 20 steel used in the acid injection line. The Carpenter 20 alloy was selected because of its resistance to corrosion; however, under the conditions encountered here, concentrated acid at high temperature, the inside diameter increased from 3/16 to 9/32 of an inch, approximately 0.094", for a corrosion rate of 34 x 10^3 mils per year! Sections of the mixing tee also showed signs of severe corrosion. This was overcome in the laboratory by using a ceramic insert.

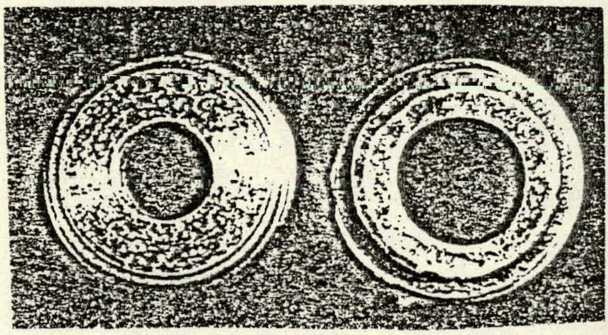


Fig. 5. Corrosion in the Acid Injection Line

The other area of concern was that of mixing of th acid with the thick slurry. Our objective is to use a highly concentrated slurry and particles up to 10 mesh, and in larger systems we would anticipate difficulty in being able to quickly mix the acid with the solids so that all the particles would have the same residence time with the same acid concentration.

To overcome these two difficulties, we proposed to develop a flow reactor in which the acid is mixed with the solids at low temperature, and in which the reaction is initiated by direct injection of steam into the slurry rather than by the injection of the acid. In this process, acid is injected while the slurry is cold and is thoroughly mixed with the slurry. The slurry is preheated indirect up to an intermediate temperature which is still too low for the reaction to take place, and then the slurry is brought up to the reaction temperature by the violent and direct injection of steam into the reactor. This avoids concentrated acid a high temperature and therefore should eliminate th corrosion problem. Furthermore, the steam injection should give adequate mixing, as indicated by flow visualization studies conducted in a 1½ inch Lexan tube. This reactor has been under development during the past year and is described in Section 4 of this paper.

3.

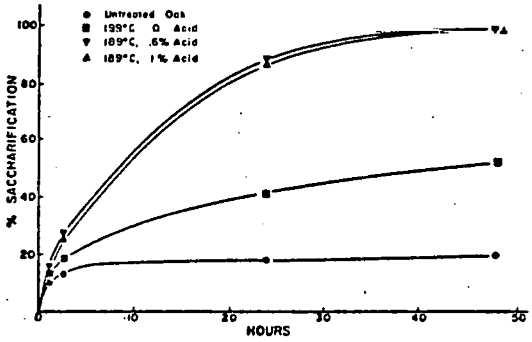
Since single pass acid hydrolysis thus far has bee limited to a saccharification yield of about 50%, and for a number of substrates the yields from matic hydrolysis are also in this range or low

studied the use of dilute acid hydrolysis as a
treatment for the substrate which is subsequently
cted to enzymatic hydrolysis. The flow reactor
eloped by Thompson was used by Knappert [6,7,8]
carry out these pretreatments. The pretreated
bstrate was then subjected to enzymatic hydrol-
is with enzyme from Trichoderme reesei provided
Natick Laboratories. These hydrolyses were car-
ed out at a temperature of 50°C, for a period of
up to two days. The results for a number of sub-
rates are summarized in Table 1.

able 1. Effect of Dilute Acid High Temperature Pretreatment on Yield from Enzymatic Hydrolysis

	Maximum % Saccharification	
Substrate	Without Pretreatment	With Pretreatment
Natural Materials		
Oak	21	90
Poplar	16	55
Corn Stover	58	100
Mechanically Processed Material		
NEP-40	58-70	93
Chemically Processed Materials		
Solka Floc	88	91
Kraft Pulp	80	71

e detailed results of the enzymatic hydrolysis
lowing pretreatment are presented for oak and
n stover in Figures 6 and 7. In order to verify
se results, corn stover was pretreated and the
duct frozen and flown to Professor Wilke's labs
the University of California, Berkeley, where a
charification of 97% was obtained, thus confirm-
the results obtained in our laboratories. As
cussed in Section 5, the economic benefits of
increased yields much more than offset the costs
pretreatment.



ig. 6. Enzymatic Hydrolysis of Oak Slurry

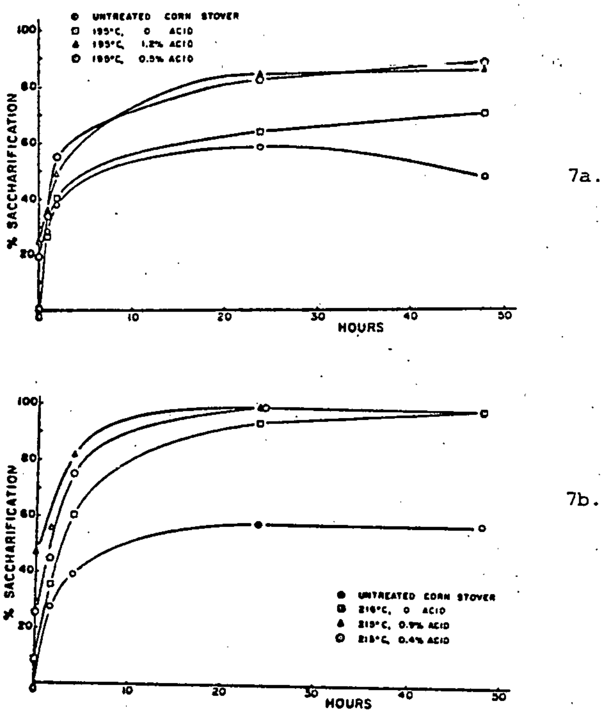


Fig. 7 Enzymatic Hydrolysis of Corn Stover Slurry

4. DEVELOPMENT OF STEAM INJECTION REACTOR

A schematic of this reactor is presented in Fig. 8.

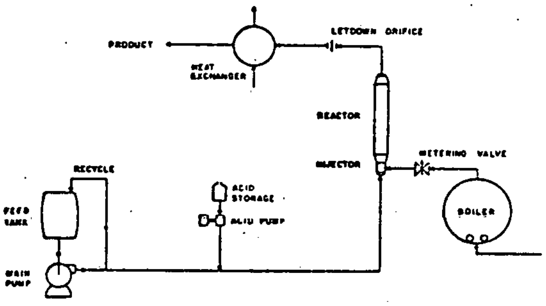


Fig. 8. Schematic of Steam Injection Reactor

A good bit of attention during the past year had to be paid to the way in which the steam is injected into the reactor. This was first attempted by using an injection lance located along the center line of the tube leading to the reactor. We were unable to operate this in a stable manner at temperatures high enough to be of interest. Steam tended to coalesce into a large bubble, and the flow and temperature regime in the reactor was not homogeneous and, furthermore, varied substantially over time. These problems have been overcome by going to a system in which the steam is injected through the pipe wall at a large number of points along the pipe wall. This has permitted us to operate in a stable manner up to temperatures of 230°C at 0.8 l/min, and we are now conducting hydrolysis experiments in the reactor. Preliminary runs have been successfully carried out on poplar, Solka floc, corn stover, cassava and cassava starch. We are now in a position to accurately measure conversion.

5. ECONOMIC EVALUATION OF DILUTE-ACID PRETREATMENT

Light fraction of municipal refuse

For the light fraction of municipal refuse Knappert et al [7] estimate that pretreatment results in a net reduction in ethanol costs of 27¢/gal, as indicated in Table 2.

Table 2. Effect of pretreatment on the cost of ethanol from the light fraction of municipal refuse via enzymatic hydrolysis with and without pretreatment
(no by-product credits: 61% glucan in feedstock costing 10$/ton)

	With Pretreatment	Without Pretreatment
	(¢/gal)	
Conversion of potential glucan, %	90%	50%
Substrate	13.0¢	23.4¢
Pretreatment	16.6	0
Hydrolysis	40.1	72.2
Sugar Concentration	1.5	2.7
Ethanol Production	30.3	30.3
	101.5	128.6

Difference = 27¢/gal

The purpose of Table 2 is to estimate change in ethanol costs caused by the pretreatment. The absolute values of ethanol cost are not to be taken as the final estimates, since credit is not taken for by-products. The substrate cost is based on 10$/ton for the feedstock. The difference in substrate portion of the ethanol cost in Table 2 is due to the difference in conversion. The pretreatment costs were developed by Knappert [6,7] and based on the acid hydrolysis design of Grethlein [2]. The hydrolysis sugar concentration and ehtanol production costs for the case without pretreatment are taken from Wilke [9]. With pretreatment, the hydrolysis and sugar concentration costs are reduced in proportion to the increased yield. This analysis indicates that pretreatment would result in a significant net reduction of ethanol cost.

Wood

A similar analysis for the case where wood is used as the substrate is presented in Table 3. The effect of pretreatment is more pronounced here because the wood has a much greater cost, a lower glucan content, and a greater enhancement of the conversion by pretreatment. Again, the absolute values should not be regarded as final because by-product credit is not taken.

It should be noted that these results for wood ar based on the results for oak (90% conversion) rat than poplar (55% conversion). However, given the results obtained with other natural substrates, such as corn stover, we expect to be able to increase the conversion obtained with poplar. Experiments with poplar were limited because it frequently plugged the reactor. The new reactor is expected to facilitate such experiments.

Table 3. Effect of pretreatment on the cost of ethanol from wood via enzymatic hydrolysis with and without pretreatment, ¢/gal
(no by-product credit; 41% glucan in feedstock costing 30$/ton)

	With Pretreatment	Without Pretreatment
Conversion of potential glucan, %	90%	25%
Substrate	56.9¢	205.0¢
Pretreatment	27.7	0
Hydrolysis	40.1	144.0
Sugar concentration	1.5	5.4
Ethanol production	30.3	30.3
	156.5	384.7

Difference = 228.2¢/gal

The difference in substrate cost is proportional to the change in conversion. The pretreatment cost is taken from Knappert [6,7]. The hydrolysis and sugar concentration costs are based on the values presented by Wilke [9]. They were adjusted upward in proportion to the lower conversion. Again, the analysis indicates that the pretreatment easily covers its cost, and results in a significant net reduction of costs.

With expectation of reduced hydrolysis cost due to improvements in the microorganism and the potentia of furfural and other by-products, these results indicate considerable promise for wood as a substrate.

6. COMPARISON OF SINGLE-PASS ACID HYDROLYSIS WITH ENZYMATIC HYDROLYSIS WITH DILUTE ACID PRETREATMENT

The comparison is presented in Table 4. The range of hydrolysis cost is due to the variation between 30% by weight and 10% slurries. Again, the reader is cautioned that the values presented are for comparison of the two processes; the absolute values should not be regarded as final.

The conclusion that emerges from Table 4 is that for high quality, inexpensive substrates, such as the light fraction of municipal refuse, single-pass acid hydrolysis still is less expensive. However, for lower quality, higher cost substrates such as wood, enzymatic hydrolysis with pretreatment is very competitive.

Table 4. Comparison of single-pass dilute acid hydrolysis with enzymatic hydrolysis with dilute acid pretreatment, in terms of ethanol cost, ¢/gal

	Acid Hydrolysis Grethlein [2]	Enzymatic Hydrolysis with Dilute Acid Pretreatment
Substrate: Light fraction (61% glucan, 10$/ton)		
Conversion, %	55%	90%
Substrate	19.8¢	
Hydrolysis	29.5 - 34.3	
Ethanol production	30.3	
	75 - 84	101.5 (from Table 2)
Substrate: Wood (41% glucan, 30$/ton)		
Conversion	55%	90%
Substrate	88.4¢	
Hydrolysis	36 - 51	
Ethanol production	30.3	
	155 - 170	156 (from Table 3)

FUTURE DEVELOPMENTS

demonstrated in the above analyses, high conver-
on is especially important when high cost sub-
rates are used. High conversion can now be ob-
ined by either pretreatment followed by enzymatic
drolysis or, very likely, acid hydrolysis with
cycle. It may also be obtained either by the
velopment of new strains of the microorganism or,
ssibly, inhibition of sugar decomposition in acid
drolysis. It is anticipated that our new flow
actor will allow further evaluation of the acid
drolysis and pretreatment aspects of these
ocesses.

KNOWLEDGEMENT

e authors gratefully acknowledge the assistance
the Department of Energy's Fuels from Biomass
ogram.

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NOTES

LIQUID FUEL PRODUCTION FROM BIOMASS

J.E. Sanderson, Diana V. Garcia Martinez, J.J. Dillon, G.S. George, D.L. Wise
Dynatech R/D Company
99 Erie Street
Cambridge, Massachusetts 02139 USA

TRACT

an extension of our previous work on the produc-
ı of methane via nonsterile anaerobic fermenta-
ı, we have modified the process to produce lon-
chain aliphatic hydrocarbons instead of methane.
s product may be used without further modifica-
ı as a fuel for Diesel engines or as a substi-
ıe feedstock for the production of gasoline.
work to date uses marine algae as the primary
strate, but work is currently in progress to
ɔt the process for use with other aquatic and
restrial forms of biomass.

the anaerobic fermentation of biomass to methane,
anic acids are produced. In the modified
mantation, the approach has been to suppress
hane formation and to extract the organic acids
n the fermentation broth for electrolytic oxida-
ı (Kolbe electrolysis) to aliphatic hydrocarbons.

liminary economic analysis based on 1000ton/day
nt design and assumed biomass costs indicates
this process is capable of producing liquid
ls at a cost competitive with the most optomis-
forecasts for alternative processes.

RODUCTION

an extension of the work on anaerobic fermenta-
ı of biomass to organic acids conducted pre-
ısly under Department of Energy sponsorship,
rogram for the production of liquid hydrocarbon
ls from biomass was initiated. The process,
presented in Figure 1, consists of non-sterile
erobic fermentation of the biomass to mixed
anic acids, extraction of the higher organic
ds (primarily butyric, valeric, and caproic
ds) and electrolytic oxidation (Kolbe Electrly-
) of these acids to aliphatic hydrocarbons,
marily hexane, heptane, octane, nonane, and
ane. The fermentation process to convert marine
ae to organic acids was developed in the pre-
ıs program, giving rapid rates and high
version yields. Figure 2 is a plot of total
ıcing equivalents of acid versus time, where
asymptotic value of 0.34 reducing equivalent
liter represents 90% of the total carbon in
substrate.

ever, in the course of this work, it was found
en other substrates were used, methane

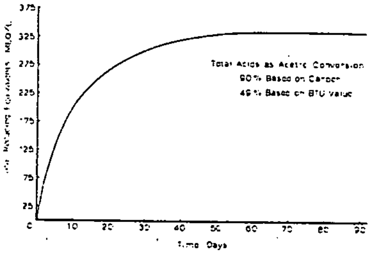

Fig. 1 Organic Acids Concentration Vs. Time
5% Chrondus crispus

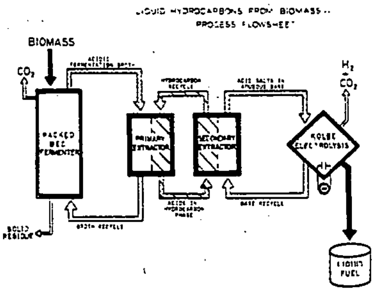

Fig. 2 Liquid Hydrocharbon from Biomass...
Process Flowsheet

Current program work has concentrated on the suppression of methane so that the fermentation process can be broadened to include other sources of biomass, particularly fresh water aquatic weeds and agricultural residues. In the initial phase of expanding the biomass usbstrates into agricultural and industrial residues, onion and garlic waste was tested. Preliminary results show (Fig. 3) that onion and garlic resiudes can be successfully converted into acids by the microbial population present in dairy manure. An interesting feature of this substrate is that no methane is formed as part of the products. Since this substrate is known to contain some antibacterial components (1), it is apparent that only methanogenic bacteria are affected by their presence.

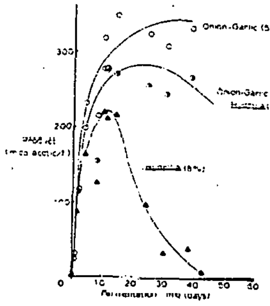

Fig. 3

As Fig. 3 indicates, addition of onion-garlic residues to Hydrilla fermentation is a potential preventive measure of methane formation. This peculiar property may be exploited by using onion-garlic residues along with other substrates in order to prevent methane formation.

It has now been demonstrated that 2-bromoethanesulfonic acid is an effective methane inhibitor for Hydrilla fermentations, and it is anticipated that this compound will also suppress methane formation from other substrates as well.

A number of approaches to methane suppression have been evaluated including fermentation at low pH (pH 5), addition of methane analogues $(ChCl_3)$, and inclusion of materials that have been reported as effective suppressors of methane formation in cattle rumen (Salinomycin, Pfizer). But the best results to date have been obtained with 2-bromoethane-sulfonic acid.

The results from a set of static Hyrdrilla fermentation experiments illustrating the efficacy of this compound in suppressing methane formation are illustrated in Fig. 4. The conversions achieved were significantly higher when the methane inhibitor was added--35%-40%--versus 25%-29% in the control. In addition, at 60 days fermentation in flasks containing the inhibitor, 100% of the acids still remained in the solution.

ENERGETICS OF THE PROCESS

In the course of developing a process for converting biomass to liquid fuels, one of the primary

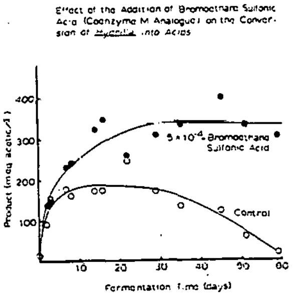

Fig. 4 Effect of the Addition of Bromethane Sulfonic Acid (Coenzyme M Analgoue) on the Conversion of Hydrilla into Acids

considerations is the energy balance of the proces which may be defined either of two ways:

$$E_B = \frac{E_O}{E_S + E_P}$$

where E_O is the energy output, E_S is the energy content of the substrate and E_P is the energy consumed in the process.

In the case where the energy of the substrate is essentially unreclaimable by other means:

$$E_B = \frac{E_O}{E_P}$$

Table 1 presents a comparison of the estimated energy yield of a wood, low moisture biomass with a high moisture form. As can be seen from this example, the energy content of the high moisture material is essentially unusable in thermal processes. Howerver, energy balances discussed herein are defined in the former manner.

Table 1 BIOMASS ENERGY YIELD FROM FERMENTATION VS COMBUSTION

	Composition %				BTU/lb (DAF)	
Substrate	Cellulose	Lignin	H_2O	Ash	Fermentation	Combustion
Woody Biomass	50	25	20	5	$\frac{3750 + 0 + 0}{.75} = 5000$	$\frac{3750 + 3175 - 250}{.75} = 9900$
High Moisture Biomass (Short rotation plants, aquatic biomass, some agricultural residues.)	12	0	35	3	$\frac{900 + 0 + 0}{.12} = 7500$	$\frac{900 + 0 - 1100}{.12} = -1700$

As with other fermentation processes to produce liuid fuels from biomass, this process is anaerobic. Table 2 presents estimates of energy output from an aerobic cellulose fermentation versus an anaerobic one. The aerobic process is highly exothermic, the remaining heating value being ily in the cell mass produced. The anaerobic ----, on the other hand, is only mildly exother- :his is an important distinction since the ect of the process is to produce fuels, not low lity heat.

e 2 AEROBIC VS ANAEROBIC FERMENTATION OF CELLULOSE

Aerobic (~10% Cell Mass Production)

$$C_6H_{10}O_5 + 6O_2 \rightarrow 6CO_2 + 5H_2O \qquad -600 \text{ Kcals}$$

Anaerobic (~1% Cell Mass Production)

$$C_6H_{10}O_5 + H_2O \rightarrow 3CH_3COOH \qquad -38 \text{ Kcals}$$

ike other fermentation processes, however, this cess is conducted in a nonsterile manner. The t required to sterilize the substrate before mentation can be considerable. Table 3 presents imates of the energy required to autoclave high low moisture biomass. Particularly in the case the high moisture biomass, the energy required sterilization is a serious burden to the pro- s. Fermentation of biomass to liquid fuels via anic acids has other advantages over competitive cesses (e.g. ethanol fermentation) besides ng suitable for nonsterile operation.

Table 3 ENERGY REQUIREMENTS FOR STERILIZATION

Autoclave 50% Fermentables

Heat Requirement = 240 Btu/lb. Fermentables

Autoclave 10% Fermentables

Heat Requirement = 1200 Btu/lb Fermentables

Prehydrolysis of the substrate to soluble components is not required. Yeast fermentations to ethanol have either acid or enzymatic hydrolysis of the substrate as a prerequisite step; that is to say, these organisms do not produce extracellular celluloytic enzymes.

Pentose fractions as well as hexose fractions are converted to organic acids. Yeast fermentations convert only hexose sugars to ethanol. Since most forms of biomass contain hemi-cellulose as well as cellulose, this limits the potential yield of the process, and yield is an important consideration in the economics and energy balance of a process. Table 4 shows a cost breakdown of this process by components. From this table it is seen that at a biomass cost of $30/ton and a 75% conversion, the contribution to the biomass is approximately 50%. At lower yields, the contribution would be proportionately greater.

Organic acids, particularly the longer chain acids, may be removed from the fermentation by solvent extraction using solvents which are nontoxic to the microorganisms responsible for acid production. Ethanol is generally removed from fermentation broths by distillation, the most energy-intensive part of the process. Solvent extraction is a much less energy intensive process but we are unaware of any suitable extraction system for ethanol that is biocompatible.

Table 4 BREAKDOWN OF TOTAL HYDROCARBON PRODUCT COST BY COMPONENT

COMPONENT	COST, $/MMBTU
BIOMASS @ $30/TON D.A.F.	3.12
LABOR AND LABOR-RELATED COSTS	.29
CAPITAL AND CAPITAL-RELATED COSTS	2.01
ELECTRICAL POWER AND ROYALTY	.63
TAXES AND OTHER OPERATING COSTS	.37
TOTAL COST, $/MMBTU	6.42 @ $30/TON BIOMASS

The primary negative factor in the process under discussion is its requirement for electricity. Although the formation of hydrocarbons from organic acids is thermodynamically spontaneous, in practice, it has been observed that activation energy is required to produce the desired product. Table 5 presents the free energies and enthalpies of formation of hydrocarbons from organic acids. It is one of the primary goals of this research to reduce the voltage required as much as possible.

Table 5 ENTHALPY AND FREE ENERGY (WITH CORRESPONDING OXIDATION POTENTIALS) FOR ELECTROLYSIS OF ORGANIC ACIDS

KOLBE REACTION

$$RCO_2H \rightarrow \frac{1}{2} R\text{-}R + CO_2 + \frac{1}{2} H_2$$

	H		H	
ACID	KCALS/MOLE	VOLTS	KCALS/MOLE	VOLTS
ACETIC	9.0	-0.39	-4.4	0.19
PROPIONIC	10.8	-0.47	-4.6	0.20
BUTYRIC	4.8	-0.21	-5.1	0.22
VALERIC	3.9	-0.17	-5.8	0.25
CAPROIC	8.3	-0.36	-6.5	0.28

PROCESS DESIGN

Fermentation

In the process as described, the fermentation step is envisioned as occurring in a fixed packed bed fermenter. The mass transfer ordinarily accomplished by stirring in conventional digesters can be equally well fulfilled by a stream of liquid of the proper composition moving at the required rate through the fermenter. This same liquid stream may be used to control the temperature, add nutrients, control pH, and to remove reaction inhibiting products and toxic components. In addition, the piping arrangement involved may be employed to dewater the undigested material when the digestion state is completed.

Figure 5 shows one possible arrangement for the fixed packed bed fermenter. In this case it may be used as a batch reactor for organic acid production. Ther fermenter, A, is filled with solid substrate. An aqueous medium containing nutrients and microorganisms is added in sufficeint amounts to saturate the substrate and to fill the digester biomass, are countercurrent devices. Furthermore, the hydrocarbon or fluorocarbon phase must be augmented with or replaced by a substance with a higher partition coefficient for the lower molecular with acids such as trioctylphosphine oxide (2).

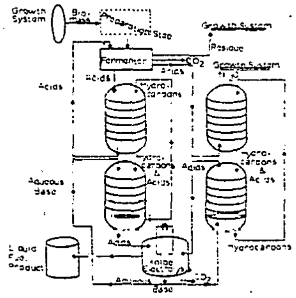

Fig. 8

Electrolysis

Although the mechanism of Kolbe electrolysis has been a source of disagreement for many years, one suggested reaction scheme, presented in Fig. 9, (3), is suitable for current purposes. Basically the organic acid anions are oxidized to the corresponding radicals. These radicals decarboxylate

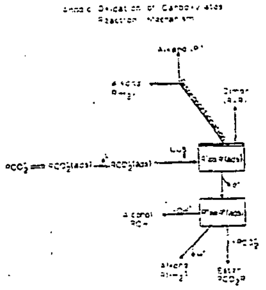

Fig. 9 Anodic Oxidation of Carboxylates Reaction Mechanism

to form alkyl radicals which may either dimerize form the Kolbe product, disproportionate to form an alkene-alkane mixture, or may be oxidized further to form alkene, alcohol, or ester. Recent experiments involving the electrolytic oxidation of mixed organic acids produced by the anaerobic fermentation of Chondrus crispus gave the product spectrum shown in Figs. 10 and 11. Figure 10 give

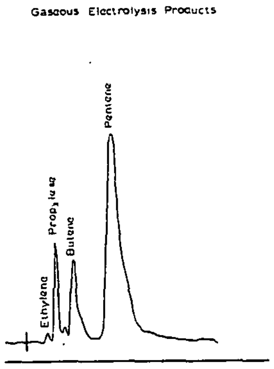

Fig. 10 Liquid Electrolysis Products

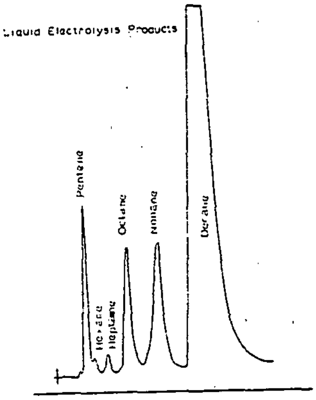

Fig. 11 Gaseous Electrolysis Products

the gas chromatogram of the liquid products, and Fig. 11 shows the gaseous products. The gaseous products are labeled as alkenes but the correspor ding alkanes would not be resolvable under th analysis conditions. This particular experi

run at pH 6, an applied voltage of 4.65V and
dic current density of 0.8 amps/cm^2. The
lysis was performed in the flowthrough cell
n Fig. 12. A flowthrough cell configura-
used to minimize polarization effects at
h the anode and cathode. It consists of a pla-
um wire anode inside a platinum tube. The elec-
de spacing is 300 μm. An additional platinum
e electrode is used to measure the open circuit
rating voltage while the electrolysis is taking
ce. The difference between the operating vol-
e and the open circuit voltage is a measure
the IR drop in the cell and other current den-
y dependent overpotentials.

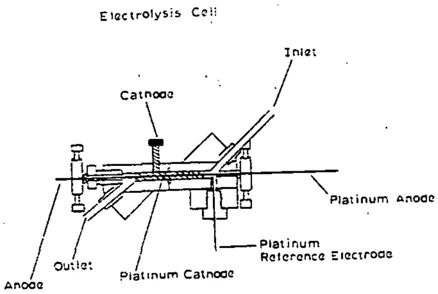

Fig. 12 Electrolysis Cell

se current density-dependent potentials give
e to an optimum current density. Figure 13
ws a plot of product cost versus current density
h a minimum at approximately 0.3 amps/cm^2. Low-
current densities have higher prorated capital
ts whereas higher current densitites give rise
excessive electrical requirements.

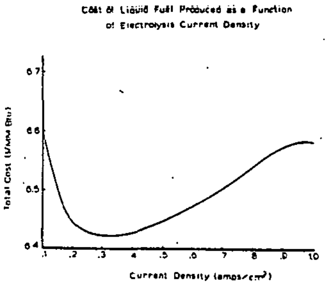

Fig. 13 Cost of Liquid Fuel Produced as a Function of Electrolysis Current Density

rgetics are critically dependent on the effici-
of the electrolysis process. Figure 14 presents the cost of liquid fuel produced as a function of applied voltage. A voltage of greater than 4V may be used before a cost of $7.00/MM Btu is reached. The slope change at approximately 1.5V occurs because at this potential sufficient resistance heating in the electrolysis is reached to supply all of the process heat requirements.

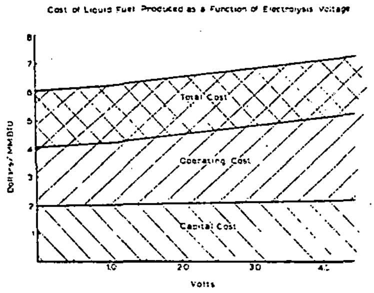

Fig. 14 Cost of Liquid Fuel Produced as a Function of Electrolysis

APPENDIX

Geoffrey Barnard and Sara Molyneaux of Dynatech performed most of the experimental and construction work, and the contribution of Nancy Scarpace was invaluable to the manuscript preparation. Dr. John H. Ryther, Woods Hole Oceanographic Institution, Woods Hole, Massachusetts, provided most of the aquatic biomass used in this work except for *Chondrus crispus* which was obtained from Marine Colloids, Inc., Rockland, Maine. We are indebted to Professor J.G. Zeikus, University of Wisconsin, Madison, for bringing bromoethanesolfonic acid to our attention.

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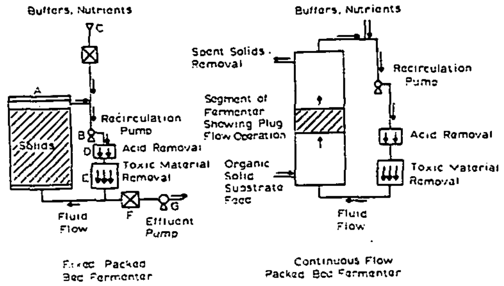

Fig. 5

head space and circulation loopa as shown. The pump, B, is used to circulate the liquid through the the loop and through the packed mass of comminuted solid substrate containing fermentation microorganisms. As fermentation proceeds, required buffers and additional nutrients may be added through the inlet port, C; the acid is removed in sub-system D, which may be an ion exchange column, liquid-liquid extractor, or membrane device. Toxic compounds may also be removed from the aqueous phase by passing it through a cleaner, E, which may contain, for example, activated charcoal. When fermentation is economically completed, the undigested material is partially dewatered by opening valve F and pumping liquid out by means of pump G. The partially dewatered material is then removed, fresh substrate added, and the cycle above repeated.

Application of the fixed packed bed concept to a continuous plug flow system is presented in Fig. 5. Here solid substrate is fed into the bottom of the batch-type packed bed system, fermenter liquid is circulated through the bed and organic acids are recovered in an external sub-system.

Currently in operation is a 300 liter fixed packed bed fermenter. It is constructed of 12-inch diameter glass tubing and is 15 feet in height. It is supported by a triangular framework of a 1 1/4-inch pipe and is enclosed with plywood and plexiglass. The temperature is controlled at 100°F by circulating warm air inside the enclosure. A flow rate of 3 gallons per hour is sufficient to fluidize the substrate, initially at 10% solids, in the reactor to reduce the operating pressure at the pump (approximately 7 feet below the top of the column) to 3 PSIG.

Figure 6 is a schematic diagram of the type of device designed to add substrate and reserve solid residue in a continuous manner. The primary element is an auger with a tapered shaft that will separate liquid and solid fractions by squeezing. A bench-model feeding device has recently been constructed. It consists of a 4-inch earth auger built up to the proper taper with 1/8-inch rubber strips encased in a plexiglass structure. Testing of this device is in progress prior to converting the 300 liter fermenter to continuous operation.

As was stated earlier, the overall product cost is

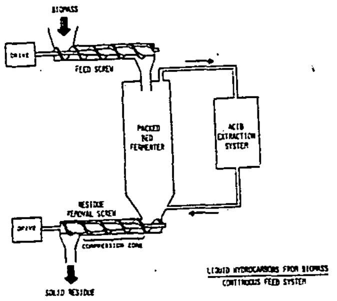

Fig. 6 Liquid Hydrocarbons from Biomass Continuous Feed System

very sensitive to conversion yield. This fact is illustrated more fully in Fig. 7, which is a plot of product cost versus retention time and indicates that the added cost due to a larger fermenter volume is outweighed by the reduction in conversion yield only after about a 60 day retention time, which corresponds to 83% of the fermentable material converted.

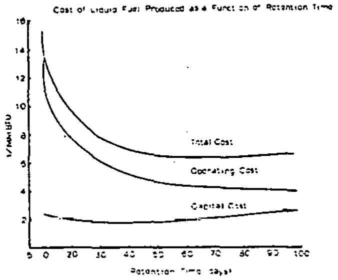

Fig. 7 Cost of Liquid Fuel Produced as a Function of Retention Time

Extraction

Removal of product acids is accomplished by liqui liquid extraction using preferably a hydrocarbon or alternatively a flurocarbon solvent. Figure 8 shows the extractor set-up for a large-scale operation. The primary extractors, which remove the higher acids from the fermenter effluent and transfer them to the aqueous stream which is then fed to the electrolysis cell. Since these extrac tors are designed to remove a small portion of t acids from a large volume of fermentor effluent, they are concurrent units. The secondary extractors, which are designed to remove most of the acids from that portion of the fermentor effluent which is returned to the growth system as it i displaced by the water introduced with the fr

Session II C

AN OVERVIEW OF BIOMASS LIQUEFACTION

Sabri Ergun
Lawrence Berkeley Laboratory
University of California
Berkeley, California

STRACT

e construction of a process development unit DU) at Albany, Oregon for converting biomass to l was completed in December of 1976. The design the facility was based on a process conceptual- ed by the Bureau of Mines in the early 1970s. e PDU was sized to produce oil at a rate of out five barrels per day.

is presentation deals with the background of quefaction efforts and progress made at the bany facility from January 1977 through Septem- r 1978.

ERVIEW

nversion of biomass into oil or gas is not a new ea, nor is the recognition that synfuels derived om wood wastes or wood grown for such a purpose uld meet a substantial portion of America's needs energy and chemicals. I have not researched cently how far back the idea dates. Suffice it state that there was a genuine interest in the rly 1940's for fuels from wood, which was lost on the discovery of huge oil fields in the abian peninsula soon after.

e interest in synthetic fuels from biomass was vived some thirty years later for reasons that ed no elaboration. The current biomass lique- ction program of DOE has evolved from the lig- te liquefaction program of the Bureau of Mines the late 1960s and early 1970s. Noting the despread interest in municipal waste disposal, e Bureau of Mines researchers explored the fectiveness of a process particularly well ited for lignite to liquefy organic municipal ste. They found that organic municipal waste d other biomass wastes such as sawdust and manure, quefied with greater ease under milder conditions an did lignite. Based on their research they nceived a process for wood liquefaction. In the ocess, 30 parts of dry wood flour is blended with parts of a vehicle oil, and 7.5 parts of 20% ueous solution of sodium carbonate, then heated about 700°F under carbon monoxide or synthesis s pressure to a final total pressure of about 500 ± 1,200 psi and for a total residence time nging from 45 to 60 minutes. If completely con- rted, 30 parts of wood would yield about 17.4 rts of oil (58%) having a heating value of about ,500 Btu/lb.

couraged by the results of bench-scale research the economic feasibility analysis provided by the Blau Knox Chemicals Division of the Dravo Corporation, Bureau of Mines proceeded with the design and construction of a process development unit (PDU) capable of processing 3 tons of wood daily or to yield roughly 5 bbl oil daily. The plant site chosen was adjacent to the Experimental Metallurgical Station of the Bureau of Mines at Albany, Oregon. The Albany PDU is illustrated schematically in Figure 1. Wood chips are first dried, milled to produce a -40 mesh flour, blended with a recycle oil, and injected into the reactor assembly comprised of a scraped surface preheater and stirred tank reactor along with sodium carbonate solution and synthesis gas. The exit stream from the reactor is cooled and sent to a pressure letdown vessel. The flashed liquid goes into a three-phase centrifuge to free the oil from unreacted solid residue and ash as a sludge and from an aqueous phase containing sodium salts. The clean oil is sent back to the blender after a slipstream (about 20%) is withdrawn as the product. Initially the slip stream is largely composed of the start-up oil, e.g., anthracene. As the run progresses, the start-up oil will be depleted and eventually replaced by the oil derived from wood. It may take about a week or more to obtain an oil almost free from start-up oil.

The plant construction was completed in December of 1976. However, by mid-1976, the facility was transferred to the Solar Energy Division of ERDA, now DOE, due to a reorganization at Washington. ERDA contracted with Bechtel Corporation for the acceptance of the facility, to commission the facility, perform shake-down tests and conduct test runs. The original contract period was one year, which later was extended to about 18 months. Plant commissioning and shakedown did not prove to be easy. Although small, the facility is extremely complex and contains process units operating under conditions heretofore unencountered. In addition, the plant was conceptualized by one company (Blau Knox), detailed engineering design was provided by another (Rust Engineering), and was constructed by a third company (Maecon, Inc.). Since Pittsburgh Energy Research Center turned its attention to lignite liquefaction upon transferral of responsibility, efforts involved in the various stages of plant design and construction could not adequately be coordinated. Bechtel did succeed in commissioning the plant and in conducting a limited number of test runs. However, a major process unit (i.e., the stirred tank reactor) could not be operated properly due to a design deficiency, and another one, the centrifuge, had to be bypassed due to operational difficulties. Therefore, the results obtained by Bechtel did not shed much light on the chemical, technical, and economic feasibility of the process. The contract in response to the ensuing RFP was awarded to Rust Engineering Company, which earlier had provided the detailed engineering design of the facility. A representative of Rust Engineering Company will present to you the results of their activities in this session.

Recognizing that further investigation of the process chemistry on bench scale would be beneficial to the development program, as Bureau of Mines had contemplated, Pacific Northwest Laboratories (PNL) proposed to ERDA that PNL would undertake bench-scale supporting activity. After conducting some experiments under conditions described by the Bureau of Mines Researchers for purposes of comparison, PNL widened the scope of investigation to conditions not covered before, provided analytical support to Bechtel, recommended the necessity of an on-site analytical facility at Albany, and conducted bench-scale tests under conditions suggested by Rust. In this session, PNL will present the results of their recent activity.

The University of Arizona noted that the injection of solids into high pressure vessels was a formidable unit operation faced in coal liquefaction as well as biomass liquefaction and, accordingly, proposed to undertake research to develop extrusion techniques for biomass injection into high pressure vessels. Professor Don White will make a presentation in this session on the subject of biomass extrusion into high pressure vessels.

In October 1977, DOE contracted with LBL to monitor the development program at Albany and the supporting activity to provide a third-party objectivity in the evaluation of results, to make recommendations regarding the future course of the program, and to provide assistance and appraisal to the new operators of the facility should a different company become a successful bidder to the RFP to be issued, i.e., to assure continuity. The results of the LBL efforts will be presented next.

The causes of the operational difficulties encountered at Albany remained controversial. Fluctuations in the performances of the various equipment, e.g., pumps, feeders, flash tank, etc., collectively could abort a run by causing a breakdown elsewhere. The system was operated in various bypass modes, first the reactor was bypassed, then both the reactor and centrifuge were bypassed, and finally only the centrifuge was bypassed. In other words, the system was not operating the way it was designed to. On the other hand, the product obtained could have been a material solubilized in anthracene such that when anthracene concentration reached a level below 50%, it simply plugged the system. DOE concluded that the operational difficulties encountered at Albany, Oregon proved to be a formidable obstacle to assessment of the technical feasibility of the process and that the chemical feasibility of the process remained in question. Accordingly, in FY79, LBL was given the responsibility of providing the operating directives for the test runs to be conducted at Albany specifying the conditions of operation (e.g., temperature pressure, wood/oil ratio, wood/catalyst ratio, wood/syngas ratio, slurry feed rates, residence time, etc.) to assess the chemical feasibility of the process. If one considers the fact that any operating condition that is deemed to be desirable from chemical or kinetic considerations, must also be technically feasible (about which the engineers in charge of the operation of the facility will have something to say), one appreciates the checks and balances aspect of the DOE approach.

LBL was also given the responsibility of monitoring the supporting research activity on biomass liquefaction to utilize the results in the development program at Albany. Also, LBL proposed to DOE to conduct supporting engineering and development studies to provide optional approaches to biomass liquefaction. The results of the LBL studies conducted in FY79 will be the last paper presented in this session.

D.O.E. CONTRACT ET-78-C-06-1092
OPERATIONS OF THE BIOMASS LIQUEFACTION FACILITY
ALBANY, OREGON
WHEELABRATOR CLEANFUEL CORPORATION
JULY 1, 1978-JUNE 30, 1980

WALTER L. BERRY, JR.
THE RUST ENGINEERING COMPANY
P.O. BOX 101
BIRMINGHAM, ALABAMA 35201

BSTRACT

he objective of this contract is to obtain definiive process information sufficient for commercial esign and economic evaluation purposes for the onversion of biomass to syncrude. This objective s to be accomplished through the execution of xperimental and developmental programs in the Govrnment-owned process development unit (PDU) loated adjacent to the Bureau of Mines Metallurgical esearch Center at Albany, Oregon.

peration of the plant is as directed by the Govrnment and consists of the following phases of ork:

hase I: Preparatory Work, including an Environmental, Health, and Safety Assessment

hase II: Operating Adjustment Period

hase III: Continuation of Process Investigation

hase IV: Modification of Plant Configuration and/or Process Flowsheets

hase V: Process Development Operations

hase VI: Shutdown and Phaseout

hases I and II have been completed. Phase III is nderway. Test runs utilizing anthracene as the arrier vehicle have been underway since October ith product oil concentrations of 38% produced. esting continues to improve equipment reliability nd to separate product oil from carrier vehicle.

dditional equipment is being added to the system o improve performance. The first test run incororating pretreatment of the wood with acid and lkaline hydrolysis, utilizing water as the carrier ehicle has been successfully completed. Addition1 quantitative test runs are planned for this conept.

hases IV, V, and VI will be accomplished when suficient data has been accumulated.

BACKGROUND

Charcoal is produced by burning wood in an oxygen-deficient environment, and some tar is simultaneously produced, thus we can say that biomass liquefaction has been known for a long time. Early in this century, Bergius studied thermal decomposition and hydrogenation of cellulose, producing small quantities of oil which resembled petroleum.

In the early 1930's, hydrogenation of sawdust and cotton in a tetralin medium produced a liquid product. This work was reported in the Canadian Journal of Research. Various other investigators performed related studies about that same time.

The Department of Energy and its predecessor organization, the Energy Research Development Administration (ERDA), and the Pittsburgh Energy Research Center (PERC), which is part of the United States Bureau of Mines (USBM), early recognized the importance of developing alternate fuels and energy systems. Thus began serious efforts to convert biomass to a liquid fuel. Laboratory research was carried out at PERC beginning in the late 1960's. These bench-scale investigations, carried on into the early 1970's, soon showed the potential for producing a synthetic crude oil from cellulosic waste which included such materials as wood, urban refuse, agricultural wastes (corn cobs, manure) and other organic wastes.

With the technical possibility established, a preliminary economic feasibility study of the process was performed.

Following the encouraging results of this study, the USBM commissioned The Rust Engineering Company (Rust) to design a Process Development Unit (PDU) to process three tons per day of wood chips. This PDU was to be constructed using commercially available components.

Further studies confirmed the desirability of testing the process in the PDU. Under contract to the USBM and its successor organization, ERDA, the PDU was constructed in 1974-75 by Maecon, Inc., in Albany, Oregon at the Bureau of Mines Metallurgical Research Center. Rust had no role in either final equipment selection or construction supervision during this phase of work.

From January 1977 to June 1978, the Albany PDU was operated by Bechtel Corporation, pursuant to the overall objective of DOE's program to obtain definitive process information to permit commercial design of a plant for conversion of biomass to synthetic crude oil.

As is often the case with PDU's, there have been numerous mechanical and operating problems, but the process has shown sufficient promise for DOE to continue the operations for an additional two-year period beginning July 1, 1978. Wheelabrator Cleanfuel Corporation, with Rust as its prime subcontractor, was awarded the contract for this period.

The significance of the project, and the considerable interest that has been shown in the PDU in the recent past, is readily understood when it is considered that the nation generates three billion tons of solid organic waste annually. If only half of this waste were processed in liquefaction plants, approximately one and one-half billion barrels of oil could be produced.

Solar energy advocates would naturally consider the possibility of cultivating fast-growing crops to supply the biomass required as feedstock for a large commercial plant of, say, 5,000 or more tons per day.

The process can be represented by two basic equations which involve reacting carbon monoxide and hydrogen with cellulosic materials at temperatures up to 700°F and at pressures up to 4,000 psig in the presence of sodium carbonate as a catalyst. Carbon monoxide reacts with sodium carbonate in the presence of water to form sodium formate which, in turn, reacts with cellulose to form oil and regenerate sodium carbonate. The following reactions are believed to be typical:

$$(1)\qquad Na_2CO_3 + H_2 + 2CO \rightarrow 2HCOONa + CO_2$$

$$(2)\qquad 2C_6H_{10}O_5 + 2HCOONa \rightarrow 2C_6H_{10}O_4 + H_2O + CO_2 + Na_2CO_3$$

The oil resulting from the process is low in sulfur, suitable for use in power plants, or as feedstock for conversion to gasoline and diesel fuels. Laboratory-produced samples show the following properties:

Specific Gravity		1.1
Viscosity, CP (140°F)		515
Heating Value, BTU per pound		15,000
Composition, %	Carbon	76.62
	Hydrogen	7.05
	Nitrogen	0.13
	Sulfur	0.14
	Oxygen	20.05

Potential applications as process feedstocks are possibly even more important economically than fuel uses.

Carrying out the apparently simple deoxygenation reaction in the PDU, however, involves several unit operations functioning in somewhat difficult conditions because of constraints such as temperature, pressure, small size and the requirement for co mercial availability of all items of equipment. its simplest terms, the primary operating mode sists of:

- * Drying the biomass (wood in the PDU at present) to less than 4% moisture.
- * Grinding to a -35 mesh flour.
- * Slurrying the wood flour with carrier vehicle oil.
- * Pressurizing and preheating.
- * Injection into the heated reactor at pressures up to 4,000 psig and temperatures up to 700°F.
- * Injection of catalyst solution into the slurry of wood and oil.
- * Injection of carbon monoxide and hydrogen into the slurry of wood and oil.
- * Pressure letdown.
- * Separation of solids (catalyst and unreacted wood) from the oil.

In the most recent operations, both catalyst solution and reactant gases have been injected ahead c the Preheater.

PROCESS AND EQUIPMENT DETAILS

The biomass feedstock in present use is Douglas F wood chips purchased from a nearby paper mill. These chips are stored in an open shed. The chip are moved by a front end loader to the Feed Conve or for transfer to the Wood Storage Bin. The woo chips are withdrawn from the storage bin by a controlled-rate table feeder and fed into a gas-fire rotary dryer where their moisture content is redu from about 50% to a maximum of 4%, using hot comb tion products for heating. The dried chips are then fed to a hammer mill for size reduction, and the resulting flour is screened to be sure it pas through a 35-mesh screen.

The dried wood, as well as the grinding and subsequent wood processing operations, are kept under nitrogen atmosphere to minimize moisture pick-up and explosion hazard.

The wood flour is conveyed pneumatically to an el vated Wood Flour Storage Bin in a closed system which uses nitrogen as the conveying medium.

The catalyst, sodium carbonate, is prepared as a 10-20% aqueous solution for injection into the reaction system by means of a conventional high pre sure metering pump.

Carbon monoxide and hydrogen are received as compressed gases in standard over-the-road tube tra ers. A carbon monoxide-hydrogen mixture is compressed to reaction pressure in a non-lubricated, non-contaminating diaphragm-type compressor. The compressed mixture may be fed into the system con tinuously either at the Preheater or the Reactor.

Wood flour is continuously fed from the Wood Flou Storage Bin through a Weigh-belt Feeder to the Wo Oil Blender.

thracene oil is used as a carrier vehicle for
art-up in the PERC process.

hracene is pumped to the Wood-Oil Blender, where
is mixed with wood flour in a cone-shaped blend-
which is equipped with helical screws that sweep
e side walls.

ter start-up, product oil serves as the carrier
hicle.

e wood-oil slurry from the Blender, at 10-20%
od content, is pumped into the reaction system
rough a Feed Slurry Circulating Pump and a High
essure Slurry Feed Pump; thence through an elec-
ically heated, scraped-surface feed Preheater to
electrically heated Reactor, where temperature
d pressure are maintained and reaction occurs.

talyst solution and the carbon monoxide-hydrogen
xture are injected ahead of the Preheater. The
gh-pressure, high-temperature equipment is con-
ructed of Type 316 stainless steel. The Feed
urry Circulating Pump is a Tuthill positive-dis-
acement lobe-type pump, while the High Pressure
urry Feed Pump is a Bran and Lubbe variable-
roke type equipped with ball check valves.

e Reactor effluent is cooled in an air-cooled heat
changer prior to pressure letdown. Water and
ght oils flash and are separated in a low pres-
re flash system. The vented gas is cooled for
covery of light oils and water. Vented gas and
n-condensibles are incinerated.

the PDU was originally designed, a centrifuge
s installed to separate any sludge of catalyst
d/or unreacted wood from the product oil since
nch-scale batch experiments had shown this separ-
ion possible. To date, no product has been suc-
ssfully clarified by the centrifuge because of
e high viscosity of the oil mixture.

e process, as we have just reviewed, feeds a
urry of 10-20% biomass in vehicle oil. This
ans that approximately 80% of the circulating
ad acts only as an inert carrier, and pumping,
ating and cooling this carrier results in the
ste of considerable energy which could lead to a
nsiderable increase in plant size (and cost) for
given throughput of biomass. Recognizing this
ct, two additional methods for solids injection
re built into the PDU:

A system for feeding pretreated wood; and
A system for feeding dry wood flour directly to
the Reactor.

e first alternative system for pretreating the
ed involves a batchwise cooking of the wood chips
water in an electrically heated, agitated auto-
ave. Original PERC laboratory work indicated a
quirement of exposing the wood to a temperature
500°F and a pressure of 680 psia for one hour.

e resulting slurry is cooled in an external pump-
ound circuit through an air-cooled heat exchanger.
e cooked wood is separated from the aqueous waste
ns of a small continuous vacuum filter. The

wood solids are somewhat similar to charcoal in appearance and are more easily processed than the dried chips. The solids are dried and ground and then processed as previously discussed. The pretreated wood offers the advantage of making a slurry with improved rheological properties such that a 50% slurry in anthracene oil can be pumped. In addition, there may be enhancement of the chemical reactions.

Recent bench-scale experiments conducted by the LBL group point to using these autoclaves for hydrolysis of the wood at conditions of 356°F and 130 psig for 45 minutes, using sulfuric acid for pH control.

In the second alternative system, solids go directly into the Reactor at pressure and temperature.

For this part of the PDU, a mechanical feed arrangement incorporating a rotary feeder was designed. It basically consists of two pressure-balanced Lock Hoppers which will be used alternately. To be able to use conventional rotary feeders, the Lock Hoppers were designed so that the feeders are installed inside and, therefore, are not exposed to high differential pressure. A rotary feeder capable of withstanding a 4000-psig differential would have been preferred for better access, improved maintenance, and a simpler lock-hopper design; but the equipment vendors expressed little interest in building such a feeder.

The Lock Hopper system functions in the following manner. A Lock Hopper is filled with wood flour and pressurized to 4000 psig, using a carbon monoxide-hydrogen mixture. The solids are then fed to the Reactor by the rotary feeder. When the Lock Hopper is empty, the pressure is reduced, venting the pressurizing gas first to a surge tank, then passing the remaining gas to the incinerator. Once the Lock Hopper is vented, it is ready for the next cycle. The surge tank collects most of the carbon monoxide mixture from the Lock Hoppers, so that the gas can be recompressed and reused in the next cycle.

Heat is supplied by recycle of part of the product oil through the Preheater as in the previous operational mode. We would expect high slurry concentration with this feed method.

Work now in progress at the University of Arizona may eventually permit feeding a very high slurry concentration, thereby providing a more attractive alternative.

PROBLEMS AND DIFFICULTIES

Practically all the reaction product that was produced during the previous contract period was extremely viscous at ambient temperatures and virtually umpumpable even at elevated temperatures.

Early Rust operations were hampered by inadequate heating of electrically traced piping. This caused premature shutdowns in some of our earlier operations but installation of proper high temperature tracing has eliminated this problem.

During a recent run, numerous leaks developed in high pressure welded fittings and screwed connections; these required shutdown for repair.

The wood processing equipment has generally functioned well.

Moyno pumps used for oil-wood slurry circulation were originally supplied with Buna stators which had poor resistance to the anthracene oil used as start-up vehicle. Replacement with Viton stators showed improvement.

Replacement of the Moynos with Tuthill lobe-type, all metal positive displacement pumps was eventually necessary, however, because of temperature limitations of the Viton.

Numerous failures of polymeric gaskets, rings and seal parts were experienced; these required replacement by such materials as Teflon, graphite, metallic parts, etc.

High Pressure Slurry Feed Pump check valves had short life until the stainless steel valve seats were replaced with Stellite. Packing life has been short, but is improving with changes in material.

Serious mechanical deficiencies with the scraped surface Preheater have existed. The mechanical shaft seal was modified extensively to provide a double seal with intermediate pressure between the seals. The original scraper blades were mechanically inadequate and required frequent replacement. Spring-loaded scraper blades, supplied by Armstrong Engineering Associates, have been installed and are expected to provide better heat transfer and improved service life.

As supplied, the Reactor head-to-shell closure was not adequate for the cyclic operations which are common in a new process pilot plant. Replacement with a GrayLoc closure eliminated this problem.

One major problem that has existed in all runs to date concerns the Reactor agitator shaft seal. Instead of the packless, magnetic drive specified, a conventional double seal with elastomeric parts was installed. This seal has been inadequate. The previous contractor removed the agitator, and when the Reactor was used, its contents were not agitated. A magnetically driven, packless agitator is being installed and will be placed in service in the immediate future.

Reactor product pressure letdown valves have had short life. This is not surprising when the extremely severe one-step drop from 4000 psig to 7 psig is considered. The small valve required for low capacity throughput and high pressure differential compound the problem. Parallel valves were installed to permit replacement without process interruption. Various valve trim materials have been used, and Stellite has been the most successful to date.

Several failures of the carbon monoxide compressor diaphragms occurred. Head modification by the manufacturer has partially solved this problem.

Because it has not been possible to seal the Re
tor agitator shaft, no effort has been made to
operate in the Lock Hopper feed mode. Use of t
equipment is scheduled later in the program.

As previously mentioned, problems with product vis
cosity have precluded use of the centrifuge to dat
However, recent results have been encouraging, inasmuch as lower viscosities have been observed, ar
these lower viscosities should lend themselves to
utilization of the existing centrifuge. Tests of
the centrifuge system are now underway.

No separation of product oil from the anthracene
oil used as start-up vehicle has been accomplishec
to date. In sustained operation, the anthracene
would disappear by dilution. However, distillatic
tests recently performed in our laboratory indicat
the potential of both upgrading product oil and re
moving catalyst by this means.

CONCLUSION

Two promising systems for producing syncrude from
wood have been demonstrated to date in the PDU.
Anthracene as a carrier vehicle has been used to
produce an oil concentration of 38%. Syncrude ha
also been produced by pretreating and hydrolizing
wood in slightly acidic water, and further reacti
in the presence of a CO and H_2 gaseous mixture an
sodium carbonate catalyst. Further testing is
underway to quantify process variables and assemb
data for commercial design and economic evaluatio
purposes.

DIRECT LIQUEFACTION OF BIOMASS - CORRELATIVE ASSESSMENT OF PROCESS DEVELOPMENT

Carlos Figueroa and Sabri Ergun
Lawrence Berkeley Laboratory
University of California
Berkeley, California

Abstract

wrence Berkeley Laboratory was assigned by DOE
e responsibility of directing and technical
nitoring of the projects dealing with direct
qeufaction of biomass. LBL found it desirable
fulfill its responsibility by undertaking a
ntinual correlative assessment of process devel-
ment activities and by initiating and under-
king studies to fill the msssing gaps in order
bring the biomass liquefaction program to a
ccessful conclusion speedily. This presentation
als with the correlative assessment efforts of
L in the development of an oil-from-biomass
chnology.

RRELATIVE ASSESSMENT OF PROCESS DEVELOPMENT

e wood-to-oil process development unit (PDU)
cated at Albany, Oregon was designed on the basis
bench scale batchwise experiments conducted
the Bureau of Mines. The translation of the
sults of batchwise experiments into a continuous
it required a lot of guesswork. Major process
its such as blender, preheater, pressure letdown
ssel, and centrifuge were designed on the basis
meager or no data. Successful operation of
ese units and their modifications required engi-
ering R & D before they could be operated
ccessfully.

equal concern was the fact that the fate of
e biomass liquefaction program depended upon
e fate of a single process conceptualized by
e Bureau of Mines. The Albany PDU was primarily
signed to evaluate the technical feasibility of
e Bureau of Mines process and to gather suffi-
ent data in order to assess the economic feasi-
lity of the process and to provide a data base
r the design of a demonstration unit. If was
citly assumed that the chemical feasibility
the process was a certainty. Preliminary
periments conducted at Albany cast some doubts
the chemical feasibility of the process.
cordingly process modifications and/or new pro-
ss options had to be researched and a data base
ovided for testing at Albany.

wrence Berkeley Laboratory (LBL) was given the
sponsibility of directing and technical monitory
the projects dealing with direct liquefaction
biomass. In discharging this responsibility
L interfaced with the Biomass Energy Systems
anch of DOE, the operators of the Albany, Oregon
U and with institutions providing supporting re-
[illegible] and other services. Interfacing with the operators of the PDU required undertaking complementary studies, providing engineering support, providing input from supporting basic research, and undertaking any other activity to speed the development program. This presentation deals with the correlative assessment activity undertaken by LBL.

The results of Phase I operations (modify, commission, and conduct preliminary experiments) more or less identified the areas of concern. They pertained to:

- Mechanical problems
- Unit operations
- Chemistry of the process(es) and physical chemical aspects,
- Multiplicity of the variable parameters.

Mechanical problems proved to be formidable; they were largely handled by the operators of the facility. The unit operational problems encountered were both of mechanical and chemical engineering nature. In this respect LBL interfaced with Rust Engineering Company. For example, our mechanical engineer provided the design for the modification of the wood flour feeding system and worked with the Rust staff in the analysis of problems encountered with the preheater, pressure letdown vessel and thermal stresses, etc. Recognizing that the design of major process units such as preheater, pressure letdown system, and centrifuge was based on meager data or guess work, LBL developed procedures for the evaluation of the performances of the various process units. We may cite the centrifuge as an example. The centrifuge of the Albany PDU, a three phase unit, did not work and was removed from the system. This centrifuge was supposed to separate the solid residue as a sludge, and an aqueous phase containing the catalyst from the flash tank bottoms to provide a clear oil for recycling and as a product. A three phase centrifuge is as complicated a unit as a fractionation column. An LBL analysis indicated that the centrifuge chosen is not likely to separate an aqueous phase; it may separate the oil formed in the process from the startup oil if the properties of the two oils are very different and eventually become inoperative or useless as the startup oil becomes depleted.

As Dr. Larry Shaleger informed you in his presentation, LBL developed a modified process option that requires pretreatment of biomass. Although two large stirred autoclaves were available at Albany for wood pretreatment, they were not designed for the operational procedures developed at LBL. Accordingly the LBL and Rust engineers worked together to modify the autoclaves available at Albany and to develop operational procedures that would meet the reaction conditions imposed by the chemistry of the pretreatment.

Dr. Manu Seth informed you of the results of LBL's investigations on the chemistry of the liquefaction of biomass. However promising, any new process or modified process option requires scrutiny regarding the adequacy of the data base provided for evaluation at Albany. The bench scale results have to be translated into the operation of a continuous unit such as exists at Albany, Oregon with minimum modifications, as major modifications are time consuming. The translation involves the specification of the following:

- o Preliminary modifications of the PDU, e.g., piping and instrumentation.
- o Startup procedures.
- o Operational procedures
- o Likely deviations from the anticipated procedures.
- o Operating conditions.
- o Data to be recorded and its frequency.
- o Equations for the analysis of the data recorded.

It is needless to elaborate that in order to develop valid equations for the analysis of the data recorded, one must consider heat and mass balances and kinetics of the reactions. The last factor is largely unknown to begin with, and in a strict sense, probably will remain unknown whether or not a viable process evolves from the effort. However, effective kinetic parameters must be formulated, realizing what can or cannot be measured or analized on the basis of their sensitivity to the operating conditions imposed. Process optimization boils down to finding the operating conditions that render the process most attractive economically. In the operating directives issued for testing the feasibilities of the Bureau of Mines process and a modified version of the Bureau of Mines process, the points raised above have been considered in detail.

The last concern that we cited in this presentation, i.e., the number of variable parameters, are shared by LBL and Rust Engineering Company. To be specific, they are listed below:

VEHICLE OIL/BIOMASS RATIO
CATALYST/BIOMASS RATIO
SYNTHESIS GAS/BIOMASS RATIO
SYNTHESIS GAS COMPOSITION
SLURRY FEED RATE (RESIDENCE TIME IN PREHEATER)
PREHEATER EXIT TEMPERATURE
PRESSURE
REACTOR TEMPERATURE
REACTOR INVENTORY (RESIDENCE TIME IN REACTOR)
TYPE OF BIOMASS
TYPE OF STARTUP OIL
TYPE OF CATALYST

The experience at Albany has shown that the number of test runs that can be conducted is limited, i.e. about one run a month. About twelve runs that can be conducted between now and August of 1980 will not be sufficient to evaluate the influences of twelve variables. Considering the chemistry of the reactions taking place, the unit operations involved in the process, and the confidence desired, we believe about 60 test runs have to be conducted. For this reason LBL has designed and constructed a mini PDU that in many respects simulates the Albany PDU. In the second part of his presentation, Dr. Seth has described two additional pieces of equipment designed to simulate continuous operations by batchwise experiments. Our plans call for screening the influences of as many parameters as possible in order to zero in on the conditions of the critical test runs that can be conducted at Albany. Of course we plan to establish a one-to-one correspondence between the results that can be obtained at Berkeley and Albany. For wood hydrolysis, for example, we started the tests in a 400 ml autoclave and graduated to one liter and later to a 10 gallon autoclave. Having established an identity in the results obtained, we requested that a test run be conducted at Albany using the 400 gallon autoclaves of the PDU. We are happy to report that the results obtained at Albany were identica to those obtained at Berkeley.

In summary, LBL found it highly desirable to fulfill its responsibility in the DOE's direct liquefaction of biomass program by undertaking a continual correlative assessment of process development activities and by initiating and undertaking studies to fill the missing gaps to speed the course of the program.

MODIFIED EXTRUDER FOR FEEDING CELLULOSIC SLURRIES TO PRESSURE SYSTEMS
Contract No: EW-78-S-05-5679

Don H. White, Department of Chemical Engineering
University of Arizona, Tucson, Az 85721, 602/626-5280

STRACT

e objective is to evaluate through experimental sts whether a modified screw extruder (as used the food and plastics industries) could be used feed a finely-ground slurry (ground sawdust in avy oils or water) into a pressure reactor sysm. This is of direct interest for the DOE ste-to-Oil pilot plant, Albany, Oregon and the E-sponsored cellulosics gasification work at llston Spa, New York. This basic overall goal s achieved during the first year of the contract 1978, utilizing Albany, Oregon process facility wdust, giving results of, (a) pumped feed slures in extruder developing up to 8,500 psi presre head, whereas 3,000 psi needed, and (b) mped up to 70 weight percent feed slurries, ereas pilot facility piston pumps are limited to -30 weight percent sawdust. During 1979, scale factors are being determined, which involve truder design, extruder operation conditions, operties of sawdust feed and properties of the cessary carrier oil. The work to date has en successful using single screw extruders. ter in 1979 or 1980, field tests will be conduct- upon several alternative types of extruders.

NTRACT OBJECTIVE

e objective is to evaluate through experimental sts whether a modified screw extruder (as used the food and plastics industries) could be ed to feed a finely-ground slurry (ground sawst in heavy oils or water) into a pressure actor system. Thick, concentrated cellulosic urries have high apparent viscosities and velop a pressure head in such extruders. Slurry operties, machine operating characteristics, and cessary scale-up criteria are being investigated. typical single screw extruder is shown in Fig. 1 d a food extruder for macaroni is shown in Fig.2.

is project is of direct interest for the DOE omass Liquefaction Facility, Albany, Oregon and e DOE-sponsored cellulosics gasification work Ballston Spa, New York. It appears that this truder feeding system can greatly simplify the ocess by eliminating the separate grinding, ying, slurrying and piston-pumping steps by mbining all operations in one machine. Further e feeding of a much higher biomass solids contration will reduce processing costs by ducing the quantity of recycle oil.

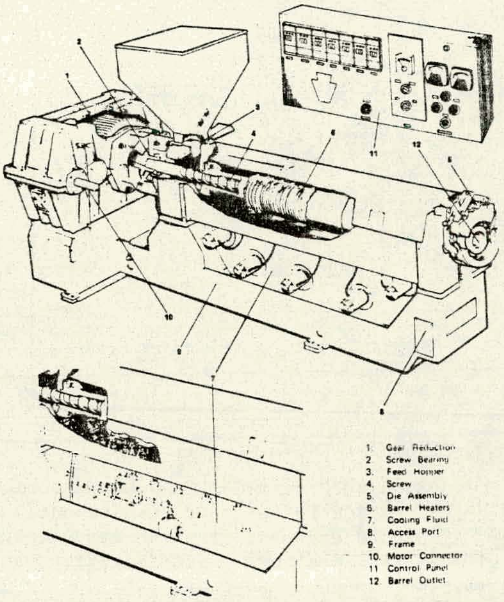

Fig. 1. Typical Single Screw Extruder

CONTRACT TASKS

Summary of Contract Tasks for the first year were as follows:

- Task 1 - Determine rheological and related properties of concentrated cellulosic slurries in oils and water.
- Task 2 - Evaluate pumping and pressure head developed in single screw extruders.
- Task 3 - Evaluate stiff cellulosic slurry masterbatching and pelletizing.
- Task 4 - Evaluate alternative types of extruders in outside field tests.
- Task 5 - Analyze and correlate extruder feeding performance data.

Task 1. Determination of Rheology of Fluid Cellulosic Slurries

The contractor shall measure the rheological and related physical properties of sawdust/oil and sawdust/water slurries, as a continuation of Task 1 under current Contract No. EW-78-S-05-5679. Measurements will consist of apparent viscosities of various slurry concentrations from 10% to maximum possible in conventional rheometers,

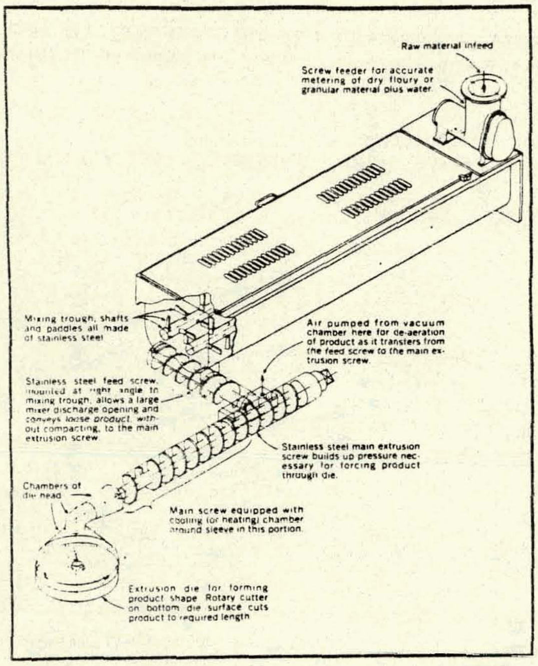

Fig. 2. Food Extruder System for Macaroni Production

temperature range 80°F to 300°F. Wood will be obtained from the DOE Waste-to-Oil Experimental Facility, Albany, Oregon and Wright-Malta Corporation, Ballston Spa, New York. It is estimated that between 40 and 60 runs, generating 200 to 300 data points, will be needed to develop the necessary information.

Task 2. Determination of Rheology of Stiff Cellulosic Slurries

The Contractor shall provide rheological data, the same as Task 1, except for covering higher cellulosic slurry concentrations, in the range of 30 weight percent up to probably 80 weight percent. It will be necessary to construct a high-pressure large-diameter tube rheometer, suitable for use in an existing Instron Testing Machine as a part of this task. It is estimated that 10 to 20 runs can be made in the given period.

Task 3. Compression of Cellulosics in Pasadena Press

The Contractor shall measure the compressibility and void volume available for the slurry oil for sawdust/oil and sawdust/water slurries over the conditions as follows:

a. Pressure, 0 to 10,000 psi
b. Temperature, 80°F to 300°F
c. Cellulosic slurry concentrations from 10 weight percent to maximum possible

A simple "pressure mold" with bottom outlets for "squeezed-out" oils or water will be constructed for use in an existing Pasadena Press, normally used for molding plastics. It is estimated that 10-20 runs will be made in this simplified apparatus.

Task 4. Design and Construct Laboratory Compression Cell for Simultaneous Slurry Fluid Flow

The Contractor shall design a unit, based on the experience gained in Task 3, and construct said unit that can be used to hold concentrated slurry samples for simultaneous compression and flow of the slurry fluid, adaptable for use in the Pasadena Press and/or an existing Instron Testing Machine. Pressure drop and flow rate data will be taken under the conditions of Task 4, similar if possible to the runs made under Task 4, for use in correlating and comparing results.

Task 5. Measurement of Friction of Stiff Cellulosic Slurries

The Contractor shall design, construct and obtain experimental data on the (a) friction between stiff cellulosic slurries and steel plates, and if possible (b) design the tests so that some effects of friction between particles within the slurry itself can be measured. This task is exploratory in nature, and it is estimated that only 5-20 runs will be made in the time available under conditions as follows:

a. Pressure, 0-20 psi compressive forces.
b. Friction force, lower range only (measured by strain gauges).
c. Room temperature only.

Task 6. Design and Indicate Feasibility of an Extruder Barrel Simulator

The Contractor shall design an extruder barrel simulator with wall configurations as used in food extruders. It will not be constructed but rather its feasibility be indicated by simplified, makeshift laboratory tests.

Task 7. Determination of Stiff Cellulosic Slurry Frictional Forces by Extruder Screw Simulator

The Contractor shall determine the surface frictional forces of stiff cellulosic slurries, in the form of solid masterbatches, for an estimated 10-2 masterbatch samples. This activity will be performed under subcontract to Dr. Chung, Rensselaer Polytechnic Institute, Troy, N.Y., the originator of the only known extruder screw simulator (developed under funding from the National Science Foundation).

Task 8. Formulate Cellulosic Mixtures with Minimum Void Volume for Slurry Liquid

The Contractor shall determine the minimum void volume of uncompressed cellulosic mixtures of different sizes of sawdust by the following te[illegible]niques:

Screen dried Albany pilot plant sawdust (pro-
:ed by hammer mill) and recombined in
:ferent proportions.
< "fibrous" cellulosic particles with more
:anular" types of particles.
Repeat item a. with a portion of sawdust being larger particle size than normally produced in Albany pilot plant.
Grind dried wood chips from Albany pilot plant by means of existing plastics scrap grinder, and repeat some of the formulations under item a. for comparison.

e Contractor shall check a selected number of e samples formulated herein for their void lumes under compression, using the apparatus nstructed under Task 4.

sk 9. Operation of Laboratory Extruder at gher Speeds

e Contractor shall operate a laboratory single- rew extruder at higher than normal speeds, per- rming a parametric investigation to determine e relationship between the extruder output pres- re and the extruder operating variables. These clude temperature, pressure, throughput rates d power consumption, at screw speeds from about rpm up to highest possible with a given ex- uder. The Contractor shall make an estimated 10 runs on the larger $1^3/_4$ inch-dia. extruder r comparison purposes.

sk 10. Chemical Pretreatment of Cellulosics der Extruder Conditions

e Contractor shall make 5-10 exploratory runs utilizing the extruder, with its inherent erating advantages of elevated temperature, essure and shear forces, to determine the effects light chemical pretreatments on the cellulosics. boratory "tubing reactions" shall be performed select the more promising chemical pretreat- nts.

sk 11. Abrasion-Resistant Coatings Bench-Scale ster

e Contractor shall adapt a known abrasion- sistant laboratory tester, so that it can be ilized for materials suitable for coating truder screws handling slurries. Three rasion-resistant formulations will be selected demonstrate the performance of the unit.

sk 12. Cellulosic Slurry Masterbatching Using ker-Perkins Intensive Mixer

e Contractor shall formulate laboratory master- tches of required cellulosic slurries using an isting laboratory Baker-Perkins Intensive Mixer. estimated 20-30 batches will be prepared, and rther tested as needed in Tasks 4 and 7.

Task 13. Evaluation of Pelletized Feedstocks

The Contractor shall complete any outside evaluations under subcontract of pelletized feedstocks, as listed under Task 3 of current Contract No. EW-78-S-05-5679, deferred due to any approved changes in work schedule of that Contract.

Task 14. Field Testing of Alternative Types of Extruders

The Contractor shall perform these outside tests under subcontract, as defined in Task 4 of current Contract No. EW-78-S-05-5679, deferred as indicated so that higher priority could be given to single-screw extrusion and to laboratory work on cellulosic slurry properties.

Task 15. Analysis and Correlation of Results

The Contractor shall prepare a draft final report, to be submitted within 30 days of completion of the work, consisting mainly of:

a. Correlation Analysis of the Technical Data
b. Suitability of each given extruder to develop a pressure head, and under what conditions.
c. *Conclusions and Recommendations*.

PERFORMANCE TARGETS

This basic overall goal was achieved during the first year of the contract in 1978, utilizing Albany, Oregon process facility sawdust, giving results of (a) pumped feed slurries in extruder developing up to 8,500 psi pressure head, whereas 3,000 psi needed and (b) pumped up to 70 weight percent feed slurries, whereas pilot facility piston pumps are limited to 25-30 weight percent sawdust. During 1979, scale-up factors are being determined which involve extruder design, extruder operation conditions, properties of sawdust feed and properties of the necessary carrier oil. A good start has been made upon understanding the rheology of cellulosic slurries over a wide range of sawdust concentrations. Apparent viscosities increase sharply at concentrations above 15 weight percent, and are difficult to measure in conventional rheometers. Modified pressure rheometers and "frictional" devices are being constructed to solve this problem.

APPROACH TO CURRENT WORK

It was decided that for the second year (1979) the order in which the tasks are carried out should be revised to accomplish the ultimate goals of the project. In view of the good success to date in meeting the ultimate project goals by means of single screw extruders, it is believed that this approach should be pursued at a higher level of effort. Simultaneously, it is obvious that there should be an expansion of tasks related to the physical and rheological properties of sawdust per se and of various sawdust slurries. In order to accomplish these tasks it was necessary to defer

(a) the outside tests on alternative types of extruders (food, twin screw and tapered screw extruder, and (b) some of the evaluations of pelletized food stocks (masterbatch lots), until the latter part of 1979. Thus, it makes sense to concentrate at this time more heavily upon single-screw extrusion and quantifying the factors involved in the development of a pressure head in the extruder when feeding stiff cellulosic slurries. One can then plan more intelligently on the outside tests under subcontract for evaluating pelletized masterbatch feedstocks and three alternative types of extruders.

With respect to the factors affecting pressure head in the extruder, the void volume available for the slurry fluid is a major factor. How small is this volume at 5,000 psi ? Can it be changed by particle size distribution of the sawdust ? The friction between two sawdust particles helps build up a pressure head. Is this a major factor compared with the effect of void volume or does the slurry oil coating the cellulose merely act as a good lubricant ? How necessary is such a lubricant effect in order to keep the stiff cellulosic slurry moving, rather than just plugging by squeezing the oil out of the sawdust ? These and many other points discussed need to be answered.

One could argue that since the extruder apparently works as a feed pump for very stiff (dense) cellulosic slurries, why not do all further work in the extruder itself. It should be obvious by now that such an approach would be highly empirical and dangerous to use in any scale-up for design purposes.

KEY RESULTS TO DATE

We have achieved the basic overall goal of the project, just a little under 8 months after its start. This basic overall goal is that a cellulosic slurry has been pumped under specifications exceeding those needed in the Albany cellulosics liquefaction pilot plant, namely, (a) at a feed pressure in the range of 5,000 to 8,5000 psi, when about 3,000 psi is needed, and (b) to date a sawdust concentration of 70 weight percent has been used (and pumped) in the feed slurry, whereas the best the piston pumps of the pilot plant can do is handle a slurry in the range of 25-30 weight percent sawdust.

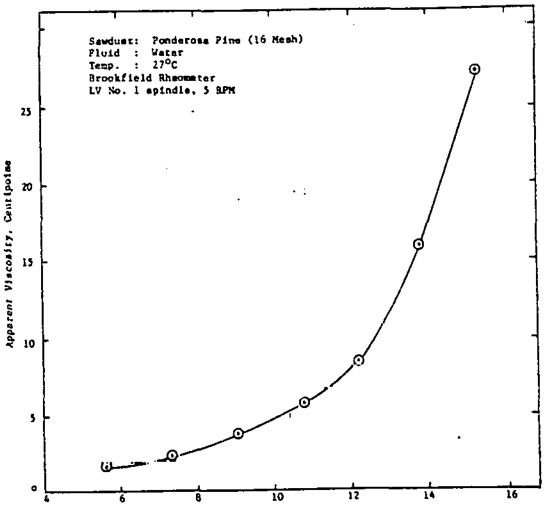

Fig. 3. Rheology of Medium-Sized Cellulosic/Water Slurry

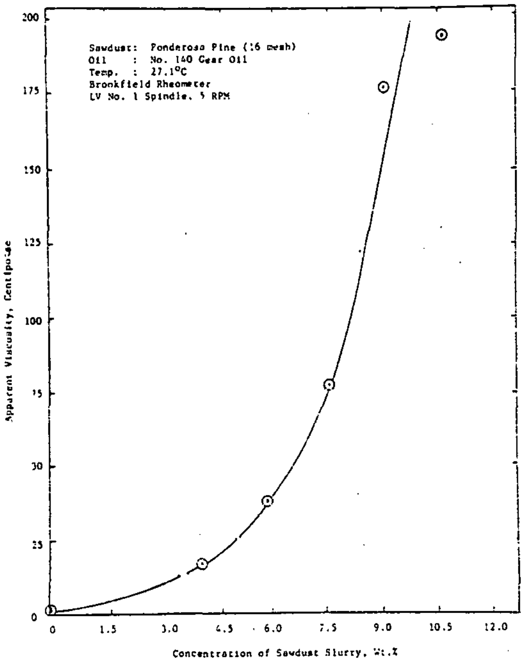

Fig. 4. Apparent Viscosity of Medium-Sized Cellulosic-Gear Oil Slurry

A good start has been made upon understanding the rheology of cellulosic slurries over a wide range of sawdust concentration. We know that the apparent viscosity of sawdust slurries increases rapidly at concentrations of sawdust greater than about 10 weight percent. This increased apparent viscosity occurs in both water and gear oil slurries as shown on Figures 3 and 4, respectively. However, it is not known as yet how high these apparent viscosities are for the very concentrated sawdust slurries, because it has been difficult to measure them in conventional pressure "capillary" rheometers. This is being overcome by modifying a pressure rheometer used successfully at Washington University on fiber glass/polymer liquid mixtures (slurries).

As suspected from the beginning of this project, there are several additional factors (in addition to apparent viscosity of the slurry oil and the fineness of the sawdust) that contribute to developing a pressure head in a single-screw

ruder feeding a cellulosic slurry. A start
n made upon the significance of these fac-
measuring in the laboratory some of the
ies of sawdust, slurry oils and various
/oil slurries. The more important pro-
rties determined to date are as follows:

Bulk Densities and Absolute Densities. The
k density of sawdust, as one might predict,
ries widely with particle size and particle size
stribution. As a further illustration of
fferences in bulk densities, a spherical poly-
hylene bead has an absolute density of 57.3
/Ft^3, a bulk density of 36.3 lb/Ft^3 and, if
ground, its bulk density drops to 18.1 lb/Ft^3.
se factors affect the masterbatching of con-
ntrated slurries, such as sawdust/oil, and
eding characteristics of single-screw extruders.

Effect of Sawdust Compressibility Upon Minimum
id Space for Slurry Liquid. Sawdust/oil slur-
es were compressed in a hydraulic Pasadena Press
rmally used to mold plastics) up to 10,000 psi
essure. Finely-ground sawdust slurry can be
mpressed about 6-fold at 5,000 psi and some 7-
ld at 10,000 psi, as shown in Figure 5. It
pears that there is approximately a one to one
rrespondence on the reduction of void volume
maining for the slurry oil to occupy. By cal-
lating the oil squeezed out of the slurry during
mpression, the void volume for oil remaining
10,000 psi is about 2 to 5 percent, as shown in
gure 6. Sawdust mixtures containing larger
wdust particles appear to be slightly more
fficult to compress, but the proper amount of
nes" may not have been used in these preliminary
sts.

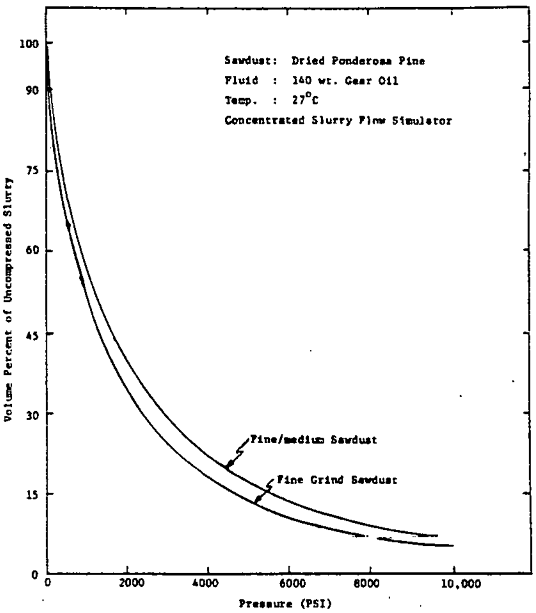

Fig. 5. Compressability of Sawdust/oil Slurry

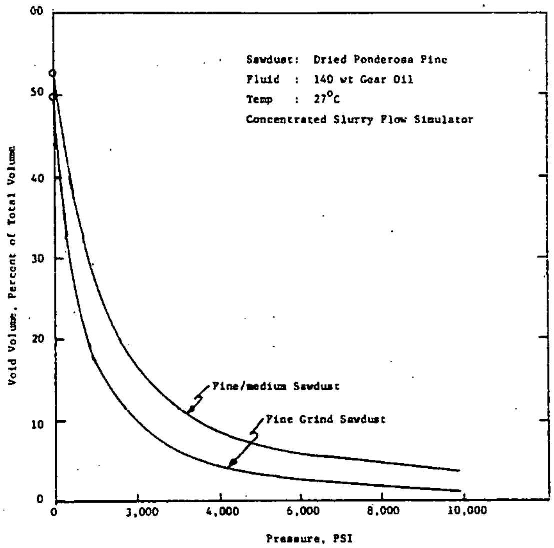

Fig. 6. Effect of Pressure on Void Volume of Sawdust

There is now experimental evidence that the particle size, the particle size distribution (PSD) and the physical form of the sawdust particles after grinding are significant factors in the development of a pressure head in a single-screw extruder. The volume of void space available for the liquid phase and the physical shape of the particles, for example, affect (a) the pressure drop of the liquid trying to flow through the sawdust voids and (b) the frictional resistance of the sawdust particles rolling over each other under shear, respectively.

The size of the particles and sawdust particle size distribution (PSD) in conventional grinding depends upon the type of grinder, screening system, recycle of oversize and degree of sawdust dryness. The shape of sawdust particles are affected mainly by (a) the nature of the wood source and (b) the type of grinder utilized. However, there are many ways of modifying the above three variables (particle size, particle shape and particle size distribution) as listed below but not necessarily limited to these techniques:

a. Mix "fibrous" cellulosic particles with more "granular" types of particles by blending and screening operations.
b. Change fibrous and/or granular nature of sawdust particles by various grinding techniques.
c. Change fibrous and/or granular nature of sawdust particles by various chemical pretreatment techniques.
d. Modify particle size distribution by utilizing proper portion of each increment of size in a blend to minimize the free void volume of bulk sawdust. This would be analogous to what is

done in concrete, which uses a fixed amount of each aggregate of a given size to minimize the void space for cement and hence a minimum of water in the premix.

e. Utilize a much larger portion of larger sawdust particles in a given feed by blending with the proper volume of sawdust fines to give a minimum of void volume for the slurry oil to occupy.

A start has been made on the "design of PSD" by measuring the physical size, PSD and aspect ratio (particle length to particle diameter, l/d) of the dried sawdust being used in this work. It has been found that the finely-ground sawdust is fibrous in nature, with a reasonably high aspect ratio (l/d), as shown in Figure 7. Thus the equivalent diameter must be defined or chosen on some formula basis of diameter, length and/or thickness.

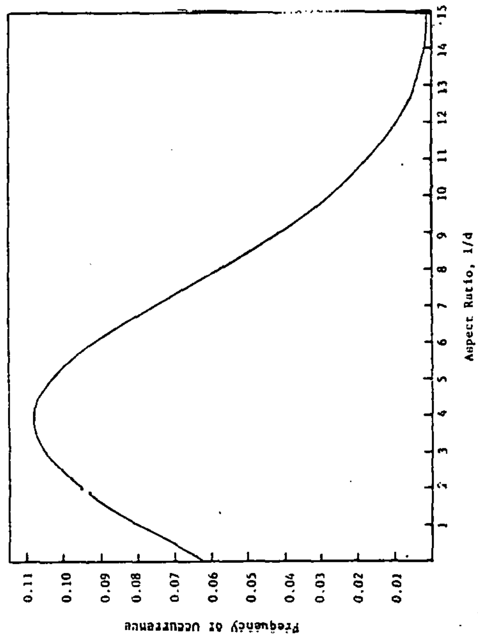

Fig. 7. Distribution Curve of Aspect Ratio of Medium-Sized, Dried Sawdust

Laboratory rheology data were obtained for Albany sawdust slurries on a Brabender Plasti-Corder in the Department of Plastics Engineering, University of Lowell, Lowell, Mass. under the direction of Dr. N.R. Schott. Some of the preliminary data on torque measurements for Albany sawdust/glycerine and sawdust/Petrolite 500 slurries ranging in concentrations from 50 to 95 weight percent cellulosic sawdust, were determined. The corresponding flow rates were also obtained. Further work using other liquids, including Allied Chemical AC-629A and AC-392, modified lignins and other low molecular weight polymer liquids, is in progress and will be reported at a later date. It should be noted that "stiff slurries" of concentrations up to 95 weight percent sawdust (which basically is dry sawdust) were handled in this unit. The extruder operated smoothly at 50 weight percent sawdust/glycerine slurry, whereas at 80 weight percent and above the throughput and torque requirements were very erratic.

Experimental pelletizing runs were made upon a laboratory California Pellet Mill, in the labs of California Pellet Mill Co., Crawfordsville, Indiana, in December, 1978, by experienced California Pellet Mill personnel and observed by Dr. Don H. White, University of Arizona. Low molecular weight Allied Chemical AC-6 polyethylene wax, powdered form, was used as "lubricant" on sawdust obtained locally in Indiana from Swain Industries. Pellets containing 50,60,70,80 and 90 weight percent sawdust were made without difficulty. Later, one drum each was made on a larger machine and the pelletized products shipped to Tucson for further evaluation in extruder tests The California Pellet Mill has world-wide acceptance for producing pellets from powders economically, ranging from dehydrated alfalfa to rabbit feed.

Brief tests were observed by Dr. White in Des Moines, Iowa, December 1978, wherein local sawdust was preblended with various amounts of glycerine and then extruded successfully in the Insta-Pro food-type extruder. However, some overheating of the sawdust occurred and maximum pressure development could not be determined due to lack of proper instrumentation and certain design features. It appears that further tests upon this extruder would be justified, but only a properly modified version of the unit.

The performance of a laboratory Henschel High-Intensity Mixer was observed by Dr. White in December, 1978 in the research labs of HPM Corporation, Mt. Gilead, Ohio, upon masterbatching Albany sawdust with various liquids. The unit was then shipped to the University of Arizona under a rental agreement for further tests in early 1979. The results of these tests will be reported at a later date.

Arrangements were made with Dr. C.I. Chung, Materials Engineering Department, Rensselaer Polytechnic Institute, Troy, New York, to determine the surface friction characteristics of two stiff cellulosic slurries in his laboratory. He has developed a unique "screw simulation" under National Science Foundation funding for use in evaluating solids conveying in a plasticating extruder, and is believed to have the only unit available. His research has contributed to an understanding of extruders in the polymers field, and if the first two series of tests are encouraging, further tests are planned for 1979.

FUTURE PLANS

Continued support at about the present level (adjusted for inflation) will be needed in 1980 and 1981 to complete the project. This involves determining the scale-up factors and designing a pilot unit by mid-1980. If such a unit were constructed and delivered to the Albany, Oreg[illegible] facility by early 1981, then simultaneous

:ions on the existing laboratory unit and the
unit would confirm scale-up and cost factors
1-1982.

he desired pressure head and pumping of higher
lurry concentrations were achieved under one set
f conditions in 1978. The effects of slurry
eed properties, oil viscosity, sawdust pre-
reatments, sawdust size and particle size dis-
ribution, machine operating conditions and
everal related factors will be determined by mid-
980. Most of this work will concentrate upon
ingle screw extruders, but field tests will be
onducted in early 1980 upon several alternative
ypes of extruders. Design and scale-up criteria
ill be accomplished in late 1980 through 1982
y constructing and operating a pilot unit.

CKNOWLEDGMENTS

e wish to acknowledge especially the assistance
f Dr. Chan I. Chung, Rensselaer Polytechnic
nstitute, Dr. George Kruder, HPM Corporation,
r. Nick R. Schott, University of Lowell, and
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ncluded John Drazkowski, Matt Frondorf, Bob
enckel, Al Hess, John Mesko, Diana Reckart and
en Yan.

NOTES

PRETREATMENT OF BIOMASS PRIOR TO LIQUEFACTION

Larry L. Schaleger, Nasser Yaghoubzadeh, and Sabri Ergun
Lawrence Berkeley Laboratory
University of California
Berkeley, California

STRACT

etreatment of biomass prior to liquefaction was
oposed by the Bureau of Mines as an optional pro-
dure for injecting biomass into high pressure
ssels. In the Bureau of Mines process the pur-
se of pretreatment was to change the rheological
operties of biomass so that biomass flour/oil
urries containing up to 50 percent treated bio-
ss could be injected into high pressure vessels.
wrence Berkeley Laboratory researchers investi-
ted the rheological and chemical changes occurring
on pretreatment of wood and conceptualized a
ocess scheme in which wood chips are hydrolyzed
der mild conditions to obtain a concentrated
ueous slurry that can be directly injected into
e liquefaction system without a carrier oil.

is presentation deals with the results of the LBL
udies on pretreatment of biomass prior to lique-
ction and the process conceptualized at LBL.

CKGROUND

. the Bureau of Mines-Albany biomass liquefaction
heme, wood is converted to oil by treating with
rbon monoxide and aqueous sodium carbonate
talyst at elevated temperatures and pressures.
me of the major developmental difficulties en-
untered at the Albany 3 TPD facility have arisen
connection with the front end of the process
ere wood chips must be reduced to a form capable
being pumped into the reaction system. This
per describes the current status of research
rected toward the development of chemical
ternatives to drying-and-grinding for the purpose
wood preparation.

e original process design called for green wood
ips to be dried, milled to a -50 mesh flour and
en slurried with vehicle oil at a solids concen-
ation of 30%. However it has proved impossible
pump a 30% slurry. Slurries of 20% solids have
en pumped with some difficulty but most of the
velopment work at Albany has been restricted to
ncentrations of 10%. The question is critical
nce the economic feasibility of the process
nges on being able to inject solids at the 30%
vel. It is clear that the lower the concentra-
on of biomass in the feed to the reaction system,
e larger the plant size for a given production
te.

addition to difficulties encountered in pumping
ncentrated slurries of wood flour, there is also
expense involved in drying and grinding. An economic analysis showed that 22% of the cost of product oil would be attributable to these operations.

That the potential for such problems existed had been recognized early. One option considered by the Bureau of Mines was chemical pretreatment. The basic idea was to subject wood chips to partial carbonization in order to increase their bulk density and thereby favorably alter the rheological properties of the slurry feed. It was hoped that, after drying and grinding, the resulting powder could be slurried at concentrations as high as 50%. An engineering data base for pretreatment was prepared by BOM and two pretreatment reactors were designed and installed at Albany.

In the BOM pretreatment technique wood chips and water (30/70)* are heated to 500°F and 700 psig for 60 minutes. After cooling, the solids are separated by vacuum filtration, dried and pulverized to -50 mesh, mixed with recycle oil and injected into the liquefaction loop.

The results of four pretreater test runs conducted at Albany were generally unsatisfactory. First, the filtering system in the plant was underdesigned and did not perform properly. Product solids proved unexpectedly difficult to dewater; drying and grinding presented additional difficulties. It appeared that new approaches to the problem were needed.

LBL PROCESS DEVELOPMENT

Several pretreatment runs were conducted by LBL on a small scale, duplicating the conditions used at Albany. Significantly, it was found that the carbohydrate fraction of the wood was almost entirely destroyed under these conditions; the product analyzed as a dehydrated char. In our view this was a serious shortcoming since carbohydrates have been shown to liquefy readily under BOM/Albany conditions, whereas chars typically are intractable.

On the other hand we noted that extensive size degradation occurred in the course of pretreatment, and we reasoned that it might therefore be possible to eliminate filtration, drying and grinding

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*Plant practice established 23% solids as a more manageable concentration.

altogether. Thus a new goal for the development of a pretreatment process was established: to produce, with no sacrifice of the original heating value, a concentrated slurry of biomass capable of being pumped directly into the liquefaction system. A five-point process development plan was formulated for the purpose of guiding the overall research effort:

1. research chemical pretreatment methods
2. research
   a. size degradation efficiency
   b. rheological properties of solids
3. research liquefaction of treated biomass
4. obtain data base for large scale test run at Albany
5. conceptualize a process.

CHEMISTRY OF PRETREATMENT

Initial guidance as to research direction was provided by an examination of the chemical composition of Douglas Fir (*Pseudotsuga menziesii*), the predominant species in the Albany area (Table 1).

Table 1. Composition of Douglas Fir (moisture free)

Cellulose	41%
Hemicellulose	26
Lignin	28
Uronic anhydride	3
Ether-soluble extractives	1.0
Acetyl	0.6
Ash	0.3
	99.9

Cellulose and hemicellulose, comprising the carbohydrate fractions, should be preserved, either as such, or in the form of their constituent sugars which are liberated upon hydrolysis. Lignin, a three-dimensional matrix of oxygenated phenylpropane units, accounts for most of the remainder.

The conditions specified by the Bureau of Mines are apparently too severe since they result in charring and destruction of carbohydrate. Milder conditions require the use of a catalyst. Obvious catalysts are acids or bases. In aqueous media the polysaccharides undergo hydrolysis catalyzed specifically by hydrogen ion. Mild acid conditions, for example 0.05% sulfuric acid at 140-180°C, suffice to hydrolyze hemicellulose to give (in the case of Douglas Fir) mannose, glucose, xylose, galactose and arabinose in the ratio of about 10:4:2:2:1. More severe acid conditions promote the liberation of glucose from cellulose but simultaneously cause the degradation of the more readily liberated hemicellulose sugars [1]. Cooking with strong aqueous sodium hydroxide, on the other hand, results in the partial solubilization of lignin as in the well-known but antiquated "soda process" for the manufacture of wood pulp. Thus either acid or base might conceivably be used to chemically disrupt the structural integrity of wood chips in order to realize the objectives of pretreatment.

RESULTS OF CATALYTIC PRETREATMENT

Sodium Hydroxide

It would be desirable to effect pretreatment under basic conditions since subsequent liquefaction also requires basic catalysts. Furthermore the effect of concentrated base is to break down and solubilize the ligneous fraction. This might be important because there is evidence to indicate that native lignin is resistant to the conditions of liquefaction.

It was found that treatment with 20% sodium hydroxide at 356°F for 60 minutes, conditions somewhat more severe than those used in the pulp and paper industry, leave the fibers of cellulose intact and give rise to viscous slurries characteristic of wood pulp. Since these slurries appeared unsuited for liquefaction, further research on basic pretreatment has been suspended.

Dilute Aqueous Mineral Acid

More promising results are obtained in the case of mild acid hydrolysis accompanied by vigorous agitation. Conditions have been found which resul in complete hydrolysis of the hemicellulose with little loss of carbohydrate values (Table 2). Under these conditions the remainder of the wood undergoes extensive size reduction, forming spherical rather than fibrous particles. The resulting slurries have been shown to possess superior rheological characteristics.

Table 2. Data Base for Hydrolysis

Wood/Water Ratio	23/77
Heat-up Time	12 min
Temperature	180°C (356°F)
Acid (pH)	0.05% H_2SO_4 (2.0)
Retention Time	45 min

Approximately one dozen hydrolysis experiments hav now been conducted in autoclaves ranging in size from 400 ml to 400 gallons (i.e., a test run at Albany) with essentially identical results. Material, carbon, hydrogen and energy balances are typically 0.85 or better.

The success of hydrolytic pretreatment in reducing wood chips to fine particles is probably due to a combination of chemical and physical action. Most of our experiments have been conducted in a 10-gallon digester equipped with a hydraulic agitator which is run at 900 rpm. The shaft contains three pitched-blade turbine impellers and a two-piece baffle assembly. Using this unit we find that about half of a 4.0 kg charge of dry wood chips, nominally 1¼" in length, is reduced to -20 mes[illegible] or finer under the conditions specified above.

se particles larger than -20 mesh can for the rt be crumbled between the fingers.

DEVELOPMENT CONSIDERATIONS

eral experiments designed to test the suitabil- of pretreated wood slurries for liquefaction e been performed. Under identical conditions ehydrolyzed wood gives slightly higher yields of l than wood flour.

order to adapt the acid pretreatment technique the Albany process, neutralization with calcium droxide and separation of precipitated calcium lfate sludge would be required. This is an vious disadvantage which could be avoided if) a basic pretreatment or (b) an acidic liquefaction method could be developed. Active research in this area is under way. Several promising acid catalysts for liquefaction have been discovered but have not yet been fully evaluated.

Acknowledgments

We are grateful to Prof. David L. Brink of the University of California Forest Products Laboratory for guidance and for the use of facilities.

References

[1] Saeman, J.F., "Kinetics of Wood Saccharification," Ind. Eng. Chem. 37, 43 (1945).

NOTES

BENCH SCALE RESEARCH IN BIOMASS LIQUEFACTION
IN SUPPORT OF THE ALBANY, OREGON, EXPERIMENTAL FACILITY

Contract Number: EY-76-C-06-1830-AJ-01-89018
Contract Date: October 1976 - Continuing

Douglas C. Elliott and Glen M. Giacoletto
Pacific Northwest Laboratory
Operated by Battelle Memorial Institute
P.O. Box 999, Richland, WA 99352

BSTRACT

ne objectives of this project are to determine hrough bench scale investigation process changes equired to facilitate fuel oil production by the O-Steam process at the Albany Biomass Liquefacion Experimental Facility; provide routine and pecialized analytical support for the facility perators; and determine process feasibility on a ench scale of new biomass liquefaction techiques which may be appropriate for development t Albany. Our work has been based on the use of batch autoclave wherein biomass is converted in oil slurry. During the past year our work has ealt with catalyst concentration, catalyst addiion technique, substitution of hydrogen donor olvents for the slurry oil, and comparison of ifferent feedstock reactivities. We have made xtensive use of analytical instruments to proide both routine and specialized analysis of the lbany products and feedstocks. We are also curently subcontracting development work for a new iomass liquefaction technique based on thermolyis of biomass in a continuously varying solvent ystem.

BJECTIVES

ne objectives of this project are threefold:) to determine through bench scale investigation rocess changes which will facilitate fuel oil roduction by the CO-Steam process at the Albany, regon, Biomass Liquefaction Experimental Facilty; 2) to provide routine and specialized anaytical support for the Albany facility operator s needed and as available at Pacific Northwest aboratory (PNL); and 3) to determine the process easibility on a bench scale of new biomass liqueaction techniques which may be appropriate for uture development at Albany.

PPROACH

ckground

t was recognized early in the history of the lbany facility that there might be variations of ne process concept which would enhance its viality in the near term. Consequently, PNL was authorized to undertake a supporting effort to investigate and confirm some of the key process parameters and effects of varying operation conditions. This support work was to be done in a batch mode in existing autoclaves at PNL. Also, we agreed to provide analytical or other requested support to the facility operating contractor. As a further expansion of our charter in the past year, we are now examining new biomass liquefaction processes which may be appropriate for future development at the Albany facility. Development of one such process, Biomass Thermolysis, is proceeding under a subcontract to Wright-Malta Corporation.

Equipment

The basic autoclave used in our process development experiments is shown in Fig. 1. It is an Autoclave Engineer's 300 ml bolted closure autoclave of 316 stainless steel with a magnedrive

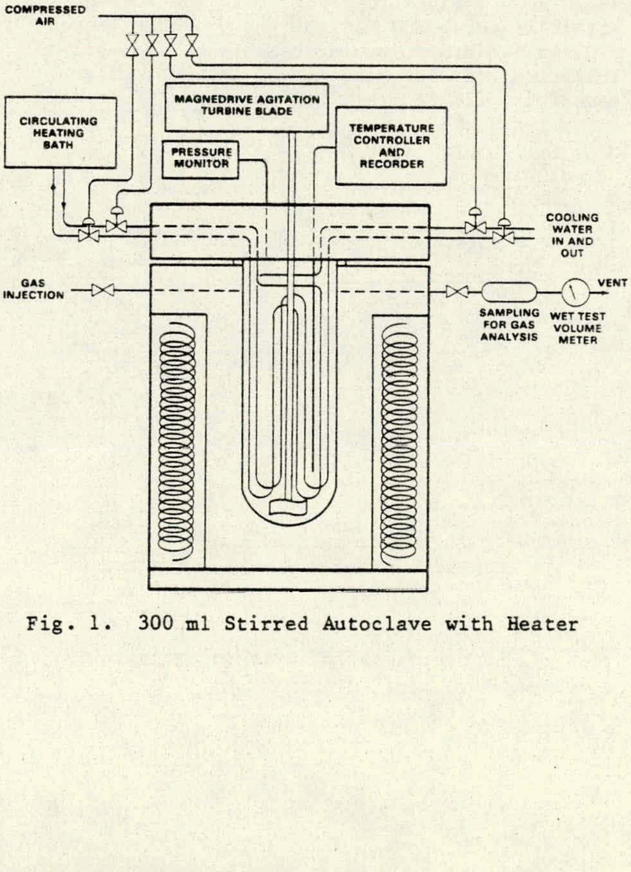

Fig. 1. 300 ml Stirred Autoclave with Heater

packless stirrer. The sampling and addition ports in the autoclave cover were used to add an internal heating coil system in addition to the standard external 1.2 kW resistance jacket furnace. A heat transfer oil, Therminol 66 from Monsanto, which is heated at up to 650°F, is pumped through the coil. The circulating fluid is heated by a 1.5 kW external electrical heater. With this combined heating system, the reaction mixture of anthracene oil, wood flour, synthesis gas and aqueous sodium carbonate can be heated quickly to the desired reaction temperature (7 minutes from 300°F to 660°F). To quench the reaction the power to the furnace is shut off, the furnace is dropped away from around the autoclave, and a fan is directed at the autoclave body. Water is then sent through the internal cooling coils to quickly lower the temperature (5 minutes from 660°F to 300°F). The heating and cooling fluid systems are remotely controlled by air-actuated valves. Temperature and stirring are remotely controlled and monitored. Pressure is also remotely monitored. After the autoclave has cooled, the gases are vented and sampled and the product oil can be recovered for further analysis.

A second experimental system is being developed as part of the biomass thermolysis task and a diagram of this thermolyzer unit is shown in Fig. 2. This unit is presently being fabricated by Wright-Malta Corporation and is intended to become operational during the summer of 1979. In the thermolyzer unit the biomass will be suspended in a mixture of organic solvents at about 570°F in a vessel built to withstand the system vapor pressure at the reaction temperature (approximately 500 psig). Screens at the top and bottom of the reactor will contain the biomass while the hot solvent mixture is pumped through the mass. As the biomass thermally decomposes at these relatively mild conditions, the solvent will act to disperse and isolate the products thereby minimizing intermolecular reactions, i.e. polymerization and char formation. Intramolecular reactions will be maximized which, in the case of biomass, will lead to the scission of carbon and oxygen bonds and the elimination of carbon oxide and water. The selection of solvents is a critical aspect of this wood liquefaction process. At every stage of decomposition, it is essential that the wood structure be as swollen as possible which is made difficult by the changing character of the decomposing wood. Ideally, the solvent mixture would diminish in polarity, as does the wood, during its decomposition. The change in solvent mixture polarity will be accomplished by the release of the more volatile polar solvents (water and light alcohols) and CO$_2$ from the top of the reactor.

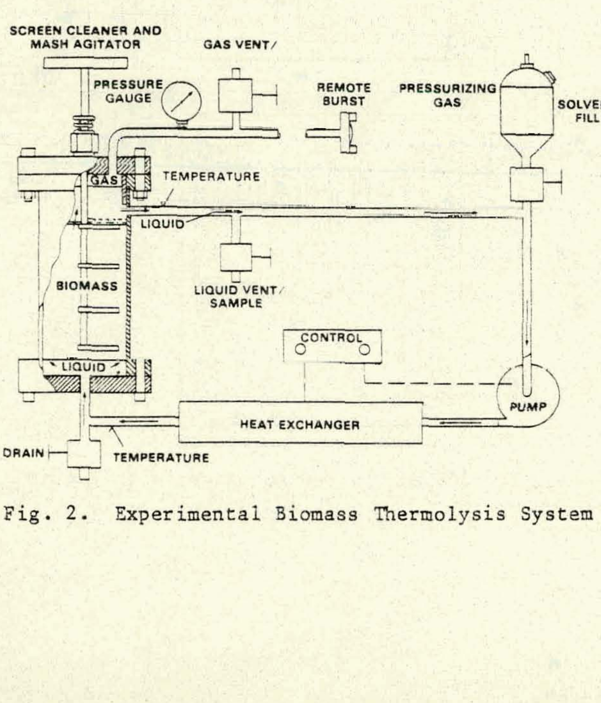

Fig. 2. Experimental Biomass Thermolysis System

RESULTS

Catalyst Addition and the Effect of Water

Early operations at the Albany facility indicated that the Na$_2$CO$_3$ catalyst was ineffective. One possible cause of catalyst ineffectiveness could have been inadequate mixing of the catalyst with the wood feed. To study this we did autoclave experiments to test 1) dry mixing of the catalyst with the wood/oil slurry, 2) solution impregnating the wood with the catalyst, 3) solution impregnating followed by drying, and 4) catalyst solution addition to the wood/oil slurry. Using these four techniques a series of experiments, summarized in Table 1, were performed. The results of these experiments support the conclusion that the direct intimate mixing of catalyst and wood is not the important parameter but that the amount of water present has a greater effect even on an oil-coated feedstock. As shown in Table 1, the amount of reactivity, as measured by solids conversion and CO consumption, and the lack of pyrolysis, as evidenced by low hydrocarbon gas production, correlate with the initial amount of water, not the catalyst application technique.

To further analyze this result a second series of experiments was performed. A graphical presentation of these results is given in Fig. 3. A dramatic reduction in residual insoluble solids is the result of up to ten times the design operational amount of water. These results are estimated to show a gain from 96 percent actual wood conversion (disappearance) to +99 percent. In the Albany facility, where conversion has ranged all the way from 50 to 90 percent per pass, this experimental result could have a significant effect on the unit's operation. Not only might higher conversion percentages be attained; but additional water, which would not be completely flashed in the reactor letdown, would also aid in lowering the viscosity of the product during the recycle, as well as in the slurry make-up stage.

Catalyst Effects

We have also performed autoclave experiments to test catalyst substitutes for sodium carbonat

TABLE 1. COMPARATIVE EFFECTS OF CATALYST AND WATER

Reaction Mixture	Acetic Acid Insoluble	Calculated Conversion	Mole CO Consumed	Percent CO Consumed	Mole HC Gas Produced
Catalyst on wood 10X water	0.43	95.2	.167	40.3	.001
Catalyst on wood 1X water	1.74	86.1	.094	19.1	.026
Catalyst on wood 1/20X water	4.38	66.7	.042	8.6	.043
Catalyst soltn into wood/oil 10X water	0.58	93.5	.087	33.7	.003
Catalyst soltn into wood/oil 1X water	1.69	86.5	.109	20.5	.021
Dry catalyst into wood/oil 0X water	∿4	∿70	.071	14.5	.075

330°C, 1-1/2 hour, 10 g wood flour, 1200 psig initial CO/H_2:60/40

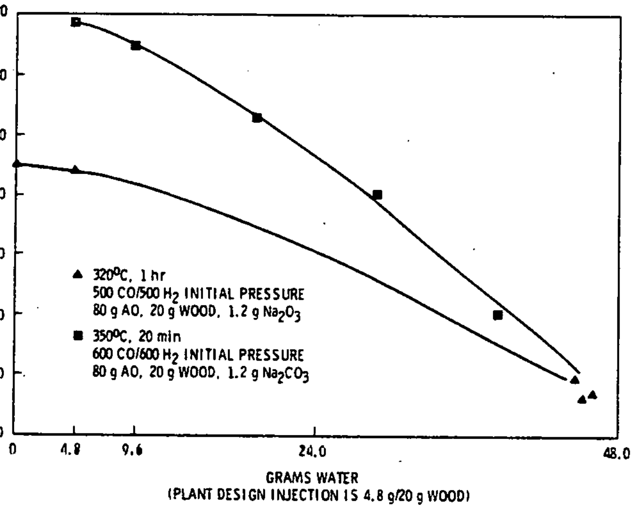

Fig. 3. Relationship of Unconverted Solids to the Amount of Water in the System

ble 2 lists those catalysts tested thus far.
ieir relative activities, as judged by the
rcent of acetic acid insoluble residue in the
oduct and by the amount of CO consumed in the
action, seem to place the catalysts into three
neral groups. Group 1 includes most sodium and
tassium compounds tested and these appear to
ve about the same activity as sodium carbonate.
oup 2 includes the lithium and magnesium com-
unds. This group has generally less activity
t still some catalytic ability. This group
ffers from Group 1 in that CO is generated with
these catalysts instead of consumed. Group 3 contains the remaining alkaline earth compounds which generate more CO and more acetic acid insoluble solids. The justification for using insoluble solids and CO consumption as tests of catalytic activity is based on the chemistry of the system. Acetic acid should be an adequate solvent for the oxygen containing oils produced from wood in this reaction, whereas the unreacted wood itself and high molecular weight chars would be insoluble. Carbon monoxide consumption provides an additional measure of the relative amount of oxygen removed from the wood as carbon dioxide, considering that the water-gas shift reaction proceeds to a different degree with each catalyst as measured approximately by the amount of hydrogen produced. We also note that the activity of the catalyst generally falls in line with its solubility in water. The sodium borate is an anomaly in the data field and this is due to the formation of borate esters with the cellulose, which in turn leads to dehydration and char formation. (This reaction would also be expected with phosphate catalysts.) To more effectively analyze the alkali metal catalysts, reduced concentration experiments are being performed to attempt to differentiate between the catalysts' effects.

Autoclave experiments to test the effect of catalyst concentration were also performed. Since the earlier work showed that no catalyst was significantly more active than the sodium carbonate which is presently used at Albany; it was decided to proceed with that catalyst for these tests. Catalyst concentrations, as a percent of organic feed, ranged from 0 to 18 percent. The results, as shown graphically in

TABLE 2. BIOMASS LIQUEFACTION WITH DIFFERENT CATALYSTS

Catalyst	Grams Catalyst	Grams Water	Solubility in Water g/100cc at 100°C	Percent Solids	Grams CO Consumed (Generated)	Moles CO Consumed (Generated)	Moles H_2 Produced
Na_2CO_3	1.2	4.9	46	0.69	0.9	.032	.013
$NaHCO_3$	2.0	4.9	16(1)	0.56	1.4	.050	.054
NaO_2CH	1.6	4.8	160	0.55	0.4	.014	.020
NaOH	0.7	3.6	347	0.67	0.6	.021	.036
Trona (dried) $Na_2CO_3 \cdot NaHCO_3$	1.4	4.8	—	0.85	—	—	—
K_2CO_3	1.6	4.8	156	0.51	1.7	.061	.056
$KHCO_3$	2.3	4.8	60(1)	0.69	—	—	—
KO_2CH	1.9	4.8	657(1)	0.64	1.6	.057	.052
KOH	1.3	4.8	18	0.96	—	—	—
$Na_2SiO_3 \cdot 9H_2O$	3.2	3.1	5	0.50	0.5	.018	.013
Li_2CO_3	0.8	4.8	0.7	0.84	(0.5)	(.018)	.007
$LiO_2CH \cdot H_2O$	1.6	4.5	57	1.36	(0.5)	(.018)	(.002)
$LiOH \cdot H_2O$	1.0	4.8	18	0.97	(0.8)	(.029)	(.004)
$4MgCO_3 \cdot Mg(OH)_2 \cdot 4H_2O$	2.7	4.4	0.01	1.24	(0.9)	(.032)	(.023)
$Mg(O_2CH)_2 \cdot 2H_2O$	3.4	4.0	24	1.83	(1.6)	(.057)	(.003)
No catalyst	0	4.8	—	3.4	0.0	—	—
$CaCO_3$	2.3	4.9	0.0019	3.72	(1.2)	(.043)	.000
$Ca(O_2CH)_2$	3.0	4.9	18	3.93	(1.6)	(.057)	(.001)
$SrCO_3$	3.4	4.8	0.065	5.34	(1.5)	(.054)	(.008)
$BaCO_3$	4.5	4.8	0.0065	5.29	(1.7)	(.061)	(.011)
$Na_2B_4O_7 \cdot 10H_2O$	4.3	7.8	170	14.41	0.8	.029	(.004)

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(1) Solubility at 60°

Figure 4, indicate that with a constant water feed the increase in the amount of sodium carbonate catalyst from the plant design specification of 6 percent up to as much as 18 percent has only a very small effect on increasing the wood conversion to acetic acid soluble product. Interpolation of the data indicates that the catalyst concentration may be decreased to as low as 3 percent before noticeable decrease in activity occurs.

Another result of these experiments was that the effect of decreasing the water content of the slurry coupled with the decrease in catalyst has a very marked negative effect on wood conversion, whereas an increase in water content coupled with an increase in catalyst concentration has a much smaller positive effect on wood conversion. This corroborates our earlier data on the effect of water in the process.

Another set of autoclave experiments was performed to test the effect of the reaction conditions on the catalyst. Sodium formate and sodium carbonate were tested in a water solution with the high pressure CO/H_2 atmosphere. The formate was heated to 350°C for thirty minutes and upon cooling was found to be converted for the most part to sodium bicarbonate. Both of these compounds decompose to sodium hydroxide below 350°C, and this test shows that the bicarbonate is the more stable species upon cooling even in an atmosphere (8 to 1, CO to CO_2) favoring the formation of the formate. A similar test performed with sodium carbonate gave evidence of very little water/gas shift reaction under these conditions. No amount of formate, bicarbonate, or hydroxide was detectable in the infrared spectrograph of the recovered product. These two experiments give evidence against the formate mechanism put forth by earlier experimenters and, in fact, conflict with some of thei other published statements to the effect that formates are synthesized in good yield by treating alkali carbonates with water and CO under th reaction conditions of the oil producing experiments. We conclude that the oil producing mechanism is the result of a combination of basic

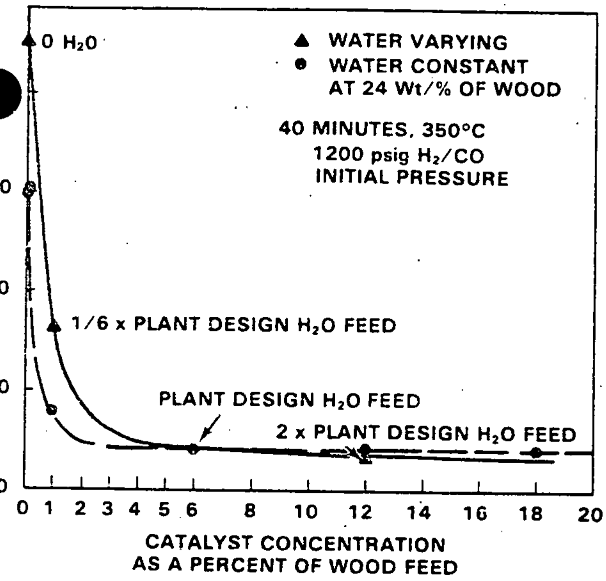

Fig. 4. Relationship of Unconverted Solids to the Catalyst Concentration

alyzed reactions (including neutralization of anic acids produced in the reaction) with mall effect possibly due to the hydrogen duced by the water/gas shift reaction.

rogen Donor Solvent Effects

resent thrust of the process development work ls with the possibility of using the hydrogen the cover gas via a hydrogen donor system rein the recycle oil (wood oil) would be hy- genated in a side stream and reintroduced to reactor to act as a hydrogen donor solvent. important questions exist with such a pro- ure: 1) will wood (or wood products) accept rogen from a hydrogen donor solvent under the ditions of the Albany system? and 2) can pro- t oil be hydrogenated and transformed into a rogen donor solvent? The present work is ed at the first question and, to that end, oclave experiments have been performed with rry oils of different hydrogen donor capabil- as substitutes for the presently used anthra- e oil. Initial experiments showed a slight rease in wood conversion which correlated with rogen donor capability, but only a small por- n of the hydrogen from the hydroaromatic por- n of the solvent was being transferred from solvent without replacement.

eries of solvents were then tested to deter- e the correlation between hydrogen donor capa- ity and the conversion and oil yield in the any reaction system with H_2/CO gas and O_3 catalyst at 350°C. A good correlation ween conversion and H-donor strength is shown the series listed in Table 3. Theory predicts the first three slurry vehicles would be hydrogen donor solvents with decreasing capability going down the table. The last four solvents would not be expected to have much, if any, hydrogen donor capability.

A second series of experiments was performed with hydrogen only and no catalyst to test the hydrogen acceptance of wood from a donor solvent (tetralin). As shown by these results in Table 4, a good correlation can be seen between time/temperature parameters and the conversion of wood and the conversion of tetralin to naphthalene. The point which is important relative to this project is that the conversion with H_2 alone is not as good as with the H_2/CO mixture with sodium carbonate catalyst.

These results support two conclusions--first, this is additional evidence for the superior reactivity of the CO-alkali-water system as opposed to hydrogen alone for converting wood from the solid form as shown earlier by PNL, and others; and second, the application of a hydrogen donor solvent to the CO-alkali-water reactant system with hydrogen present has a beneficial effect on

TABLE 3. CONVERSION OF WOOD IN DIFFERENT SOLVENTS

	Conversion	Oil Yield
Dihydronaphthalene	97.0	40.6
Tetrahydronaphthalene	92.6	22.2
Coal tar distillate	90.0	NA
Naphthalene (40 minutes)	87	NA
Heavy petroleum oil	85.7	NA
Dodecane (40 minutes)	82	NA
Decahydronaphthalene	78.2	5.0

Expt. at 60 minutes, 350°C, H_2/CO 1200 psig initial pressure.

TABLE 4. HYDROGEN DONOR EXPERIMENTS

	Wood Conversion	Oil Yield	Tetralin Conversion
Tetralin, H_2, 400°C, 60 min.	91.3	22.9	8.0%
Tetralin, H_2, 350°C, 60 min.	85.0	25.8	2.9%
Tetralin, H_2, 350°C, 20 min.	89.7	28.7	1.7%
Coal Tar, H_2, 350°C, 40 min.	73.8	34.6	NA

conversion of wood from the solid form as measured by benzene solubility.

Alternate Feedstocks

Additional tests were made to examine alternate feedstocks for use in the Albany plant. The biomass feed presently used at Albany is -60 mesh wood flour made from dried and hammer-milled Douglas Fir wood chips. Feedstocks examined include both actual biomass feedstocks which could be used in the Albany unit and also biomass components such as cellulose and lignin. The combined results of all these tests should help us develop a model to predict results obtainable with other potential biomass feedstocks. A listing of partial results is given in Table 5.

Two feedstock factors which appear important to biomass conversion in the CO-Steam oil slurry process are the compactness of the structure as measured by crystallinity, wood structure, and particle size; and the composition of the feed relative to the amounts of cellulose and lignin. By examining the table the reader can see that conversion drops as the ratio of crystalline to amorphous structure increases from the powdered celluloses, to microcrystalline cellulose, to the two starches to the two sugars. Particle size and compactness of structure are also important as shown by the relative reactivity of the three forms of blotter paper. The lignin and sodium ligno-sulfonate exhibit low reactivity. Ryegrass, which is low in lignin, exhibited higher reactivity than the Douglas Fir wood flour currently used at Albany. The oak and mesquite are both relatively lower in lignin than Fir but showed slightly lower conversion; this lower conversion may be due to the larger particle size. It may be that the compactness problem can be overcome by the use of more water in the system. Water would serve to swell the biomass structure, as well as allow catalyst penetration into the feed material. The Douglas Fir wood flour's higher reactivity relative to the quantities of cellulose and lignin present in the wood has several possible explanations. The lignin which we tested was produced by a decompression technique wherein the wood is "pop corned" in a stean

TABLE 5. FEEDSTOCK COMPARISON

Feedstock	Acetic Acid Insoluble	Calculated Conversion	Mole CO Consumed	Mole H_2 Produced	Mole CO_2 Produced	Mole HC Producec
α - cellulose	1.43	93	.110	.109	.227	.025
Pulverized chemical pulp, extracted	-	92	.089	.045	.168	.001
Micro-crystalline cellulose	2.25	89	.091	.010	.130	.004
Milled blotter(2) paper	3.37	83	.068	.033	.149	.002
1/8" squares sheared blotter paper	3.59	82	.048	.067	.156	.113
1/2" squares torn blotter paper	4.92	75	.122	.150	.230	.104
Cornstarch	2.64	87	.056	.050	.120	.002
Potato starch	2.91	85	.041	.042	.121	.002
Dextrose(2)	7.47	63	.056	.034	.151	.002
Sucrose(1)(3)	7.77	61	-	-	-	-
Ryegrass	1.31	93	.113	.076	.183	.002
Douglas Fir wood flour	2.28	89	.082	.111	.173	.023
Ground mesquite(2)(3)	2.69	87	-	-	-	-
Hardwood sawdust	3.55	82	.096	.064	.145	.004
Sodium ligno-sulfonate(3)	6.44	68	-	-	-	-
Poplar lignin (370°C)	7.69	62	.126	.039	.165	.053
Poplar lignin, extracted	-	47.9	.104	.018	.110	.010

20 g feedstock, reaction at 350°C, 1 hr., 1200 psig initial CO/H_2:60/40

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(1) 20 minutes at 350°C

(2) N_2 included in initial pressurization

(3) Autoclave leak, gas analysis inaccurate

tmosphere and the lignin is then extracted by thanol. The resulting lignin is in the form of rospheres which are likely less hydrophilic n wood flour. They may preferentially isolate hemselves in the oil vehicle away from the water oluble catalyst. Their sphereical shape would lso allow less surface area for chemical attack han the fibrous wood flour.

hermolysis

rogress in this area has advanced to the stage f experimental reactor fabrication. Future ex-erimental work will be aimed at optimizing oil uality as measured by clarity, heating value, nd viscosity by varying process parameters such s residence time, temperature and solvent omponents.

FUTURE WORK

CO-Steam process development at the Albany site is scheduled to continue at least through June, 1980. Bench scale support of that effort should be an ongoing activity to provide input for planning of the more cost intensive facility operations. In addition, new processes should be screened at the bench scale for possible future development at the Albany facility. Specific plans call for continued investigation into the intricacies of operating CO-Steam biomass conversion as a continuous operation. Utilization of the hydrogen in the synthesis gas as a catalyzed reducing agent will be evaluated. Additional process options including high temperature injection of the feedstock and fractional distillation for product recovery will be investigated as dictated by time, money and sponsor interest.

NOTES

THE POTENTIAL FOR BIOMASS LIQUEFACTION

Manu Seth and Sabri Ergun
Lawrence Berkeley Laboratory
University of California
Berkeley, California

TRACT

irst, broad, overview of biomass liquefaction presented. Four desirable chemical conversion tes that may be useful in the liquefaction of d have been identified. Process conceptualiza- n has been attempted based on an analysis of nges in physical structure, elemental composi- n and chemical transformation that may occur ing liquefaction. Possible process streams e been characterized and likely separation cedures identified.

RODUCTION

technical monitors for the Thermochemical Con- sion of Biomass, a major responsibility of the l and Biomass Group at LBL is to identify and ine promising research and developmental areas ated to the production of liquid fuels from mass. As a first step a framework for techno- nomic evaluation of developing liquefaction ions is being formulated. This paper outlines objectives, scope, approach and consequences our first overview of biomass liquefaction. sequent papers will attempt to fill in the ails and discuss experimental results and how se fit into or modify the evaluation framework.

OVERVIEW

Several biomass feed-stocks are available for conversion to desirable fuels. These feedstocks in turn may be used to obtain a broad mix of products. Our objective, then, is to look for and develop a family of conversion processes. An integrated approach would enable us to apply experience gained with one feedstock or process-concept to other biomass feeds and conversion options. The developmental scheme is well represented by Figure 1, which outlines some of the feed-stocks and products of interest.

Each biomass liquefaction scheme can be characterized by the feed being processed, the products produced and the severity of the treatment. Apart from their ability to handle various feedstocks, the processes selected for development must also span a wide range of processing conditions. High severity processes, being capital intensive, could be used for large scale production wehreas low severity processes could be used for dispersed, small-scale applications. As a first approximation, the severity of a process can be defined by the reaction temperature and residence time in the major conversion step. Figure 2 shows a hypothetical temperature-time region which future biomass conversion processes must attempt to span.

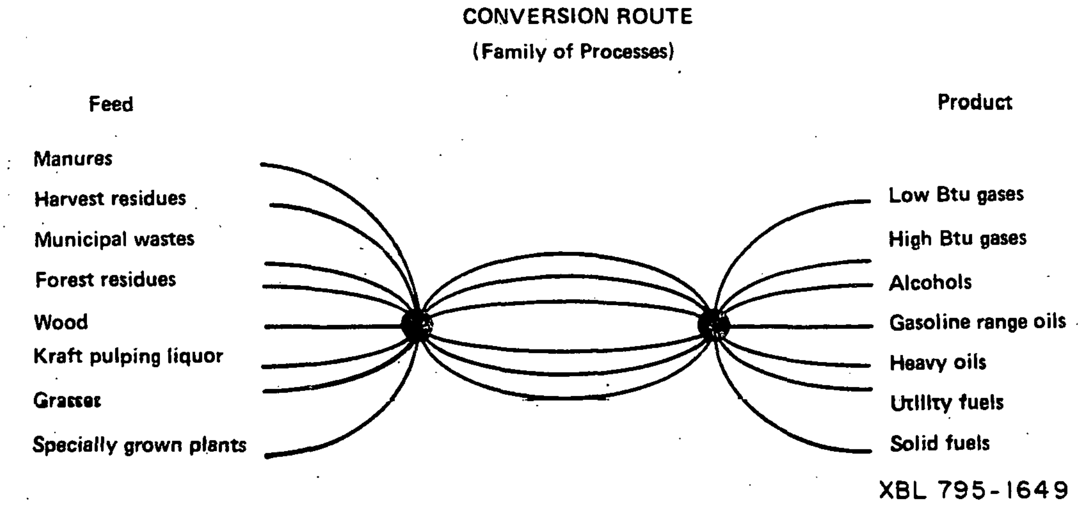
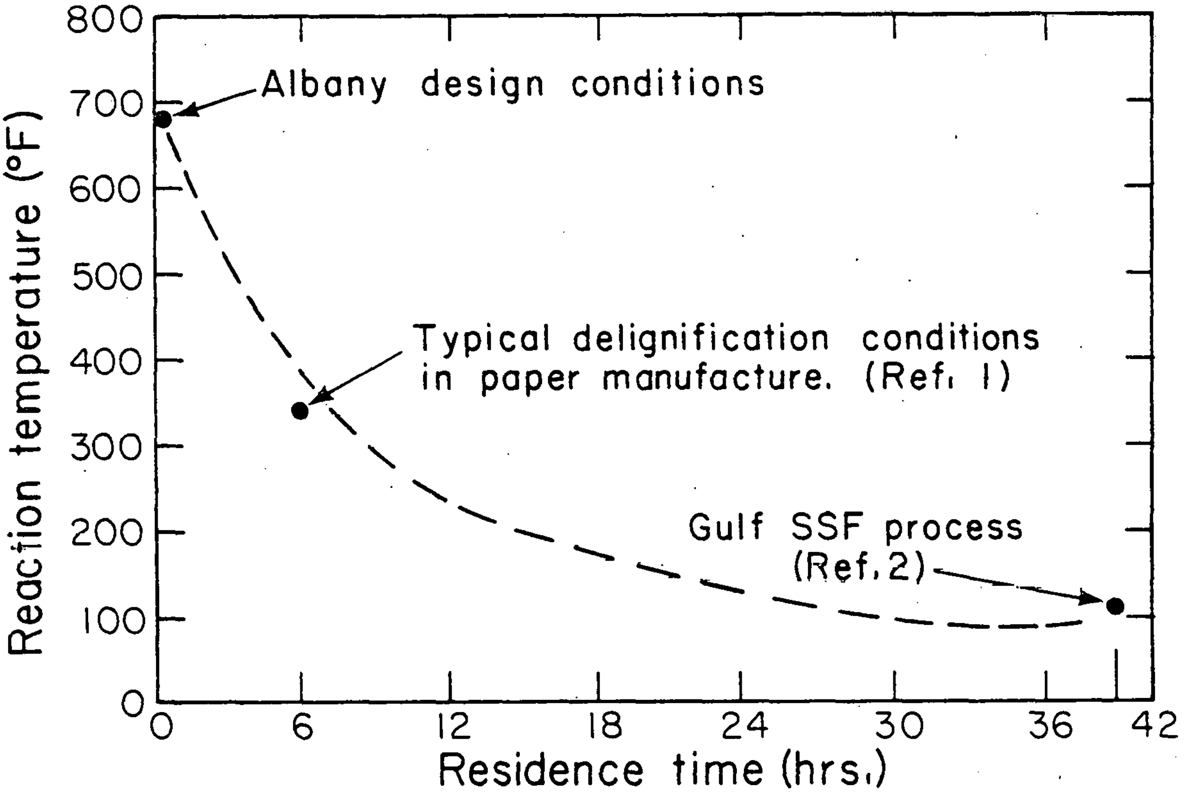


Fig. 1: Possible biomass feeds and products.



XBL 795-1650

Fig. 2: Hypothetical temperature-time region for future biomass processes.

The conversion of solid biomass, especially wood, can be viewed on three levels. The physical state, elemental composition and chemical structure of the biomass feed all change simultaneously as the material moves through a process train. Analysis of each of these transformations and identification of desirable changes leads to an approximate definition of process boundaries and flow schemes. For the sake of simplicity and clarity their discussion is limited to a single biomass feed-stock --wood.

CHEMICAL STRUCTURE OF WOOD

Wood is composed of three major components, cellulose (30-50 wt-%), hemicellulose (10-35 wt-%) and lignin (15-35 wt-%). The chemistry of wood can for the most part be described by the chemistry of its constituents.

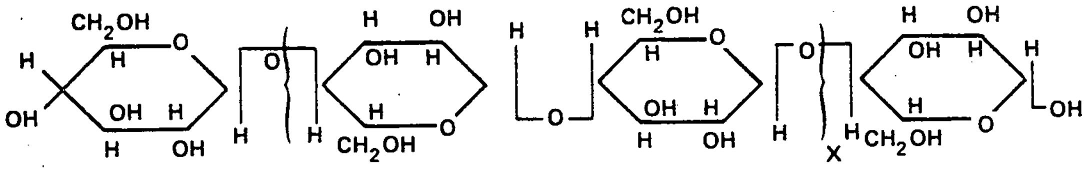
Cellulose is formed from D-glucose blocks joined by β-1,4-glucosidic bonds. Wood cellulose occurs as polymeric molecules with molecular weights generally in the range of 80,000 to 340,000. The structure of cellulose is shown in Figure 3.

Hemicelluloses are complex molecular chains of xylose or arabinose backbones. Xylans combined with substantial amounts of uronic acids are the most important hemicelluloses in wood.

Lignin is a polymeric substance whose complexity results from the variety of ways in which constituent phenyl propane and other building blocks can be linked. Several models have been proposed for the structure of lignin. One such model, by Freudenberg [3] is shown in Figure 4. The molecular weight of lignin polymers in wood ranges from 8,000 to 11,000 (or higher) corresponding to a degree of polymerization of 35 to 55.

WOOD CHEMISTRY AND ITS APPLICATION TO LIQUEFACTION

Several areas of wood chemistry have been well explored. A variety of chemical conversions were tested with a view to elucidate the chemical structure of cellulose, lignin and hemicellulose [1,3, 4,5]. Pulping of wood for the manufacture of pape is probably the most studied and best understood area of wood chemistry [1,6]. Hydrogenation of waste liquor from paper manufacturing processes has also been well investigated [4,7,8]. The hydrogenation and hydrogenolysis of wood and lign were extensively studied by Lindblad [9], Harris [10], Lautsch [11] and Hachihama [12], among othe Substantial information also exists on the selective hydrogenation of carbohydrate materials such as cellulose, sugars and polyhydric alcohols [e.g. 13, 14].



XBL 795-1647

Fig. 3: Structure of cellulose.

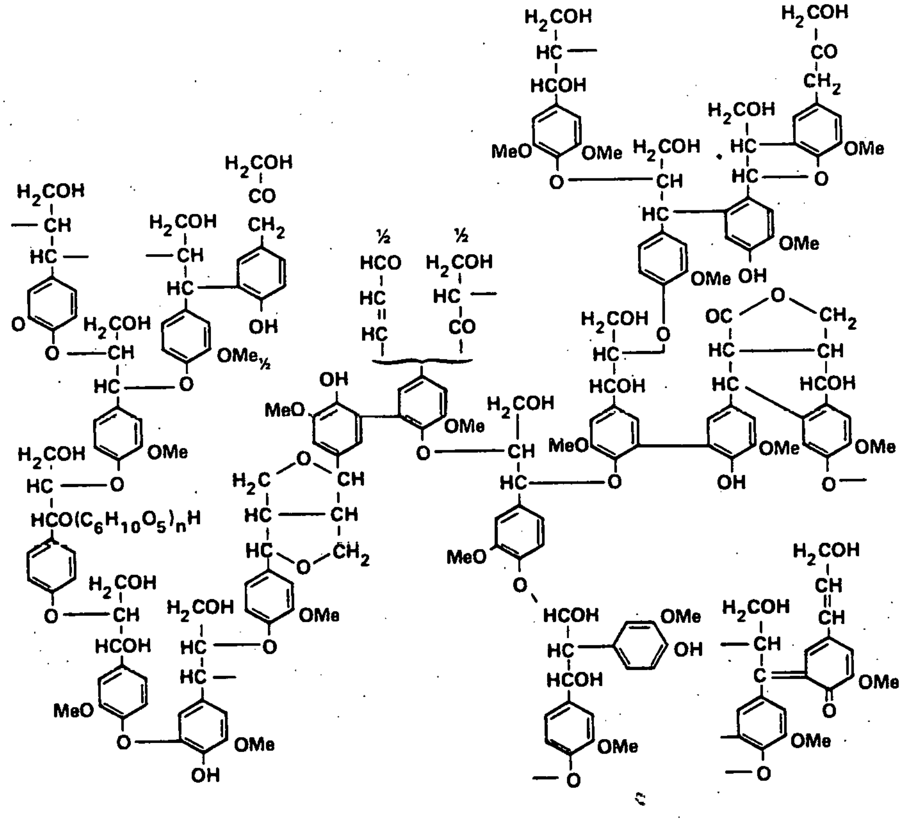


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Fig. 4: Structure of lignin.

Review of the various methods of the degradation and hydrogenolysis of wood and its components lead to the selection of four major chemical routes for liquefaction. Criteria used for the selection included acceptable process conditions and the possibility for pilot-scale testing at the DOE Experimental Facility at Albany, Oregon.

The four selected chemical conversion routes are briefly discussed below, together with illustrative examples of processing conditions, reagents and possible applications to wood liquefaction.

TREATMENT WITH INORGANIC SALT SOLUTIONS

In both the Soda and Kraft processes for the production of cellulosic pulp for paper manufacture, the lignin in wood is degraded by the cooking chemicals and its fragments dissolve in the liquor. Typically cooking times of 4 to 6 hours at 170°C are used for delignification. Several complex reactions occur during the delignification process. It has been suggested [15,16] that aromatic ether links in lignin are hydrolyzed by the alkaline cooking liquors. Etherification of specific phenolic groups are postulated to result in rapid cleavage of some of the ether groups [15], resulting in the solubilization of about 1/3 of the lignin. The presence of sodium hydrosulfide ions ($-SH^-$) in Kraft cooking liquors are assumed to cause the increased extent of lignin solubilization in this process [15]. Model experiments indicate the initial step is addition of $-SH^-$ ions to quinone methide lignin intermediates to form mercaptide ions (S^-). The mercaptide ion is a strong nucleophile, and by a complex series of rearrangemets leads to the depolymerization of virtually all the lignin [15].

Modifications of the chemistry of paper manufacture could lead to new process schemes for wood liquefaction. Research in this area must be geared to overcoming three drawbacks. Other inorganic nucleophiles should be screened to find salts that i) do not result in sulfur (or nitrogen) incorporation in the products; ii) substantially depolymerize cellulose; iii) result in products which either have no inorganic base incorporation or products from which incorporated inorganic materials can be easily removed.

Extensive screening of various catalysts by the LBL group has resulted in the identification of an acceptable catalyst. These results will be reported when testing is completed.

Mild oxidation of wood prior to treatment with alkali may also result in the simultaneous depolymerization of cellulose and lignin at 170-180°C [15]. This option is also being tested at LBL.

Hydrogen Transfer from Hydrocarbon Molecules

Hydrogenation of wood to distillable products can be achieved by treatment with hydrogen donor solvents at 320 to 400°C for 1 to 4 hrs [17]. Cyclohexanol [18], tetralin [17] and decalin [19] hav all been tested as sources of hydrogen. Liquefa tion occurs by the donation of hydrogen from solvent molecules to thermally cleaved bonds in wood. An alternative approach for the transfer of hydrogen to wood is the acid catalyzed hydride transfer reaction. In this reaction, a hydrogen atom is transferred with its pair of electrons to a carbonium ion [20]. α-Pinene has been used as a hydride donor in the presence of a strong acid at 120-150°C [21].

Both thermal hydrogen donation and catalytic hydride transfer hold promise for wood liquefaction. In each case the spent solvent would have to be hydrogenated preferably in a separate hydrogenation reactor. Two possible problems may occur with either scheme. First, incorporation of solvent by condensation reactions with products or with wood residues could lead to unacceptably high solvent losses. Second, separation of solvent fro reaction products by distillation may be difficult because of overlapping boiling ranges. To circumvent both problems wood-derived molecules should be tested as hydrogen donors or hydride transfer agents. Partially hydrogenated phenyl propane units (derived from lignin hydrogenaration) and/or alcohols (from cellulose hydrogenation) may prove to be adequate hydrogen sources.

Solvolysis

Solvolytic degradation of cellulose and of lignin have both been well studied [7,22]. Solvolysis in an acidic medium can occur under a variety of conditions ranging from 0.8 wt-% acid at 170°C to 40% acid at 40-60°C, with solvents such as water (hydrolysis), methanol, ethanol and even phenol [8]

Solvolysis of cellulose is an acid catalyzed reaction involving the rapid formation of an intermediate complex between the glycosidic oxygen and a proton; this is followed by the slow, rate-determining scission of the glycosidic bond adjacent to $C_{(1)}$ [22]. When lignin is treated with alcohols in an acidic medium new alkoxy groups are introduced at ambient conditions [7] without significant lignin depolymerization. At higher temperatures (e.g., 79°C with ethanol) this rapid hydroxyl displacement reaction is followed by degradation of lignin to monomers and soluble oligomers.

By carefully controlling reaction temperature and acid concentrations it may be possible to liquefy wood by simultaneous solvolysis of lignin and cellulose, using acceptably small levels of solven consumption. If low boiling alcohols are used for solvolysis the overall process must also include an alcohol recovery or production unit.

Organometallic Complexes as Liquefaction Catalysts

The application of homogeneous organometallic complexes for hydrogenation and hydroformulation reactions are well known. Such catalysts app hold considerable promise for wood liquefaction. ently Gaslini [23] and Nahum [24] reported the f dicobalt octacarbonyl as an effective alyst for the delignification of wood. Their jective was to produce a cellulosic pulp suit- le for the manufacture of paper. Red spruce od meal was suspended in a polar solvent and eated with 1:1 hydrogen/carbon monoxide at 0-170°C in the presence of dicobalt octacarbonyl, soluble catalyst. Tests were run for residence imes ranging from 6 to 24 hrs [23,24]. Under ese conditions about 97% of the lignin was lubilized. Extensive characterization of the roducts indicated the formation of substituted aiacols and guiacyl tetrahydrofurans, with up 70% of the soluble product boiling under 400°C. e cellulose and hemicellulose fractions of the od were essentially unchanged.

effectively utilize homogeneous organometallic mplexes (especially carbonyls) for wood liquefac- ion, two changes in the above scheme must become ssible. First, the catalysts selected must be luble in dilute aqueous acids (pH ≅ 2-4). The cid would hydrolyze the wood to soluble sugars, hich in turn may also be hydrogenated. The ctivity of catalysts must also be improved so as o achieve hydrogenation rates comparable to those f sugar degradation.

aving selected potentially attractive process hemistries we now turn our attention to process onceptualization.

ROCESS DEFINITION

rocess development involves establishing of hemical, technical and economic feasibility of a rocess scheme, usually in that order. When an rray of partially understood chemical conversion chemes exist, even a limited prior understanding f process parameters and economics can be of nvaluable assistance in formulating a reasonable xperimental program. The changes in physical tructure and elemental composition of wood as it is onverted to liquid products have been analyzed or a first overview of process concepts.

hysical Transformations

ood may be fed to a process as chips (3/8" to 1" ize pieces) or as flour (typically -60 mesh owder). Use of wood chips avoids the cost of rying, crushing and grinding associated with the roduction of flour. Chemical (or mechanical) egradation of chips to fine particles is, however, ecessary if the wood is to be fed as a slurry through high pressure pumps to pressurized reaction tanks. In the first lower pressure chemical conversion step wood chips must be transformed into a pumpable slurry and/or material soluble in the slurrying solvent. Early depolymerization of wood to soluble material is also chemically advantageous because of greater accessibility of the substrate due to reduced mass transfer limitations. Furthermore, cellulose must be depolymerized in the first conversion step, since delignification alone would result in the formation of fibrous cellulose which would still be difficult to pump.

The next major physical transformation would be the conversion of the pumpable slurry to low molecular weight product. This step could, if necessary, occur at high pressures.

In retrospect the chemical structures of major wood components involve a large number of functional groups containing oxygen. In the early stages of depolymerization a polar reaction medium could be expected to be beneficial because the oxygen-containing groups are quite polar.

Elemental Transformations

The overall mass balance for any biomass conversion scheme can be represented by:

$$C_xH_yO_z \rightarrow a\ H_2 + b\ CO_2 + c\ CO + d\ C_pH_qO_r$$

(feed) (product)

This equation is a plant-battery-limit mass balance that ignores ash, nitrogen and sulfur but includes the production of any reducing gases or wood derived reagents needed for the overall conversion (e.g., CO, H_2 or alcohols).

An essential element for wood liquefaction is the removal of nearly all the chemically bound oxygen which, together with depolymerization, can lead to formation of liquid products with a high heating value. Oxygen removed from wood would exit any processing scheme as water, carbon dioxide or carbon monoxide. Removal or addition of water to or from an organic molecule results in little change in its molar heating value, as indicated by the Dulong Formula for estimation of heating values. Spontaneous removal of carbon dioxide in the major reaction step would lower the consumption of reducing gases and hence the load on any auxiliary gasification system. Better carbon utilization would be obtained if an increased fraction of CO_2 is produced since two oxygen atoms are removed for every carbon (instead of one for one in a CO molecule).

If hydrogen, carbon monoxide or alcohols are used for depolymerization and/or deoxygenation of wood the extent to which they are consumed will affect the total wood consumption per unit of oil. Furthermore, the ability of a process to accept mixtures of CO and H_2 could result in a substantial saving in overall cost by eliminating the need for cryogenic gas separation.

Separation Processes

Separation and recovery of products from reactor effluents could be an important (and expensive) part of biomass liquefaction processes. Typical intermediates and products obtained from the depolymerization of cellulose and lignin in reducing atmospheres were examined so as to obtain some understanding of the unit operations that could be used for product separation and recovery. A partial list of cellulose and lignin decomposition products is shown in Figures 5 and 6, together with their estimated boiling points and solubilities in water.

The depolymerization of wood in reducing atmospheres results in the formation of products of progressively lower oxygen content and polarity. In general the lignin in wood decomposes to form water-insoluble, high boiling products, whereas cellulose decomposition results in the formation of water-soluble products having lower boiling points. This is a rather simplified view of wood liquefaction since it ignores the possibility of the formation of polymeric materials from condensation and degradation reactions. It does, however, enable us to begin estimation of the physical properties of process streams and selection of possible separation processes.

	Example	Boiling point	Solubility in water
Cellulose			
Oligomers			Insoluble
Sugars	Glucose	–	Soluble
Polyalcohols	Sorbitol	295(@ 3.5mm Hg)	Soluble
	Glycerol	290	v. Soluble
Alcohols	Propanol	98	v. Soluble
	Ethanol	79	v. Soluble
Alkanes			

Fig. 5: Cellulose depolymerization products in reducing atmospheres.

A two-phase liquid effluent can be expected from the liquefaction reactor(s). The light, aqueous, phase would contain any added process water (and water from wood dehydration) together with dissolved low-molecular-weight phenols and alcohols. Higher molecular weight wood-derived oils and any non-polar organic slurring solvents could be expected in the heavy, organic, phase. The organic phase would also contain any unreacted wood or high-molecular-weight condensation and degradation products.

Where the aqueous phase is a large fraction of the product stream, decantation or centrifugal separation may be considered so as to avoid high evaporation costs. Such phase separation would also permit the recirculation of water soluble catalysts.

Of the organic phase only materials lower boiling than dilignols (from lignin) and sugars (from cellulose) can be distilled. Reactor conditions should therefore be optimized for the production of substituted phenyl propanes (monoliquids) and polyhydridic alcohols. The non-distillable fraction of the organic phase could be used for the production of reducing gases and/or process heat generation.

CONCLUSION

A two-step wood liquefaction process has been envisioned using wood chips as the feed. The first low-pressure step would be used to obtain

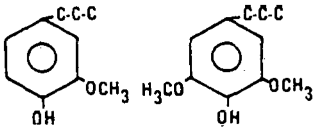
	Example	Boiling point (°C)	Solubility in water
Lignin			Insoluble
Oligomers			
Phenylpropanes		265-400	
	4-methoxy-2 methyl phenol	252	Insoluble
Substituted phenols and cresols	Cresolol Ethyl phenol	249 207	
Phenols	Phenol	182	Soluble
Cyclohexanols	Methyl cylohexanol	155	Insoluble
Cyclohexanes	Cyclohexane	81	Insoluble

Fig. 6: Lignin depolymerization products in reducing atmospheres.

pumpable slurry. The second step, which could r at higher pressures, would convert the ass to distillable oil.' Four possible chemi- l routes that could prove useful for wood quefaction have been identified. An analysis likely changes in elemental and physical com- sition during liquefaction resulted in a broad finition of process characteristics and the entification of desirable changes. Probable mponents of reactor effluent streams have also en identified and appropriate separation pro- dures suggested.

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NOTES

THERMOCHEMICAL CONVERSION OF BIOMASS TO GASOLINE

J. P. Diebold and G. D. Smith
Naval Weapons Center
China Lake, CA 93555

STRACT

s noncatalytic process involves the low pressure, ective pyrolysis of organic wastes to gases containing atively high amounts of ethylene and other olefins. er char, steam, and tars are removed at low pressure, gases are compressed to 450 psia (3100 kPa) for ification. The concentrated olefins are then further npressed to 750 psi (5200 kPa) and fed to the polyrization reactor where they react with each other to m larger molecules, 90% of which boil in the soline" range. Using organic feedstock derived from sh, gasoline was produced on a bench scale system ch had the same appearance and distillation characistics as gasoline made from pure ethylene which had unleaded motor octane of 90. Preliminary economic lyses indicate that the process is currently compeve with petroleum derived gasoline. This program has n funded by the U.S. Environmental Protection ncy.

RODUCTION

petroleum resources of the free world are becoming easingly scarce and expensive. As the search for new goes further out to sea and deeper in the ground, this oil will be increasingly expensive to produce. Many the existing oil fields in some countries apparently not be pumped at their maximum rate in an effort to that oil for future generations when the price will resumably higher. Thus, it appears advantageous to elop sources of synthetic petroleum and in particular hetic gasoline. Due to the high volumetric energy ent of hydrocarbon liquids, they would have been nted for mobile transportation if they had not been rally occurring. Usually the term "synthetic oleum" brings to mind coal and/or oil shale efaction. However, these processes will involve a iderable amount of environmentally questionable mining and are normally considered to require a e amount of scarce water resources. The process to iscussed in this paper involves the use of organic es and/or biomass in a noncatalytic selective process roduce unleaded, high octane gasoline. Byproducts fuel oil, lubricating oil, ash, water vapor, and carbon ide. Commercial sized plants would resemble small refineries in many aspects and could be widely oyed in order to be near the feedstock source.

advantage of producing gasoline rather than other s becomes apparent by comparing common fuels on a relative wholesale value basis. If a wholesale value of $1.00 per unit of energy is assigned to natural gas for interruptible commercial usage, then in 1975 the value of noninterruptible domestic natural gas was $1.42 per unit of energy. Number six fuel oil was $3.08, while the value of gasoline was $5.48 per unit of energy. Recent relative increases in the value of natural gas have reduced the ratio of gasoline to natural gas prices temporarily. However, the widely predicted continuation of today's petroleum shortage will cause the value of gasoline to escalate at a rate much greater than boiler fuels. This is because, although coal can again become the primary boiler fuel for the nation, coal fueled automobiles having today's performance are not being actively considered.

The conversion of organic waste to automotive fuel was consequently investigated for remote military installations in a Department of Defense Advanced Research Projects Agency sponsored program because of the high potential value of the product. The synthesis of methanol from organic wastes was evaluated in detail with mass and energy balances developed about two different flow diagrams. This approach would have pyrolyzed the organic fraction to form synthesis gases which would then have been compressed, purified, and catalytically reacted to form methanol. During the course of this study (1), several references were found which indicated the possibility of using pyrolysis to form a significant amount of low molecular weight hydrocarbons rather than just carbon oxides, methane, and hydrogen. These gaseous hydrocarbons of interest were predominantly ethylene, propylene, and other olefins.

In the 1930's the oil industry began to extensively pyrolyze ("crack") crude oil to increase the yield of gasoline. By-products of that process included large amounts of ethylene, propylene, and butylene. Extensive research was performed on the utilization of these by-product gases which led to the commercialization of their conversion to gasoline by both catalytic (2) and non-catalytic (3) processes. The gaseous hydrocarbons were compressed, purified, and then heated such that they reacted to link up with themselves to form the larger gasoline molecules by polymerization. This process produced a liquid which was over 75 percent gasoline. The proposed process substitutes solid organic wastes for the crude oil feedstock, but otherwise parallels the petroleum process used to make polymer gasoline. Due to the high ethylene content of the olefins formed during pyrolysis of cellulosic materials, the thermal or non-catalytic polymerization was chosen for this process because the traditional catalytic method polymerized ethylene with difficulty.

The overall process for converting organic wastes to gasoline consists of: first, the pyrolysis of the wastes to gases containing large amounts of olefins, i.e. ethylene, propylene, etc.; second, the compression and purification of the olefins; and third, the polymerization of the smaller olefins to form larger gasoline molecules. Overall this amounts to a process which removes the oxygen from the cellulosic wastes to produce a gasoline consisting of hydrocarbons. Most of the rejected oxygen is in the form of carbon dioxide.

The critical technology that this program needed to demonstrate was the pyrolysis of cellulosic organic waste materials into gases containing large amounts of olefins and their subsequent processability. If the selective pyrolysis could be verified, and demonstrated, then the remainder of the process to make gasoline was thought to be relatively straightforward due to industrial experience with similar processes. In effect, the process parallels that of an oil refinery with the most significant difference being the use of today's cellulosic waste material rather than eons-old organic matter (crude oil) as a feedstock.

The potential impact this process could have on the gasoline consumed in the United States appears to be substantial. Assuming that this process will undergo a traditional developmental or scale-up period followed by a well financed program with high national priority, widespread deployment of conversion plants could be accomplished by 1998. A recent Market Oriented Program Planning Study by the Department of Energy (DOE) estimated the U.S. gasoline consumption in 1990 at 115 billion gallons (4). If 10% of the land currently in forest, pasture, or range usage having at least 25 inches of precipitation and less than 30% slopes were to be developed for silvicultural energy farms, approximately 28% of the gasoline consumed could be produced from the resultant biomass (5). Another 16% of the projected gasoline consumption could be produced from crop residues and spoiled forages (6). If the trash generated by 120 million people were to be converted to gasoline, 4% of the 1990 gasoline consumption would result. From these three biomass sources, slightly less than half of the projected 1990 gasoline consumption could be produc by the conversion process described without impacti food production to any extent. To determine the net energy produced by the overall system including energy inputs to grow, harvest, transport, and process biomass into gasoline requires a very specific analysis for a given site and is outside the scope of this paper. However, the process would have the best economics where the biomass is a waste by-product and which now requires energy for its disposal.

Initial preliminary economic analyses looked very promising for fairly small cities to rid themselves of trash so the Environmental Protection Agency (IERL, Cinn.) was contacted to generate interest in this economical approach to trash disposal. As a result of their interest in this process, an interagency agreement (EPA-IAG-D5-0781) was written for NWC to pursue the technical and economic feasibility of converting organic wastes to gasoline by evaluation in a bench-scale unit (5 kg/hr). This EPA effort was begun in June 1975 and has been continuing.

EXPERIMENTAL

Pyrolysis

Figure 1 shows the currently evolved pyrolysis schematic. A finely ground organic fraction of municipal soli waste is continuously fed by a one-inch screw feeder. A the end of the screw feeder, the feed is fluidized an conveyed by a carrier gas stream (normally carbo dioxide) to the steam ejector. The mixture of soli waste, carrier gas, and steam then enter a long, red-ho 3/4 inch diameter tubular reactor. Reactor lengths of and 6 meters resulted in pyrolysis times of about 50 t 150 milliseconds, respectively. The char is remove from the pyrolysis stream in a three inch diamete cyclone. The steam and tars are condensed out of th pyrolysis gas stream by the water quench system. Th noncondensible gas stream volume is measured by a orifice flow meter and then either flared off or fed int a three stage compressor for storage at high pressure This system pyrolyzes a nominal 5 kg of feed per hour.

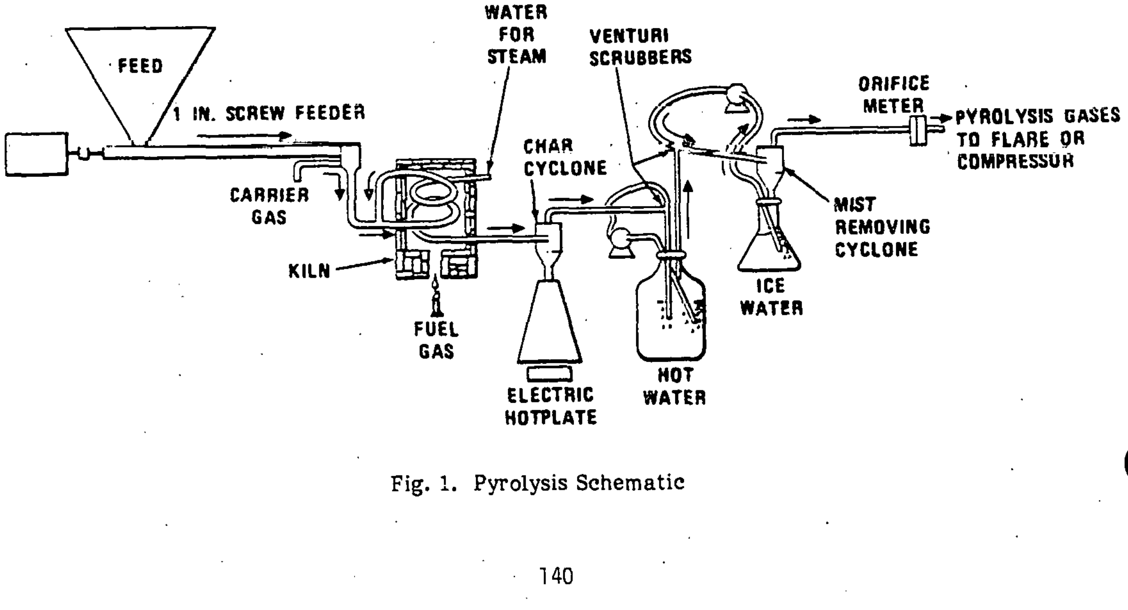

Fig. 1. Pyrolysis Schematic

ie composition of the pyrolysis products has been found be a function of the combination of temperature, nce time, and especially the dilution of the solid es by inert gases inside the reactor. As shown in gure 2, the total hydrocarbon product (other than ethane) can be varied by as much as 165% depending on the relative dilution of the pyrolysis gases. Using e more favorable combination of these variables, a rolysis product distribution such as that represented in gure 3 can be attained using organic material obtained om municipal solid wastes. Slightly more than half of e energy contained in the organic waste can be covered in the gasoline precursors. About one third of e energy is recovered in the "medium Btu" by-product s stream of carbon monoxide, methane, and hydrogen. out one-seventh of the energy is recovered in the ar, which has a heating value similar to lignite coal.

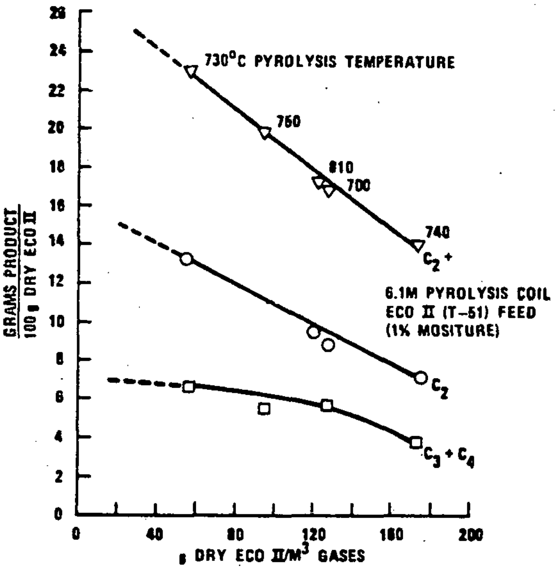

Fig. 2. Pyrolysis Products as a Function of Dilution

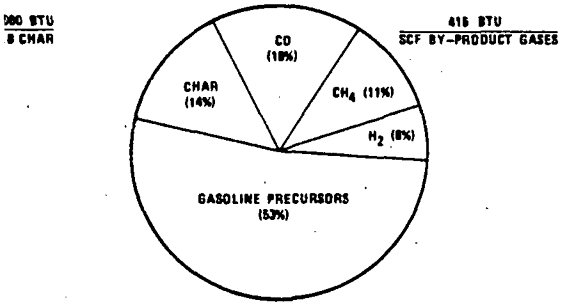

Fig. 3. Pyrolysis Product Energy Distribution

olysis Discussion

olysis of cellulosic materials has long been recognized the result of several competing reactions taking place ultaneously. At low temperatures of 200 to 300° C, the predominant reaction is of dehydration to form char and water vapor. At intermediate temperatures of 300 to 600° C the predominant reaction is a depolymerization or chain cleavage reaction to form levoglucosan tars. At elevated temperatures, gasification is the predominant reaction to form combustible volatiles. (7) This gasification reaction itself appears to have at least two major competing reactions: a) to form olefins, carbon monoxide, hydrogen, and methane; and b) to form thermally stable, water soluble, oxygenated compounds. The data suggest that the olefin forming reaction is favored when the partial pressure of the pyrolysis products is reduced by the addition of a relatively inert carrier gas, e.g. carbon dioxide, steam, methane, nitrogen, or carbon monoxide, or mixtures of these gases with hydrogen. It is interesting to note that although the quantity of gases produced can be varied by the relative dilution, the relative molar ratios of the gaseous species after water washing are relatively constant.

The by-product gases could be used to fuel stationary internal combustion engines to generate shaft power for the compressors and shredders, as well as to fuel the pyrolysis furnace along with the by-product char. As was mentioned, it was found that the use of steam to dilute the pyrolysis gases results in a higher conversion to gasoline precursors. It now appears that the process will be optimized by making as much steam as possible by energy recovery from hot gas streams as well as by utilizing all of the char and the by-product fuel gases for process energy. This will result in only gasoline, fuel oils, and lubricating oils as the products from the organic fraction. These products will be readily marketable compared to char, pyrolytic oils, or "medium Btu" gas. The generation of high pressure, superheated steam to cogenerate shaft energy and low pressure process steam may optimize the process.

It is interesting to note that the pyrolysis conditions found to be optimum for municipal solid wastes are coincidentally similar to those used by the petrochemical industry to pyrolyze naphtha or crude oil to form ethylene: e.g. about 750°C (~1400°F), steam dilution, pyrolysis times of less than a second, and long tubular reactors having inside diameters of less than five cm (8, 9, 10). It is even more interesting to note that if the carbon monoxide and carbon dioxide in the pyrolysis gas are ignored, and the remaining products are then normalized, that the relative weight percentages of the pyrolysis products from naphtha (8), solid wastes, and oil shale (11) are very similar, as shown in Figure 4. It appears that the molecular fragments from the pyrolysis of both cellulosic and large hydrocarbon molecules are very similar and that they achieve a similar short-lived "equilibrium" between the products. If not rapidly quenched, this psuedo "equilibrium" changes by the low-pressure, high temperature polymerization of ethylene, propylene, and butylene to form about half benzene and half aromatic tars with the evolution of hydrogen (12). If the pyrolysis system has a very long residence time, the benzene and tars further react to form char and hydrogen — the last mentioned being the traditional products from charcoal kilns, along with carbon monoxide.

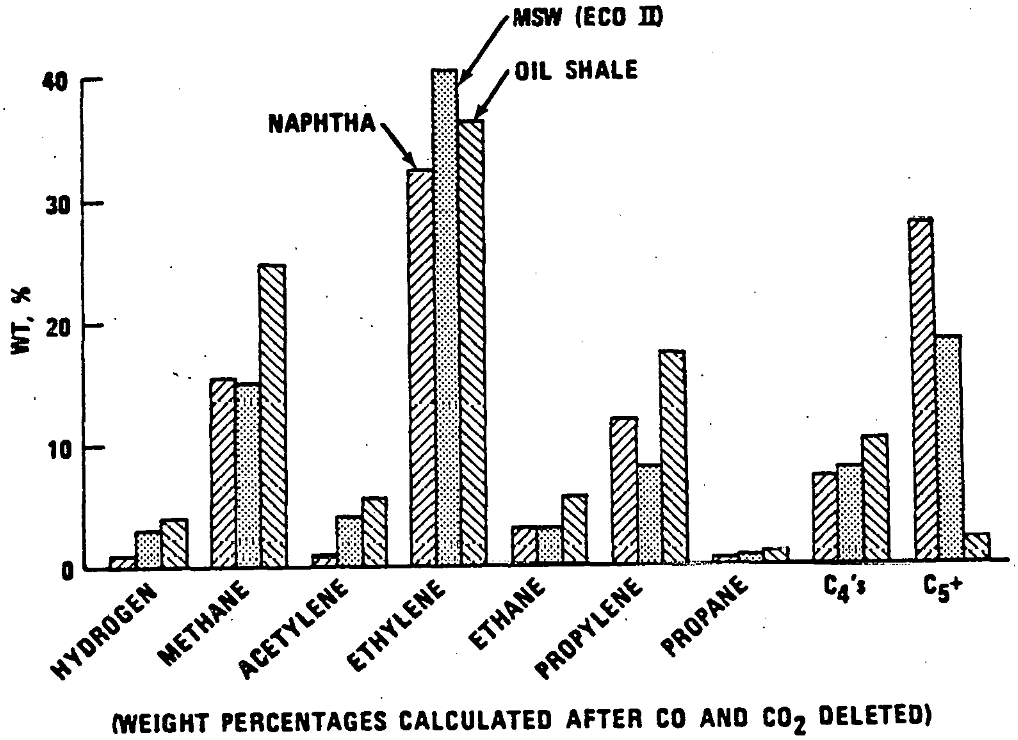

Fig. 4. Pyrolysis Product Comparison

Polymerization with Pure Ethylene

To avoid the tar formation during polymerization, lower temperatures can be employed so that the gasoline fraction in the product is optimized. However, in order to maintain reasonable reaction rates, the pressure must be increased. Temperatures of 400 to 500°C (750 to 950°F) and pressures of 4800 to 6900 kPa (700 to 1000 psi) were often employed commercially (13, 14). To gain insights into the exothermic polymerization reaction, an experimental reactor was constructed as shown in Figure 5. This polymerization effort was conducted concurrently with the pyrolysis development, so no purified pyrolysis gases at elevated pressures were available for use. Since ethylene was the largest single constituent in the gasoline precursors found in the pyrolysis gases, bottled ethylene was purchased and used for this study. The ethylene was regulated into a one-half inch diameter stainless steel tube which had been coiled and placed in a constant temperature, fluidized sand bath. The sand bath served to initially heat the coiled reactor until the exothermic reaction was initiated, after which the bath served to remove the heat generated. The ethylene slowly moved down the length of the reactor for a few minutes while polymerizing. The polymerized gases were then cooled to condense ou the gasoline and oils. Noncondensable gases were measured and then flared off. Conversions per pass were as high as 80 percent. The liquids formed had very low viscosities and when distilled produced about 90 percen

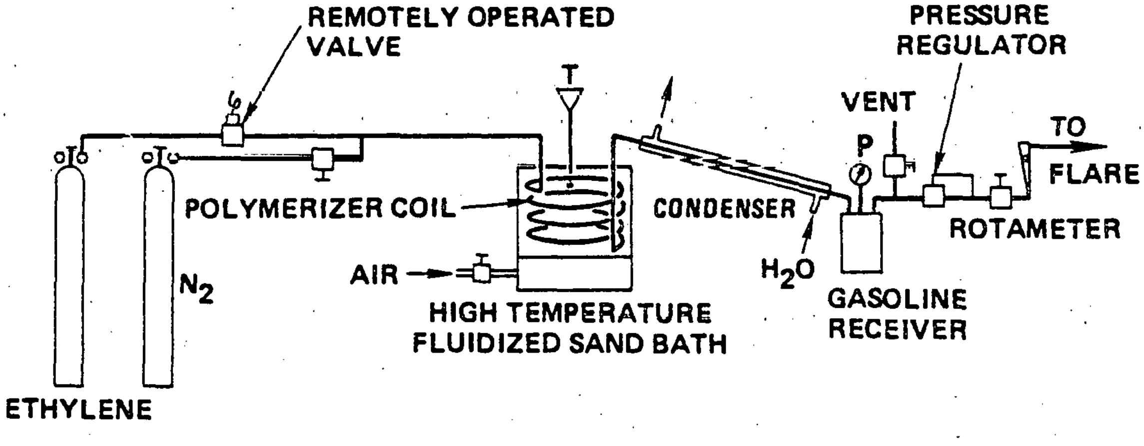

Fig. 5. Experimental Polymerization Reactor

soline. This distilled gasoline fraction was tested llowing the ASTM motor method and found to have a tane rating in the unleaded condition (the research od would have resulted in a slightly higher octane mber). The higher boiling liquids will yield a fuel oil d a lubricating oil fraction.

sed on the pyrolysis and ethylene polymerization perimentation it is estimated that 0.28 liter of gaso- e and 0.03 liter of oils will be produced per kilogram 90% by weight organic material (68 gal/ton gasoline, 8 l/ton oils). The amount of lubricating oil in the oils is rrently conjectured to be fairly small, but with a high tential total value due to its reported excellent cosity characteristics (15).

nch Scale Purification and Polymerization

e bench scale pyrolysis gas purification and poly- rization flow diagram is shown in Figure 6. The rolysis gases are first compressed to 3100 kPa (~450). The purification starts with the removal of carbon xide, hydrogen sulfide, and other water solubles with a t, aqueous solution of potassium carbonate (16). Next desirable olefins are absorbed in an organic solvent h the by-product fuel gases passing through absorbed. The olefin rich solvent is then heated to ve off the relatively volatile olefins (17). The purified fins are then further compressed to 5200 kPa (~750) and fed into the polymerizing reactor held at about)°C (~850°F). The polymerizing reactor consists of a eters long by 1.3 cm diameter 316 stainless steel tube ft by 1/2 in dia) immersed in a boiling sulfur bath. e boiling sulfur bath has an extremely high heat nsfer capability and serves to maintain a constant nperature as it removes the heat of polymerization. polymerization reaction proceeds under these ditions of heat and pressure without catalysts to form product consisting primarily of gasoline. After polymerization, the hot gasoline vapors are cooled, condensed, and stored. Unpolymerized olefins can be recycled.

This bench-scale system is in the final stages of debugging. The hot carbonate system for carbon dioxide removal is very selective and has resulted in a carbon dioxide stream having a purity of greater than 99% by volume with the primary contaminant being acetylene. The potassium carbonate solution is easily regenerated, although it does require a considerable amount of energy to boil the solution at 120°C (250°F). The hydrocarbon absorption system needs additional tuning, but has resulted in a byproduct gas stream containing 20% hydrogen, 5% nitrogen, 11% methane, 61% carbon monoxide, and 2 to 3% gasoline precursors (mostly ethylene and propylene). The hydrocarbon stripper system also needs some additional tuning, but it has resulted in an enriched stream containing as high as 48% gasoline precursors with 4% hydrogen, 1% carbon dioxide, 1% nitrogen, 16% methane, and 29% carbon monoxide as the impurities.

The effect of these impurities in the polymerization was to lower the partial pressure of the reacting olefins to slow the bimolecular polymerization reaction. A longer residence time is indicated in the polymerizer reactor depending upon the relative purity attainable from the gas purification system. Using pyrolysis gases made from organics derived from trash, a small quantity of synthetic crude oil was produced from the partially purified gases available. This synthetic crude oil was distilled and 93% of it boiled in the gasoline range. The distillation curve for the polymer gasoline made from pure ethylene was virtually identical to that for the gasoline made from trash derived organics. The physical appearance of the two synthetic gasolines is an identical very pale yellow and with a gasoline odor. Although sufficient gasoline made from pyrolysis gases has not yet

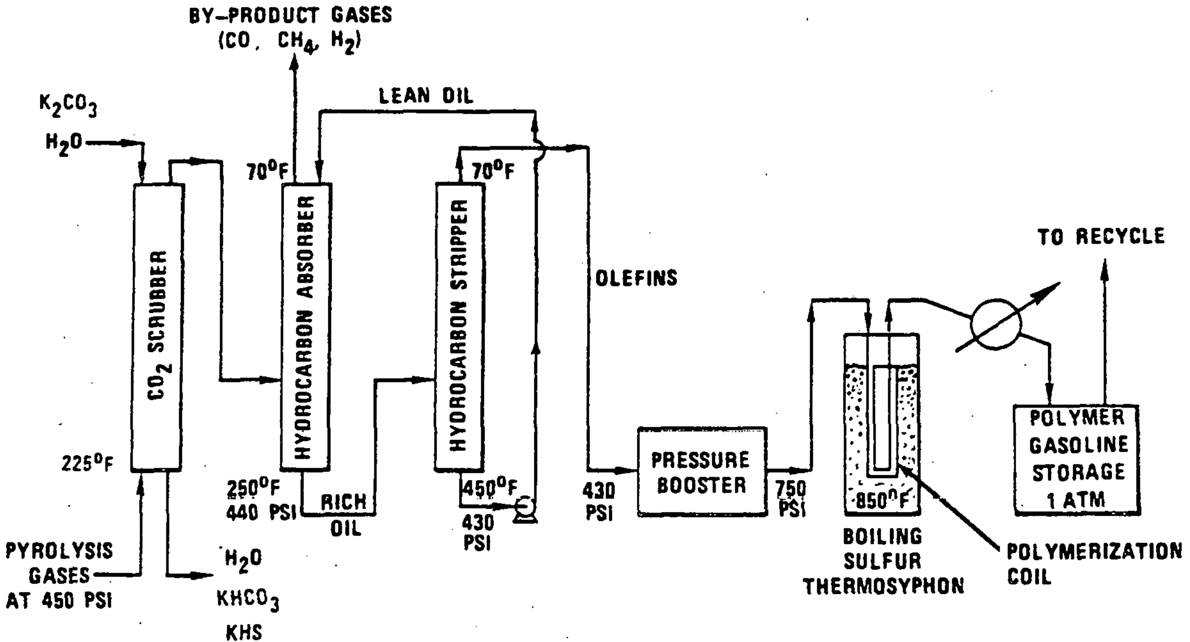

Fig. 6. Purification and Polymerization Schematic

been produced to allow an octane test to be run, it would appear reasonable to expect it to have an octane rating similar to that of the gasoline made from pure ethylene.

Additional effort needs to be expended in the bench scale system particularly in the areas of: low pressure pyrolysis gas scrubbing to more completely remove the tar mists; the hydrocarbon absorption and stripping system to verify that the reported high levels of gas purification can be attained with gases that contain carbon monoxide; the addition of a soaking chamber to the polymerizing reactor to increase the residence time to increase the conversion per pass; the addition of a recycle loop around the polymerizer to increase the overall yield; the full characterization of the synthetic gasoline and oil product; and the evaluation of feedstocks other than trash derived organics. A modest program to pyrolyze pure cellulose and pure lignin powders to be able to project biomass potential for this process has just been started with funding from the Solar Energy Research Institute.

COMMERCIALIZATION POTENTIAL

It was noted previously that the pyrolysis conditions used in development for organic waste are very similar to those used to pyrolyze crude oil and naphtha to ethylene. It follows that once the solid wastes are fluidized by the carrier gas stream, that the technology exists in the petrochemical and petroleum industry to design and build commercially-sized organic waste pyrolysis units. Since similar compression, purification, and polymerization of the gaseous hydrocarbons have all been commercialized in the past, it would appear that the process could be contracted to any one of several petrochemical or petroleum construction firms with a fairly low technical risk. For pilot plant demonstration purposes, it may be economically advantageous to add this gasoline module to an existing trash processing plant already in operation or to locate it in an area having several types of agricultural wastes available.

PRELIMINARY ECONOMICS

Municipal Waste Feedstock

To determine the economic feasibility of the process to convert municipal trash into gasoline requires a long list of assumptions. For the use of municipal trash as feedstock it was assumed to contain 60% dry, ash free organic material, 13% inorganics (iron, aluminum, and glass), and the balance moisture. A credit of $4.85 per tonne ($4.40 per ton) of trash processed was assumed for the value of the reclaimed metals and glass. A yield of 0.19 liters of hydrocarbon liquids per kilogram (45 gal/ton) of trash processed was assumed. The relative economics of the process were taken to be a function of source of capital, plant size, gasoline (hydrocarbon) value, and the dump fee per tonne of trash credited to the process. The plant size can be converted to the population served by the process by assuming a daily per capita trash generation rate of 2.27 kg/day (2.5 tons/1000 daily people). The capital and operating cost figures are based on those generated by an outside petrochemical consultant under contract to EPA for a 100 ton per day plant during their evaluation of the process. The consultant's cost estimate in early 1978 was $7.6M capital costs and $1.1M annual operating c (exclusive of debt service). The capital costs we scaled using 0.65 as the exponential scaling factor, whereas 0.20 was used as the labor cost factor.

As will be illustrated, the method of financing the construction costs has a significant impact on the apparent economics of the process. Figure 7 shows the economics for a plant financed at an interest rate of 8% and a 25 year amortization. This would be fairly typical of a municipally financed plant which could be operated either by the municipality or by a contractor. It is seen that with a $10/tonne dump fee credited to the plant that a 250 tonne/day plant would produce gasoline worth $0.10/liter ($0.38/gallon).

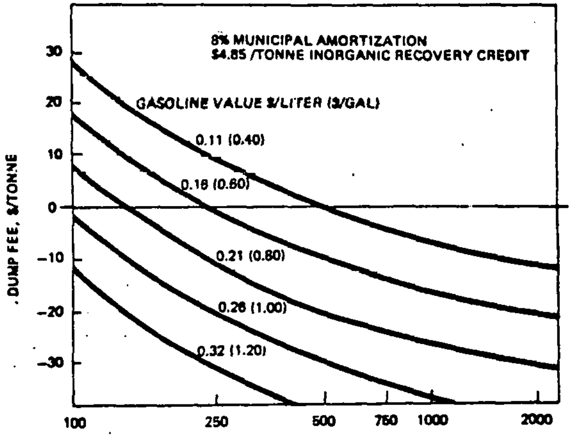

Fig. 7. Trash to Gasoline Economics with Municipal Financing

Figure 8 has the same assumptions as before, but wit private enterprise involved at a 15% rate of return. At 15% rate of return, the small 250 tonne per day plan with a $10/tonne dump fee must charge about $0.20/lite ($0.75/gallon) to meet its financial obligations. Fo private enterprise, larger plant sizes are clearl indicated. For example, at a 15% rate of return and th same dump fee, a 500 tonne per day plant size i estimated to produce gasoline valued at about $0.13/lite ($0.50/gallon).

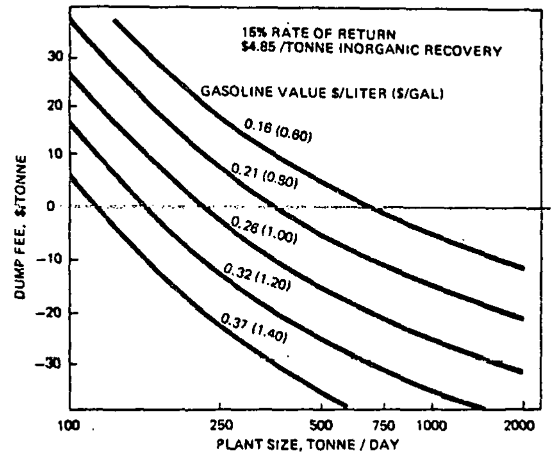

Fig. 8. Trash to Gasoline Economics with Priva Financing

ıe 500 tonne per day plant therefore appears to be an omically interesting size. One of the problems in ating the economics of this system is to project the ue of gasoline over the estimated 25 year life of the ant. Figure 9 shows the rate of return on the vestment for a 500 tonne per day plant using the same ıses as before. With a $10 per tonne dump fee and a ırrent $0.685/gallon wholesale gasoline value, a 23% te of return would be realized. If the wholesale value gasoline were $1.00 per gallon, the projected rate of turn would be 34%.

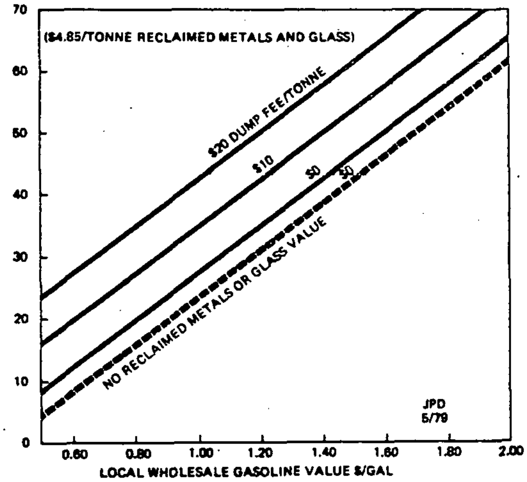

Fig. 9. Rate of Return for 500 Tonne/day Trash Plant

omass Feedstock

ıe process to convert trash to gasoline is thought to be plicable also to agricultural and silvicultural materials. ıese organics could be a mixture of wastes and/or ıterial specifically grown for conversion to fuel. The sumptions were modified somewhat to reflect the lack metal and glass recovery equipment and the lack of ese salvage products. The feedstock was assumed to ntain 90% dry, ash-free organic material. The yield of soline per unit weight of organic material was assumed be the same as with municipal trash derived organics. nilar economic trends are observed with the biomass ɔcessing plants as with the trash plants, but the onomical size of the biomass plants is larger than that the trash plants due to the difference in feedstock st.

interest private capital at a 15% rate of return, a 00 tonne per day plant could pay $30 per tonne of ganics and charge $0.72 per gallon as can be seen in gure 10. Examining a 1000 tonne per day plant in more tail, in Figure 11 it is seen that if the feedstock cost is 0/tonne and the local wholesale gasoline value is $1.00 gallon, the projected rate of return would be 31%.

ese economic predictions for the process point out ıt for the process to be able to buy the organic ɔdstock, fairly large plants will need to be constructed. wever, by commercial standards these plants at the 00 tonne/day size are in reality not very large and uld have an output of a fairly small oil refinery (less ın 2000 barrel per day). If the 1000 tonne/day biomass-to-gasoline plant were to be located at the center of a biomass producing area having an assumed annual biomass production rate of 11.2 tonne/hectare (5 tons/acre), the necessary biomass could be raised within a 10.2 km radius. Long hauling distances would not seem to be involved with this process and the gasoline produced would be consumed relatively locally.

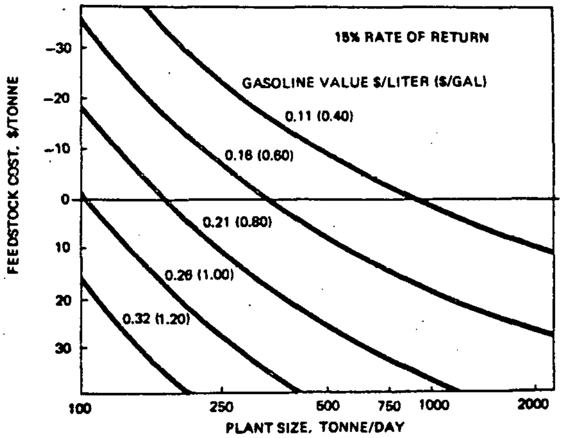

Fig. 10. Biomass-to-Gasoline Economics with Private Financing

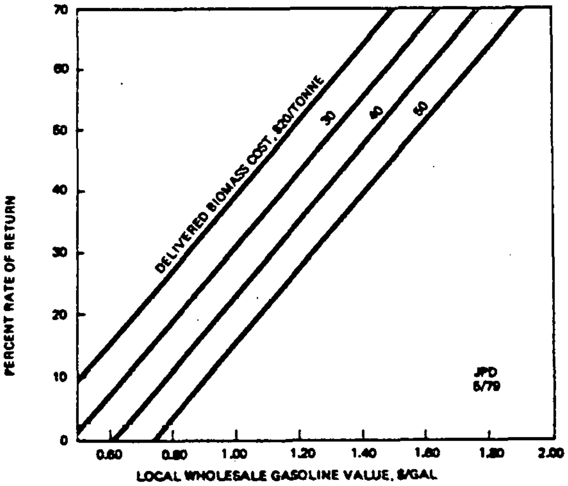

Fig. 11. Rate of Return for 1000 tonne/day Plant

SUMMARY

This process converts organic waste materials into the basic hydrocarbon building blocks, i.e. the olefins: ethylene, propylene, and butylene. Although these compounds could be used as highly valued petrochemicals, they can also be profitably converted to a high octane, unleaded gasoline with only a small amount of fuel and lubricating oils as the by-products. The process utilizes a high temperature, short residence time pyrolytic process with steam dilution to optimize the production of the desired gaseous olefins. This process is virtually identical to that used by the petrochemical industry to make ethylene and propylene from crude oil. Once the gaseous olefins are produced, the technology required to convert them into gasoline is state-of-the-art. Experimentation on a 5 kg/hr bench scale system has demonstrated the selective pyrolysis to form unusually large amounts of the desired olefins as well as their purification and subsequent polymerization to gasoline. It is projected that about 1.8 U.S. petroleum barrels of gasoline can be produced per ton of dry ash free organic material or about a 50% efficiency in energy conversion. The economic projections indicate that this process will be viable in moderately sized plants which could be widely dispersed.

Although some additional bench-scale process development work is needed, it is timely to make long range plans to exploit this technology. The outlook of rising trash disposal costs and restrictions on agricultural burning of wastes, as well as increased petroleum costs combine to make the potential of this process very promising.

ACKNOWLEDGEMENT

The authors wish to acknowledge the contribution to the effort made by Dr. Charles B. Benham (currently with the Solar Energy Research Institute), particularly in the polymerization demonstration with pure ethylene and the early development of the preliminary economics and their presentation format. Mr. Walter Liberick, Jr. and Mr. Charles Rogers of the Industrial Environmental Research Laboratory (EPA) have been the program officers responsible for the funding which made the hardware demonstrations possible. Portions of this paper also appear in the authors' paper "Noncatalytic Conversion of Biomass to Gasoline" presented to the ASME sponsored Solar Energy Conference held in March 1979 (ASME 79-Sol-29).

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ECONOMIC FEASIBILITY ASSESSMENT OF BIOMASS LIQUEFACTION

Sabri Ergun
Energy & Environment Division
Lawrence Berkeley Laboratory
University of California
Berkeley, California

Abstract

veral assessments have been made of the economic
asibility of liquefaction of organic municipal
ste and/or biomass since 1972, placing the
eakeven cost of liquefying from about $16.00/dry
n to over $58.00/dry ton cellulosic material
ocessed, exclusive of the cost of biomass, in
ants processing roughly 2,500 dry tons biomass
ily.

studied the cost assessments in order to trace
origins of disparities and to identify the unit
erations or unit processes that are costly so
at the R & D effort could be diverted to find
utions to the costly steps. The major causes
the disparity (corrected for inflationary trends)
e traced to the differences in the interpreta-
n of process chemistry. The LBL analysis
icated that the economic feasibility of bio-
s liquefaction approaches that of coal lique-
tion in spite of the fact that the biomass
uefaction plants have to be smaller by an order
magnitude.

NOMIC FEASIBILITY ASSESSMENT

s economic feasibility assessment concerns the
eau of Mines process for which the Albany facil-
was designed and built and which is now being
luated. On the basis of bench scale experiments,
rence Berkeley Laboratory (LBL) conceptualized
ee optional biomass liquefaction processes, one
which has already been tested at Albany, Oregon
found to be chemically and technically feasible.
other two process options are scheduled to be
ted at Albany during July 1979. The cost assess-
t made in this report pertains to the Bureau of
es process.

1973, Dravo Corporation was commissioned by the
eau of Mines to make a technical and economic
sibility study of biomass liquefaction by a
cess conceptualized at the Bureau of Mines and
provide the conceptual design of a process
elopment unit (PDU) that would permit the devel-
ent of the process should the process economics
roach economic feasibility. The Dravo study
ced the breakeven cost of oil produced at
00/bbl exclusive of the cost of raw material.

years later, National Science Foundation com-
sioned Bechtel Corporation to make a technical
economic desirability study of organic waste
uefaction in order to assess the operational
tification of the PDU by ERDA. The Bechtel
laced the breakeven cost at $16.13/bbl. During the same year Battelle Pacific Northwest Laboratories (PNL) made an independent analysis of the process for ERDA as a part of a study designed to explore the best use of the facility. The Battelle study placed the breakeven cost at about $12.00/bbl.

The plant construction was completed during December of 1976 and Bechtel Corporation was contracted to commission the facility, and conduct some experimental runs. The final Bechtel report (June 1978) included a cost analysis of the process which placed the breakeven cost of oil produced at $29.00/bbl, exclusive of the raw material cost, from a plant processing 2,500 TPD dry wood. Although Bechtel strongly recommended the continuation of the development program with the stated belief that the costs can be reduced through development, the high cost caused concern and LBL was asked to review the cost estimates.

Bechtel conceptualized two integrated plants, one for liquefaction, another one for gasification to provide synthesis gas for the liquefaction plant and other heat requirements. Each plant was sized to process 1,250 dry tons daily. The cost estimates made by Bechtel included the items listed below:

1. Annualized direct costs: site and yard, civil/structural, mechanical, piping and instruments, electrical, indirect (80% of labor), engineering, and uncertainty (20%).
2. Owner costs: land, startup, AFDC, and miscellaneous (4% of the total construction costs).
3. Operating costs: labor and material for maintenance, operating labor, supervision, administration and overhead, supplies, local taxes and insurance, and utilities.

The cost estimates were thorough and conservative. The totals translate into a wood processing cost of $58.00/dry ton. Bechtel has assumed that 2 bbls oil (a round figure) would be produced per dry ton total wood utilized. The $39.00/bbl figure translates into $78.00/dry ton. Subtracting the cost of wood ($20.00/dry ton), the cost of processing becomes $58.00/dry ton.

An examination of the cost breakdown for the two plants leads to the following figures for processing costs:

1. The cost of liquefying wood is $40.00/dry ton.

2. The cost of gasifying wood is $74.00/dry ton. The average being $58.00/ton.

To translate these operating costs to a breakeven cost for the oil produced, Bechtel made the following assumptions:

1. Fifty lb dry wood will yield 35 lb oil upon liquefaction.
2. Fifty lb dry wood has to be gasified to produce the carbon monoxide needed for liquefaction and process heat requirements.
3. The oil produced has a heating value of 15,000 Btu/lb.
4. The gross thermal efficiency of the process is 66%.

Neither the conceptual design nor the operating costs are in question. However, the assumptions cited above are not acceptable.

Assumption 1 requires an oil yield of 70%! Experimental results obtained at the Bureau of Mines, PNL, and LBL indicate the yield of an oil having the composition assumed by Bechtel cannot exceed 58%. If the Bechtel cost estimate is corrected for this lower yield, the cost of oil would be $49.00/bbl instead of $30.00/bbl. This assumption is over-optimistic.

Assumption 2 is a consequence of the reaction stoichiometry assumed by Bechtel. There are no data in the literature indicating that carbon monoxide consumption is over sevenfold that determined by the Bureau of Mines. This assumption is very pessimistic.

Assumption 3 is baseless. An oil having the composition that Bechtel assumed, i.e., 23% oxygen, is likely to have a heating value of about 13,300 Btu/lb and not 15,000. This assumption is very optimistic.

Assumption 4 may not be regarded as an assumption because it is the consequence of the three preceding assumptions. However, we are not sure of this. This may have been the assumption that dictated or influenced assumptions 2 and/or 3. If corrected for lower oil yield and lower Btu value, the gross thermal efficiency of the process would be 48% instead of 66%. Even a 66% gross thermal efficiency is extremely conservative for a liquefaction process (for coal it is about 80%).

In summary, the chemical and stoichiometric bases of cost estimates are fraught with many inconsistencies and invalid assumptions.

After a careful evaluation of the results of recent research on wood liquefaction at PNL and LBL, and comparing these results with those reported by the Bureau of Mines, we arrived at the chemical bases listed below:

1. Eighty ton dry wood will yield 46.4 ton oil (58% conversion efficiency) upon liquefaction.
2. Twenty ton dry wood will be gasified to pro the carbon monoxide and heat requirements.
3. The oil produced will have a heating value of about 13,500 Btu/lb.

These bases lead to a gross thermal efficiency of about 78%, i.e., do not include the electricity used.

Using the chemical bases listed above and process costs provided by Bechtel, the breakeven cost of oil is as shown below:

$$\frac{(42 + 20)80 + (74 + 20)20}{46.4}\;\frac{350}{2000} = \$25.80/\text{bbl.}$$

The factor 350/2000 converts barrels to tons.

The chemical bases listed above require a larger liquefaction plant (80 vs. 50) and a smaller gasification plant (20 vs. 50) than those assumed in the Bechtel conceptual design. If we adjust the processing costs for changes in plant sizes (42 becomes 36.5 and 74 becomes 97), the cost increas by only $1.00. However, we cannot take the $97.0 figure seriously.

In summary, had Bechtel interpreted the reaction chemistry properly, they would have arrived at a breakeven cost of about $26.00/bbl and an oil yield 33% higher than they projected.

In terms of heating value of the oil produced, th latest Bechtel cost estimate places the breakeven cost of oil at $6.13/MM Btu, exclusive of the material cost ($8.25/MM Btu including the cost of the wood). The cost estimate revised by LBL plac the corresponding costs at $3.86 and $5.46/MM Btu respectively. In the Bechtel report the cost of oil produced from coal by the SRC II process is placed between $3.00 and $4.00/MM Btu including the cost of coal. If we assign $1.00/MM Btu for the cost of the coal and 80% for thermal efficien the cost of the SRC. II oil would be between $1.7 and $2.75/MM Btu. If the latter figures are tran lated into plants processing 3,000 TPD coal inste of 30,000 TPD, the cost of processing would nearl double, i.e., $3.50 to $5.50/MM Btu. Considering the fact that biomass is much more reactive than coal, liquefies at much lower temperatures (680°F vs. 860°F), does not require pure hydrogen and contains no elaborate cleanup system in the lique faction process because it contains little or no sulfur, nitrogen, and ash, the cost figure revise in this report is very conservative. In its conceptual design Bechtel incorporated an expensi product oil recovery system and a very expensive gas cleanup system without a valid data base.

Banquet Address

THE BURNING ISSUE
(Banquet Address)

Paul F. Bente, Jr.
Executive Director
The Bio-Energy Council

dies and Gentlemen:

e most urgent issue of our times is the rapidly panding population of the world. Two symptoms of is phenomenon are 1) intensive farming to produce od and 2) the exploitation of the world's non-newable resources. Both symptoms have a common aracteristic -- heavy consumption of liquid and seous fossil fuels.

n that be all had? Our food production is undant, even sufficient to forestall famines in her parts of the world. Certainly, a lavish pply of material goodies has placed us in the lap luxury. From that perspective one might say o far, so good."

is reminds me of the story you have probably ard, but since I have the microphone, I'll tell anyway. An optimist attended a cocktail party the top floor of a high rise apartment building. more and more guests arrived, he finally got owded off the balcony. As he fell, people on the wer floor balconies heard him repeating to him-lf, "So far, so good. So far, so good!"

s self assurance seems to me to be at the heart the problems I want to discuss. His rationale typical of mankind's obstinate rejection of ality and mankind's collective failure to make me hard decisions. These decisions are necessary reconcile demands for more and more energy and terials with the inevitable depletion of the n-renewable resources from which they are derived. en production of food must be undertaken with due re so as not to destroy renewable pastures and oplands. This fate has already befallen vast eas in the world because of overuse and careless rming that causes erosion. By causing a tremen-us amount of deforestation, food production has ten been counter productive.

t me return to the basic issue of population owth. Within my lifetime modern advances in dicine and sanitation have spread worldwide. is has improved the health and increased the ngevity of all peoples remarkably. Babies for-rly doomed to die now survive to become parents. nsequently, population has expanded at a mind ggling rate (sometimes referred to as geometric owth).

Throughout early unrecorded and recorded ages of time, world population grew very slowly, finally reaching the one billion mark by the year 1830. It took only 100 years more for the population to double. In the next 30 years it added the third billion. The fourth billion arrived after 15 more years -- that is, by 1975. Today, the world population has already passed 40 percent of the fifth billion.

People who are now 70 years old have lived through a growth of 2-1/2 billion in world population. If such a growth rate were to continue, and I seriously doubt if it could be sustained, those who are born today and live threescore years and ten will see the world population increase about 5 times more than did the septuagenarian of today. How to cope with the population explosion is mankind's greatest challenge. As one wag said, "We simply can't grow on like this!"

The last century has produced an equally mind boggling technology explosion. This has both enabled and induced mankind to tap the world's non-renewable and renewable resources at fantastic rates. Over the years, on the average, each individual of the expanding population is consuming ever greater quantities of materials and foods. Those who haven't yet started to consume more, are aspiring to do so. Those who already have an abundance have developed an insatiable appetite for more. They often refer to the "have-nots" as having "rising expectations." This phenomenal growth in consumption would not have been possible without cheap energy derived from fossil fuels.

We could spend all night on statistics documenting increasing consumption of materials wrested from the earth with cheap energy. But I shall cite only a few which typify the trend. World-wide production of automobiles had by 1975 grown to about 25 million per year, and that of buses and trucks to about 8 million. These have a ferocious thirst for liquid fossil fuels. The U.S. fleet of about 100 million autos consumes about 100 billion gallons of gasoline per year. That's about twice as much as can be produced from our country's oil wells, Alaska's included. In other words, half of the motor fuels we consume comes from imported petroleum.

Society today has not only become very mobile. In addition it has become dependent on export/import of countless materials and products from all parts of the world. Shipping this seemingly inexhaustible supply of goods also depends on liquid fossil fuels. We might say that today's way of life is founded on petroleum, which we now finally understand is disappearing.

The first thing we must do is use less energy by wringing more goods and services out of the energy we use. This means more than eliminating waste or simply cutting back on what we do in order to conserve, both noble objectives in themselves. Improving efficiency in use of energy must become a new way of life.

The Council on Environmental Quality has presented this picture clearly in its recent report "The Good News About Energy." It points out that growth in energy use by the end of the century can be reduced to no more than 10 to 15 percent, instead of the 250 percent increase over 1970 usage that was predicted before the oil embargo occurred.

This change can occur as we apply cost effective measures that increase energy productivity. This, in turn, will have a positive effect on employment, provide a high standard of living, cause continued economic expansion and help fight inflation. Further, this will benefit the environment by causing less pollution. Before moving on to other themes, let me take the following as a given, namely, that regardless of the source of energy, we will learn to use it more efficiently, getting more goods and services from all the energy that is used.

Although in our society one can still use as much energy as one wishes, or as one can afford, obviously a person can only eat so much. Food, after all, is a renewable resource. If that reasoning were correct, we could conclude that at least our eating habits aren't compounding the depletion of non-renewable resources. But that reasoning overlooks the fact that there has also been a large per capita increase in food consumption in the more developed countries of the world.

This has occurred because of advances in agriculture which have also occurred in my lifetime -- namely, growing huge amounts of corn and other grains as feed for livestock in order to have plenty of well marbled meat in the marketplace. Specifically, U.S. farms produce about 7 billion bushels of corn per year on 65 million acres of good farmland. Ninety-five percent of that corn is fed to livestock.

Farm production has been more than doubled by genetic improvements coupled with intensive agricultural practices. These require much less manpower than earlier methods of farming but a great deal more energy, again mostly in the form of petroleum fuels. These are used to run tractors, combines and other farm equipment, to manufacture fertilizers, insecticides, herbicides and fungicides and to provide necessary transportation.

The non-farm population in the food system then uses large amounts of energy to process, package, distribute and merchandise the food. The consumers also use large amounts of energy driving to the supermarkets and finally cooking the food.

Altogether, about 13 percent of the energy consumed in our country (or about 10 Quads) is used to put food on our tables. Only about one-fourth of this is actually used on the farms. To think that our foods are safe because they are produced from renewable resources is a myth. Our food supply is now heavily dependent on non-renewable energy. If we run short of petroleum, we may not only have to drive less, we may have to eat less, or at least eat less corn-fed beef.

Without doubt, history will record that in our day and age we consumed most of the world's petroleum resources and quite possibly most of the natural gas resources. Will history also eventually record that, before the fossil fuels ran out, we managed to switch completely to renewable energy? That is the burning issue which we must address.

I congratulate you for having dedicated yourselves to that task. You have an assured place in history as having been among the earlybirds who took the first serious, though somewhat faltering steps in that direction. It will probably turn out that many of these steps were rather meandering and often led down blind alleys. But some, I believe, will blaze a pathway to success.

We are still, relatively speaking, in the dark ages in terms of making efficient and effective use of the radiant energy sent out by the sun. We still have a great deal to learn in order to tap the vast storehouse of green energy collected by plants.

The amount of solar energy is simply staggering. Let me give you the round numbers for the bio-energy portion. Although plants capture only about 1 percent of solar radiation, they store up some 20 times as much energy in a year as the whole world now uses. Today biomass is the main source of fuel for half of the people in the world. It has been estimated that, used efficiently, 10 percent of the world's current yearly production of biomass could readily meet the entire projected year-2000 requirements for food and energy, but not if the whole world lives on as grand and wasteful a scale as we do.

A tremendous amount of progress must be made before bio-energy can realize its potential. There is already considerable activity going on in the bio-energy field, but we lack an adequate inventory of what is occurring.

The Bio-Energy Council, with which I am affiliated, has started to address this problem by publishing the Bio-Energy Directory. The May 1979 Directo gives one-page summaries of almost 500 activit going on in the North American sector. There a viously many more activities in North America to mention other regions of the world) which ven't yet learned about and some which we know out, but which the organizations carrying out o-energy activities decline to have publicized.

me activities reported are efforts that have been stained over a period of many years. Some have ready been commercialized involving major investnts. Others are just getting started or report ly the current level of effort. Some programs ve been proceeding which duplicate those of others derway with the same goals. At best, it is a xed bag.

vertheless, it is instructive to compare the gnitude of the effort that has been launched by e Federal versus the non-federal governments and the private sector, as compiled in the Directory. ding all kinds of dollars together, that is exnditures for past work, investments made, rrent level of spending and forecast expendires gives the following picture.

deral funding accounts for $180 million; nonderal government funding amounts to $160 million; d private sector expenditures are about $600 llion, for a total of about $940 million. Hopedr future funding, as distinct from forecast exnditures, would add $30 million to the Federal lly as well as to the non-federal government gure, and $585 million more to the private ctor total, if realized. Adding together all mmitted forecast and hoped-for expenditures vered in the Directory gives a grand total of arly $1.6 billion.

e lion's share of this, 60 percent of the mmitted and all of the hoped-for funding, falls the area of thermal conversion of biomass. is consists mostly of combustion processes. arly 30 percent more of the committed funding lls into the area of biomass production, chiefly lated to forestry. Microbial conversions comise 8 percent of fundings. Certainly if unrerted investments and R and D expenditures made by e wood processing industry for internal use of o-energy could be included, the grand total uld be several times the $1.6 billion reported.

e of bio-energy in the United States is already eater than most people realize. It is reported be about 1.8 Quads per year. That is equivalent about 10 percent of the petroleum imported, and out 2.5 percent of the total energy consumed.

st month, Secretary Bergland set some targets for velopment of bio-energy. He set the goal of ving all U.S. agricultural production and all rest production and processing of forest products come self-sufficient in net energy by 1990 witht impairing productivity. I am glad to see one m of the government finally set a goal for conrsion of biomass to energy.

cording to the Secretary's estimates, achieving is goal will require production of only 4 more of energy from biomass, assuming all the energy comes from that source. This goal is judged to be "achievable" because the "ultimate potential," is stated by the USDA to be "well above 10 Quads." This is a potential which won't preempt or interfere in any way with present food production and its system of land set-aside safeguards. After all, the Emergency Agriculture Act of 1978 prohibits growing grain for alcohol on set-aside land. The goal of 4 Quads impresses me as being a safe bet on a business-as-usual basis, picking up the easiest part of the more than 10 Quad potential that is now untapped.

What about also setting out to use farmlands and forests to enhance biomass production? The week before last at the Purdue meeting, we were informed that by simply fertilizing the so-called marginal lands that lie east of the Mississippi River, that is, the 30 million acres of haylands, the 42 million acres of permanent pastures, production of grasses could be increased by over 200 million dry tons above present needs, enough to yield about 3 Quads of bio-energy. The energy yield would be 8 times the energy put in to achieve this increase.

Is this worth doing? It is equivalent to 10 percent of all liquid fuels consumption or 20 percent of fuels consumed on the highways.

We harvest about 330-340 million acres of cropland annually. There is more land that could be cultivated but is set-aside so that farms don't produce more than the market can absorb. Three out of every 10 acres of croplands produce products for export to other countries. Isn't the day dawning when we and others in the world should stop feeding cattle and other farm livestock 95 percent of the corn produced on U.S. farmlands? Why not add a 1990 goal of making alcohol from half of this grain and feeding the protein-rich residues to the cattle? Of course, there are not now large export markets for distillers dried grains. But, with the tremendous world demand for foods and feeds and with no other country having the farms that can meet this export demand, it seems that economically rewarding export markets for the protein-rich residues left over from alcohol production could be developed. That should be part of the target.

We should set much higher bio-energy goals for the farm and forestry sectors, for these are the basic sources of biomass. For starters, the USDA 1990 goal of 4 Quads ought to be tripled.

In addition we should set year-2000 and 2020 goals which include production of biomass by aquaculture, that is, both marine and freshwater farming of water plants. Likewise we need goals for production on land of trees and crops from genetically improved stocks developed specifically for production of biofuels.

Let's not forget energy that can be recovered from municipal and industrial wastes. Nationally, we should be able to recover at least one Quad of energy from such wastes. We generate 2.3 lbs. of waste per person per day in urban areas and 1.3 in rural areas. If there were a world prize for waste generation, we'd win it hands down. Moreover, current waste handling procedures are counter productive. Consider, for example, that it costs the city of Columbus, Ohio between 1/2 and 3/4 million dollars per year to bury grass clippings!

The USDA goal should be integrated with yet-to-be defined goals of the Department of Energy, the Environmental Protection Agency, the Department of Defense (the largest user of energy) and other government agencies. We need a National Goal.

President Carter alluded to the importance of bioenergy over a year ago when he visited SERI on Sun Day (May 3, 1978). He promised then that by Labor Day a Domestic Policy review would produce an assessment with some goals. We don't have it yet. Then the Department of Energy announced that by the end of 1978 it would release an alcohol study, presumably also with goals. We still await this report. On May 4, a year and a day after President Carter spoke here on Sun Day, he made a speech in Iowa endorsing Gasohol and on-farm production of ethyl alcohol. He announced $11 million in funding for 100 small-scale plants to produce alcohol as a transportation fuel, namely Gasohol. He also referred to $30 million of already guaranteed loans to construct a large Gasohol plant in Florida and another in Texas.

He said, "Our best calculations are that our Nation can produce 300 million gallons of Gasohol annually by 1982 -- and double that again by 1985." The people applauded, but did they understand? By my calculations, the 300 million gallons of Gasohol will require 30 million gallons of ethyl alcohol. If we take a gallon of ethyl alcohol as equivalent to a gallon of gasoline, the 30 million gallons of alcohol would extend the Nation's 100 billion gallons of gasoline consumption by only 0.03 of one percent. In truth it would be not much more than half of that if corrected for the fuel the farmer uses to grow the grains needed to make the alcohol.

Based on the May 16 testimony of Alvin L. Alm, DOE's Assistant Secretary for Policy and Evaluation, President Carter should be given the benefit of some doubt. It appears that he really meant to say 300 million gallons of ethyl alcohol, not 300 million gallons of Gasohol, as the White House press release states. That amount would extend net gasoline supplies by nearly 0.2 of one percent.

I believe that the new biomass managers in DOE, SERI and the regional centers recognize the urgency of this situation and want to break out of this business-as-usual syndrome. I think it is high time that we shift into a much more determined pace.

For "so far, so good" is suicidal logic -- which brings me back to the optimist who fell off the high rise balcony. The trouble with his logic is that, although it was absolutely irrefutable during his fall, on his arrival at ground level, it revealed a fatal flaw. For as the old saw puts it, "It's not the fall that kills you, it's the sudden stop!"

Nero fiddled while Rome burned. Are we fiddling while our petroleum is burning up? Is anybody running scared? The burning question, I repeat, is this: "Can we learn to replace petroleum with renewable energy?" Let's not worry yet about replacing coal. We have to take this one step at a time. Let's use the tremendous resource of solar energy, and in particular, the energy constantly being stored up by earth's original, enduring, and ubiquitous energy factory -- photosynthesis. We have to arouse this sleeping giant, bio-energy, to come to our rescue and keep mankind from being crowded off his balcony. For if that happens, it's too hard to change course on the way down, picking up speed.

I don't have to tell you that similar challenges were a way of life in our past and the dedication to the solution of these challenges is what made our country great. I hope we haven't lost the vigor and the vision to surmount the dilemma we are in.

Let's keep our eyes on the sun, for it will not only bring us tomorrow as a new day, it will show us the new way. This hopeful thought has been captured by a perceptive high school senior in the following verses which I now quote, hoping that this peotic expression will inspire you to develop bio-energy so that tomorrow will be a worthwhile time for living.

I watch the sun quietly rising
Giving birth to a new day.
Tomorrow has come, a time for living.

Wishing for youth everlasting,
Dreaming of children at play,
I watch the sun quietly rising.

With a saddened heart I cling
To the past, grasping it, wishing to stay,
But tomorrow has come, a time for living.

Memories are passing.
Clutching them, I consider today
As I watch the sun quietly rising.

This is the day for silent thinking,
Deciding on a different way.
Tomorrow has come, a time for living.

I listen to the swallows chirping,
Welcoming the warm sun rays.
I watch the sun quietly rising
Tomorrow has come, a time for living.

Lisa

Session III

BIOMASS PROCESSES
SYSTEMS EVALUATION AND ECONOMIC ANALYSIS

phael Katzen*
nn R. Messick
orge D. Moon, Jr.

Raphael Katzen Associates
Cincinnati, Ohio

celerating depletion of the world supply of non-
newable energy resources has brought into focus
e need to employ renewable resources for future
obal energy and chemical feedstocks. Biomass,
a myriad of forms, can supply a large portion
that demand.

e creation of technically and economically feasi-
e biomass energy systems will require the effect-
e coupling of existing technologies with bold new
ientific concepts. Concentrated interdisciplin-
y research efforts supported by large-scale fund-
g, from both governmental and industrial sources,
e being brought to bear upon the problem of as-
ssing the total biomass potential and how it may
realized. Many of those present here are in-
lved with these efforts in some manner.

e engineering community is faced with the chal-
nge of coupling research and development results
th "best available technology" to produce viable
mmercial biomass-based energy systems. The stra-
gy which we at Raphael Katzen Associates use to
complish successful commercialization, or indeed.
jection, of proposed processes is the subject of
is paper.

illustrate the methodology which we employ, we
ve chosen, as an example, a grass-roots facility
produce 50 MM gal/yr of motor fuel grade alcohol
om corn. In late 1978, we were asked to prepare
major study for the Department of Energy; the re-
lts of which were reported last week at the Third
ternational Alcohol Fuels Technology Symposium at
ilomar, California. All the examples used here
ve been taken from that report.

establishing the commercial viability of a pro-
sed venture, the engineer must follow a methodi-
l sequence of steps. These steps may be categor-
ed into:

- Study of Process Alternatives and Technologies
- Develop Process Design
- Prepare Economic Analysis.

The first category, Study of Process Alternatives and Technologies, may involve only a cursory review of existing or proposed technology before selecting the best process; but for many evaluations, several processes may have to be studied in parallel. Once the best process or processes technologies have been selected, the engineer proceeds with the process design.

Slide Number 1 outlines a typical sequence of events.

<u>STEPS FOR PROCESS EVALUATION</u>

1. **Establish the Design Basis**
2. **Accumulate and Assemble All Available Process Data**
3. **Prepare Preliminary Block Flow Diagrams**
4. **Prepare Conceptual Flowsheets**
5. **Prepare General Material and Energy Balances**
6. **Revise Process Based on Insights Gained in Steps 3, 4 and 5**
7. **Establish Final Process Design Including Detailed Material and Energy Balances.**

SLIDE NO. 1

The design basis must be established first. Next, it is a good idea to accumulate and assemble all available process data and information, including, if necessary, a computer-aided literature search. This prevents disruption of the engineer's thought pattern or delays in making process decisions due to not having data at ones finger tips.

The preparation of preliminary block diagrams and conceptual process flowsheets is just that - PRELIMINARY and CONCEPTUAL. This is absolutely necessary in order to establish an optimum design. <u>Flexibility</u> in design is the key at this point. There is nothing more wasteful and demoralizing than to spend all of ones time, effort, and money establishing a process flowsheet with detailed material and energy balances, only to discover later that a cursory look at the situation would have

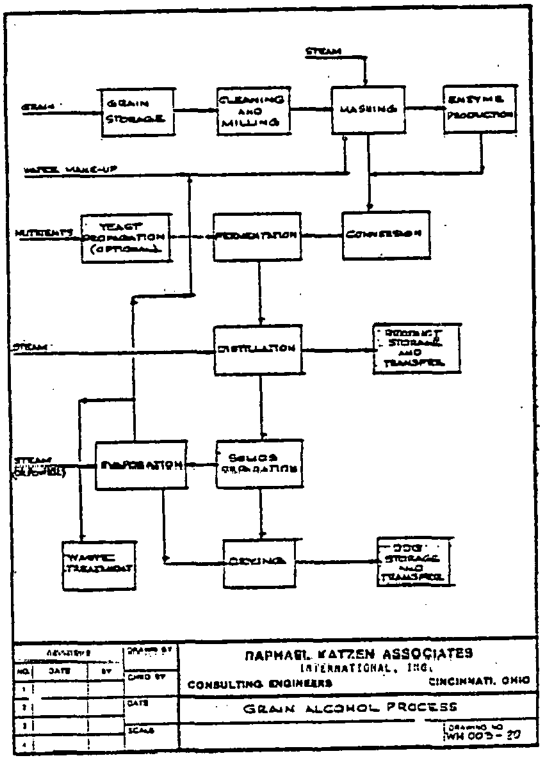
aker

shown other options to be more desirable. At this stage of the study one must decide on such options as:

- Vapor recompression evaporators versus multiple effect evaporators
- Use of reboilers, direct steam, or compressor to provide vapor traffic for distillation columns.
- Co-generation of electricity versus purchased power
- Staging of distillation columns versus single column operations.

In most situations, these options can be decided by evaluating the single unit operations. Sometimes, however, the option may be so complicated that parallel studies have to be performed.

Slide Number 2 illustrates a block diagram of the basic process modules in the production of motor fuel grade alcohol from grain.



SLIDE NO. 2

From this diagram an engineer sets out to establish the best process under the conditions prevailing at the proposed site.

After optimizing the overall design, the process flowsheet is finalized. Only after finalizing the process does the engineer prepare the detailed material and energy balance.

At this point, the engineer has established the most desirable process or processes. His process decisions have been based principally on elementary economic rules. The next step is to prepare a complete economic analysis.

The first step in developing an economic analysis is to determine the accuracy requirement of the project. An analysis may be developed ranging from "back of envelope" quality information, to a detailed analysis based on complete drawings, specifications, accurate material and energy balances and actual freight and utility costs. The first may be accurate to ±50%, whereas the latter may be accurate to ±5%. Between the two extremes, the engineer selects an analysis quality dictated by the needs of the project.

After establishing the accuracy, the engineer determines the type of financial analysis required of the project. Historically, engineers have relied chiefly upon three methods:

- Simple return on original investment
- Payout time or cash recovery
- Return on average investment

These methods are useful (though they are not without drawbacks), when comparing alternative processes, but they are inappropriate when evaluating a capital expenditure program based on the time value of investments and money returned to the project.

Over the past two years, Raphael Katzen Associates has developed a method for evaluating the true financial return of a venture; or alternatively, the appropriate selling price required for a company to realize a satisfactory interest rate of return on investor equity. We call the method a "Preliminary Budget Authorization Analysis (PBAA)". The accuracy is in the range of ±10% to ±25%, depending on the accuracy of the data and cost projections.

The steps necessary to complete a PBAA economic analysis are shown in Slide 3:

STEPS REQUIRED FOR ECONOMIC ANALYSIS (PBAA)

1. Establish raw materials & chemicals usage, and labor requirements
2. Prepare listing of total utility requirements
3. Prepare battery limits investment estimate
4. Determine total plant capital
5. Determine cost projections for all purchased items
6. Develop financial analysis.

SLIDE NO. 3

e first five steps establish the necessary infor- n to prepare the financial analysis. Let's a closer look at these steps.

v materials, chemicals and utilities requirements e obtained from the detailed material and energy lances. Adjustments must be made for peak loads d process control limitations. The flowsheets ovide average usages. Labor requirements are es- nated from the number of control loops and the tent to which assistance is provided the opera- r (use of computer control for example).

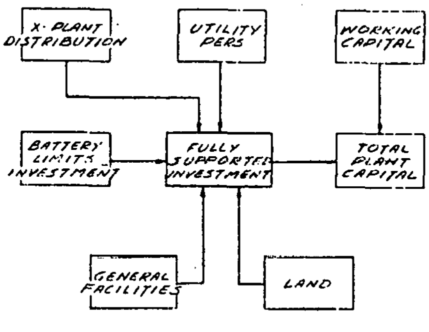
e battery limits investment estimate is only a rt of the total plant investment. The limits y circumvallate the total plant, or more typical- , just the operating units. The procedure for termining the battery limits investment is:

- Size all major equipment
- Determine major equipment purchase price
- Use cost factors to establish the installation cost
- Apply appropriate inflation escalation to construction schedule
- Determine construction cost and contractor fees
- Estimate overhead expenses
- Estimate engineering, start-up and project management cost
- Apply contingency factor.

ide 4* shows the investment, by process section, r the Grain Motor Fuel Alcohol plant evaluated the DOE report. Slide 5* gives a more detailed eakdown for each part of the plant.

e total plant capital required is the sum of sev- al components. Many mistakes can be made in de- rmining the plant investment by not taking into count all cost components. Slide 6 demonstrates ese components:

TOTAL PLANT CAPITAL



SLIDE NO. 6

We have already talked about the battery limits investment. Let us now discuss what the other terms mean.

X-Plant distribution is the investment associated with feedstock receiving or product shipping outside the boundaries of the plant. It includes items such as trucks, garages, and storage depots. Utility pers are investments associated with utilities not included in the battery limits investment. An example is the purchase of steam from a central boiler. The investment must show its share of the cost. Investments for general facilities and land must also be included.

Working capital is sometimes left out or grossly under-estimated. Components of working capital are shown in Slide 7.

WORKING CAPITAL

1. Raw materials inventory (1 month)
2. Products in stock (1 month)
3. Accounts receivable (12% annual income)
4. Operating cash (2 months)
5. Accounts payable (negative - 12% raw material cost)

SLIDE NO. 7

The factors associated with each component of working capital will vary depending on the requirements of the project. For example, in the DOE report, working capital was minimized by providing for only one week's grain storage. This meant the grain would be purchased from local grain elevators, probably at a slightly higher price than what one would pay if long-term storage were provided.

The key to developing an accurate capital requirement is to take all factors into consideration.

After the engineer has completed tasks associated with the first five items listed in Slide 3, he is prepared to proceed with the financial analysis.

If the engineer uses one of the simpler analysis techniques mentioned earlier, his remaining work is relatively easy. For example, the steps associated with percent return on investment are:

- Develop the single year operating cost
- Determine total income
- Determine gross profit for one year
- Establish percent return on investment after taxes.

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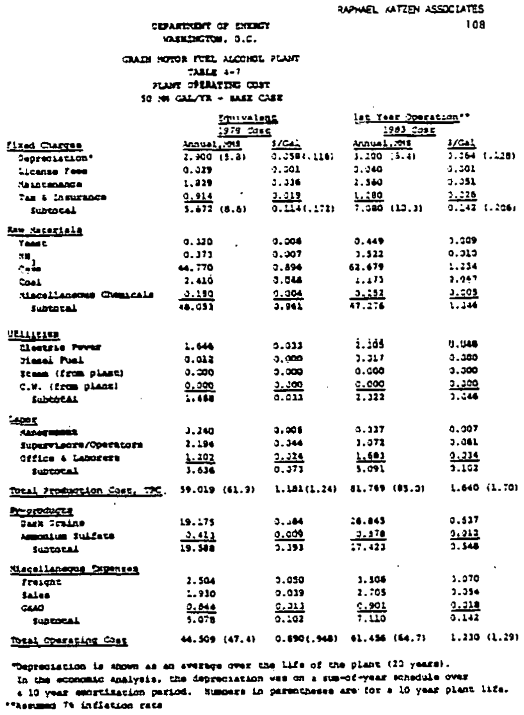
des 4 and 5 appear at end of paper.

Slide 8 indicates the cost items that must be considered in establishing the single year operating cost.



RAPHAEL KATZEN ASSOCIATES
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DEPARTMENT OF ENERGY
WASHINGTON, D.C.

GRAIN MOTOR FUEL ALCOHOL PLANT
TABLE 4-7
PLANT OPERATING COST
50 MM GAL/YR - BASE CASE

	Equivalent 1979 Cost		1st Year Operation** 1983 Cost	
	Annual, MM$	$/Gal	Annual, MM$	$/Gal
Fixed Charges				
Depreciation*	2.900 (5.8)	0.058(.116)	3.200 (6.4)	0.064 (.128)
License Fees	0.029	0.001	0.040	0.001
Maintenance	1.829	0.036	2.560	0.051
Tax & Insurance	0.914	0.019	1.280	0.026
Subtotal	5.672 (8.6)	0.114(.172)	7.080 (10.3)	0.142 (.206)
Raw Materials				
Yeast	0.320	0.006	0.449	0.009
NH_3	0.373	0.007	0.522	0.010
Corn	44.770	0.894	62.879	1.258
Coal	2.410	0.048	3.173	0.063
Miscellaneous Chemicals	0.190	0.004	0.252	0.005
Subtotal	48.063	0.961	67.276	1.346
Utilities				
Electric Power	1.644	0.033	2.305	0.046
Diesel Fuel	0.012	0.000	0.017	0.000
Steam (from plant)	0.000	0.000	0.000	0.000
C.W. (from plant)	0.000	0.000	0.000	0.000
Subtotal	1.656	0.033	2.322	0.046
Labor				
Management	0.240	0.005	0.337	0.007
Supervisors/Operators	2.194	0.044	3.072	0.061
Office & Laborers	1.202	0.024	1.683	0.034
Subtotal	3.636	0.073	5.091	0.102
Total Production Cost, TPC	59.019 (61.9)	1.181(1.24)	81.769 (85.0)	1.640 (1.70)
By-products				
Dark Grains	19.175	0.384	26.845	0.537
Ammonium Sulfate	0.411	0.009	0.578	0.012
Subtotal	19.586	0.393	27.423	0.548
Miscellaneous Expenses				
Freight	2.504	0.050	3.506	0.070
Sales	1.930	0.039	2.705	0.054
G&AO	0.644	0.013	0.901	0.018
Subtotal	5.078	0.102	7.110	0.142
Total Operating Cost	44.509 (47.4)	0.890(.948)	61.456 (64.7)	1.230 (1.29)

*Depreciation is shown as an average over the life of the plant (20 years). In the economic analysis, the depreciation was on a sum-of-year schedule over a 10 year amortization period. Numbers in parentheses are for a 10 year plant life.
**Assumed 7% inflation rate

SLIDE NO. 8

Operating cost items include:

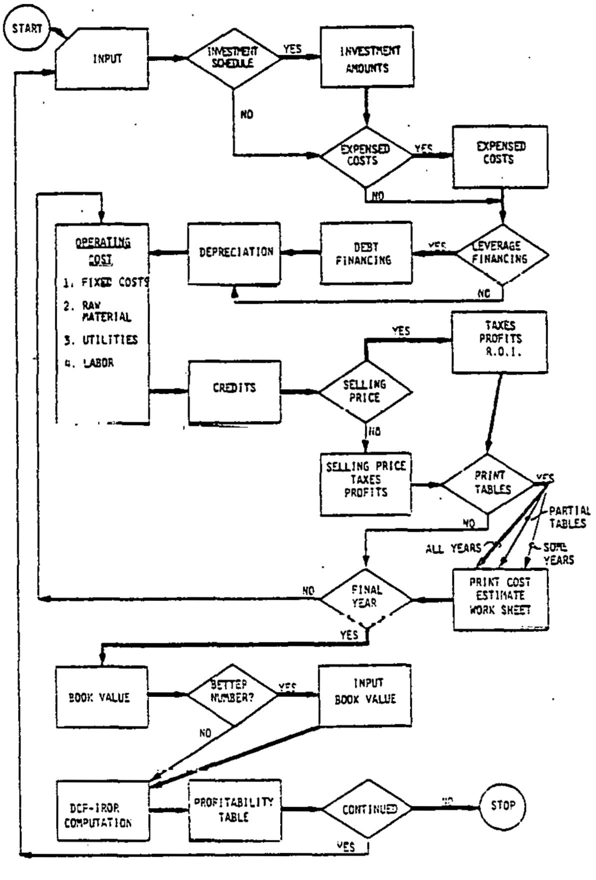
- Fixed charges
- Raw materials
- Utilities
- Labor
- Miscellaneous expenses
- By-product credits.

In our discussions earlier we stated that Raphael Katzen Associates has developed a financial analysis method called "Preliminary Budget Authorization Analysis (PBAA)". We developed this method because the conventional analyses often do not represent the real life situation correctly. The PBAA analysis has the following specific advantages:

- Time value of money taken into account
- Inflation is included as an input variable
- Each cost item may be inflated at different rates
- Present value of product selling price is established
- Complete profit picture of the venture (both actual and discounted) is established over its projected life.
- The effect of leveraged capital can be easily studied
- Tax credits and other governmental incentives can be adequately evaluated.

As you can see, the engineer must have a multidisciplinary background in order to successfully carry out a financial analysis. In order to reduce the engineer's work load to a manageable level, we developed a computer program to do the tedious but routine computations, prepare annual cost tables, and summarize results. The program compresses several man-weeks of work into just a few hours. The program contains the latest tax computations, credits, etc; but the engineer still has to supply input variables associated with these computations. With the program, the engineer is forced to take all factors into consideration. Slide No. 9* demonstrates the amount of information that must be provided to the computer before meaningful results can be obtained.

Slide No. 10 demonstrates the computer flow logic.



SLIDE NO. 10

* Slide 9 included at end of paper.

steps of the program are shown in Slide 11. nnual operating cost information developed by e computer is the same as that shown in Slide 8. separate table is prepared for every operating ar.

PBAA PROGRAM FLOW LOGIC

1. Input Data File
2. Input Investment Schedule if Required
3. Input Additional Expenses Charged to Project
4. Input Debt or Leveraged Capital Financing
5. Compute Debt Financing Schedule
6. Compute Depreciation Schedule
7. Compute Amount Operating Cost (Slide 8)
8. Compute By-Product Credits
9. Establish if Product Selling Price Known or % ROI Required
10. Compute Taxes
11. Compute Tax Credits
12. Print Annual Financial Tables
13. Determine if all Operating Years have been Evaluated
14. Establish Salvage Value of Plant
15. Compute DCF-IROR for Project
16. Print Project Profitability Tables
17. Continue to Another Analysis
18. Stop - End of Analysis

SLIDE NO. 11

example of the summary printout from the PBAA ogram is shown in Slide No. 12.

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               PROFITABILITY, CASH FLOW, AND IROR

                      FOR SAMPLE PROGRAM

YEAR    TCI     SP       PAT    CUM.P   CUM.D   CSH.FL   CDI    CDCF

1979   21.22   0.        0.      0.      0.      0.     21.22   0.
1980   21.22   0.25     1.79    1.79    1.77    3.55   21.22   2.91
1981   21.22   0.24     2.03    3.82    3.54    7.35   21.22   5.73
1982   21.22   0.26     1.76    5.58    5.31   10.80   21.22   8.10
1983   21.22   0.30     1.45    7.03    7.07   14.11   21.22  10.06
1984   21.22   0.32     1.62    8.65    8.84   17.58   21.22  11.92
1985   21.22   0.35     1.80   10.46   10.61   21.07   21.22  13.70
1986   21.22   0.37     2.00   12.45   12.38   24.83   21.22  15.39
1987   21.22   0.40     2.20   14.66   14.15   28.80   21.22  17.01
1988   21.22   0.42     2.42   17.08   15.91   32.99   21.22  18.56
1989   21.22   0.45     2.66   19.74   17.68   40.96   21.22  21.22
         **                                  *

INTEREST RATE OF RETURN (IROR) =  10.5%
PAYOUT PERIOD =   6.0 YRS AFTER STARTUP FOR R.O.I.
*INCLUDES SALVAGE VALUE AND RETURN OF WORKING CAPITAL.
     (I.E., SALV=    3.54 & ROWC=    0.  ).
**T.F.I. PLUS ANY COMPANY FINANCED WORKING CAPITAL,
     WHERE WORKING CAPITAL ESCALATES AT   7.0% PER YEAR.

     TCI   - Total Company Investment, $MM
      SP   - [redacted] Selling Price, $
     PAT   - Annual Profit After Tax, $MM
     Cum.P- Cumulative Profit, $MM
     Cum.D- Cumulative Depreciation, $MM
     CSH.FL- Cumulative Cash Flow (undiscounted), $MM
     CDI   - Cumulative Discounted Investment, $MM
     CDCF  - Cumulative Discounted Cash Flow, $MM
```

SLIDE NO. 12

In addition to this printout, the program prints all pertinent information for each operating year. The equation for calculating the discounted cash flow - interest rate of return is shown in Slide 13.

DISCOUNTED CASH FLOW ANALYSIS

FORMULA

$$FI^{*} + \frac{WC_{i+1}}{(1+I)^{K-i}} + \sum_{j=2}^{n} \frac{(WC_i - WC_{i-1})}{(1-I)^{K-1}} = \frac{WCE + S}{(1+I)^{N}} + \sum_{j=1}^{n} \frac{ATN_i}{(1+I)^{K}}$$

Where:

FI = Plant Investment (w/o working capital)
WC = Working Capital
WCE = Working Capital Last Year
S = Salvage Value
I = Discounted Cash Flow - Interest Rate of Return
j = Year Index Counter; j=1, 1st year plant operation
K = Year Index Counter; K=1, 1st year capital required
N = Total Number Years from Time Capital Required
n = Plant Service Life
ATN = Net After Tax Cash Flow for Operating Years

*FI may be input over several years as

$$FI_{K=1} + \sum_{K=2}^{a} \frac{FI}{(1+I)}K-1$$

Where:

a = Number of years for plant investment input.

SLIDE NO. 13

The DCF-IROR solution is by trial-and-error. The left hand side of the equation represents the discounted investment and the right hand side represents the discounted cash flows.

The ability to evaluate the real effect of inflation, tax credits and leveraged capital over the entire "life cycle" economics of a venture is an absolute necessity. For example, all of us realize that the inflation rate of petroleum based products is substantially different from non-petroleum products. The financial analysis must be able to demonstrate this effect on venture profitability.

Another example is the effect of leveraged capital. One of the governmental incentives being offered for motor fuel alcohol plants is loan guarantees. An example of how leveraged capital can affect a products selling price necessary to achieve an adequate interest rate of return is demonstrated in Slide 14. This slide is from the DOE report.

RAPHAEL KATZEN ASSOCIATES
AN 493
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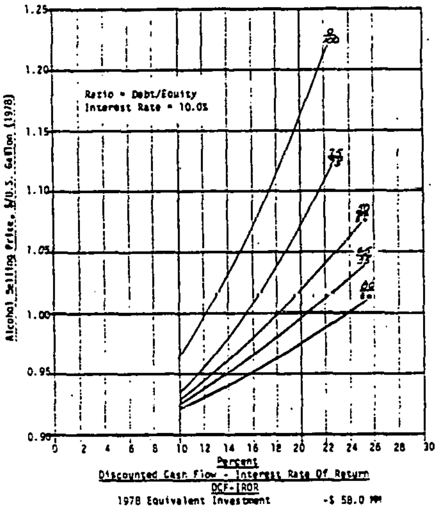
FIGURE 8-8

DEPARTMENT OF ENERGY
GRAIN MOTOR FUEL ALCOHOL PLANT

LEVERAGED CAPITAL
SENSITIVITY ANALYSIS
ETHANOL SELLING PRICE VS. DCF-IROR
20 YEAR PLANT LIFE
1978 COSTS



1978 Equivalent Investment -$ 58.0 MM
1981 Equivalent Investment - 64.0 MM
1983 First Year Operation - 7.9 MM

SLIDE NO. 14

For example, let's look at the effect of a 20% DCF-IROR. By increasing the debt to equity ratio from 0/100 to 80/20, the alcohol selling price, a 20% DCF-IROR is reduced from $1.17/gallon to $0.98/gallon, 1978 prices. The profitability in the leveraged capital case is based on a 20% company equity. The leveraged capital case has a lower selling price for a specific DCF-IROR due primarily to tax laws which permit interest payments to be deducted from the tax base profits; whereas dividends to investors cannot be deducted.

As you can see from the step-by-step procedure we have gone through this morning, the determination of commercial feasibility is a difficult task. Not only must one have a good procedure, but the cost factors must be representative of "real life" situations. It takes years of experience to develop the factors and techniques.

DEPARTMENT OF ENERGY
WASHINGTON, D.C.

GRAIN MOTOR FUEL ALCOHOL PLANT

TABLE 4-5
50 MM GAL/YR PLANT INVESTMENT
BASE CASE

Section No.	Identification	December 1978 Cost ($ MM)	Percent of Investment
100	Receiving, Storage & Milling	2,086,800	3.96
200	Cooking and Saccharification	2,824,300	5.36
300	Fungal Amylase Production	3,485,900	6.61
400	Fermentation	4,195,600	7.96
500	Distillation	5,123,800	9.72
600	Dark Grain Recovery	13,018,400	24.69
700	Alcohol Storage, Denaturing & By-Product Storage	4,399,900	8.34
800	Utilities	15,090,000	28.62
900	Building, General Services and Land	2,494,000	4.73
		52,719,000	100.00
	+ 10% Contingency	5,272,000	
	Total Plant Cost	$ 57,991,000	

DEPARTMENT OF ENERGY
WASHINGTON, D.C.
GRAIN MOTOR FUEL ALCOHOL PLANT
TABLE 4-6
50 MM GAL/YR - INVESTMENT BREAKDOWN
BASE CASE - 1978 COST

RAPHAEL KATZEN ASSOCIATES

		Equipment	Field Materials	Direct Labor	Construction Overhead	Sales Tax & Freight	Contractor Fees	Engineering	Buildings & Structures	Total
100 Receiving, Storage and Milling	E	725,600	182,400	244,000	146,400	49,800	29,300	250,000	200,000	2,086,800
	NE	259,300								
	T	984,900								
200 Cooking and Saccharification	E	32,100	302,900	593,000	356,000	56,700	65,000	247,000	450,000	2,824,300
	NE	721,600								
	T	753,700								
300 Fungal Arylase Production	E	0	401,500	665,000	400,000	94,100	81,000	350,000	225,000	3,485,900
	NE	1,267,300								
	T	1,267,300								
400 Fermentation	E	1,313,500	478,000	695,800	417,500	107,500	83,500	350,000	300,000	4,195,600
	NE	449,800								
	T	1,763,300								
500 Distillation	E	0	801,500	1,182,100	709,200	105,000	141,900	518,200	190,000	5,123,800
	NE	1,475,900								
	T	1,475,900								
600 Residue Feed Processing	E	64,200	1,079,400	2,308,000	1,150,000	348,500	215,000	1,200,000	600,000	13,018,400
	NE	6,052,300								
	T	6,116,500								
700 Alcohol Storage, Denaturing and By-Product Storage	E	1,200,000	580,000	957,000	574,200	95,200	111,800	385,000	300,000	4,399,900
	NE	193,200								
	T	1,393,200								
800 Utilities	PK	15,090,000	0	0	0	0	0	0	0	15,090,000
900 Buildings and General Services		0	0	0	0	0	0	0	1,744,000	1,744,000
Totals	E	3,135,400								
	NE	10,519,400								
	PK	15,090,000								
		28,344,800	3,325,700	6,644,900	3,753,300	856,800	733,500	3,300.700	4,009,000	51,969,000
									Land	750,000
									Contingency	5,272,000
										57,991,000

E = Erected
NE = Not Erected
T = Total
PK = Packaged Unit

TOTAL PLANT INVESTMENT (1978)

SLIDE #5

DOCEWSK DATA INPUT SHEET

	DATA	LINE DATA
100	JOB, TITLE	*100___,______________________
110	TFI,BASIS,DYRS,RINF,TAX,YR1,YRCA,ROIPC	*110____,____,____,____,____,____,____,____
120	NRM,NBYP,ISP,IDEP,IDEBT,ISWCB,IVTFI,IADDC	*120____,____,____,____,____,____,____,____
130	FEES,FYRS,MAINPC,TALPC	*130_____,_____,_____,_____
140	DEBT,DBYRS,YRL,IROL	*140_____,_____,_____,_____
150	IOWC	*150____
200	RM(1,1),RM(1,2),RM(1,3),RM(1,4)	*200_____,_____,_____,_____
210	RM(2,1),RM(2,2),RM(2,3),RM(2,4)	210_____,_____,_____,_____
220	RM(3,1),RM(3,2),RM(3,3),RM(3,4)	220_____,_____,_____,_____
230	RM(4,1),RM(4,2),RM(4,3),RM(4,4)	230_____,_____,_____,_____
240	RM(5,1),RM(5,2),RM(5,3),RM(5,4)	240_____,_____,_____,_____
250	RM(6,1),RM(6,2),RM(6,3),RM(6,4)	250_____,_____,_____,_____
260	RM(7,1),RM(7,2),RM(7,3),RM(7,4)	260_____,_____,_____,_____
270	RM(8,1),RM(8,2),RM(8,3),RM(8,4)	270_____,_____,_____,_____
280	RM(9,1),RM(9,2),RM(9,3),RM(9,4)	280_____,_____,_____,_____
290	RM(10,1),RM(10,2),RM(10,3),RM(10,4)	290_____,_____,_____,_____
300	STM(1),STM(2),STM(3),STM(4)	*300_____,_____,_____,_____
310	CW(1),CW(2),CW(3),CW(4)	*310_____,_____,_____,_____
320	ELC(1),ELC(2),ELC(3),ELC(4)	*320_____,_____,_____,_____
330	FUEL(1),FUEL(2),FUEL(3),FUEL(4)	*330_____,_____,_____,_____
340	PW(1),PW(2),PW(3),PW(4)	*340_____,_____,_____,_____
400	LABR(1,1),LABR(1,2),LABR(1,3),LABR(1,4)	*400_____,_____,_____,_____
410	LABR(2,1),LABR(2,2),LABR(2,3),LADR(2,4)	*410_____,_____,_____,_____
420	LABR(3,1),LABR(3,2),LABR(3,3),LABR(3,4)	*420_____,_____,_____,_____
500	BYP(1,1),BYP(1,2),BYP(1,3),BYP(1,4)	*500_____,_____,_____,_____
510	BYP(2,1),BYP(2,2),BYP(2,3),BYP(2,4)	510_____,_____,_____,_____
520	BYP(3,1),BYP(3,2),BYP(3,3),BYP(3,4)	520_____,_____,_____,_____
530	BYP(4,1),BYP(4,2),BYP(4,3),BYP(4,4)	530_____,_____,_____,_____
540	BYP(5,1),BYP(5,2),DYP(5,3),DYP(5,4)	540_____,_____,_____,_____
600	SALEPC,GAOPC,FBR(1),FR8(2),FR8(3)	*600_____,_____,_____,_____,_____
700	SELP(1),SELP(2),SELP(3)	*700_____,_____,_____
800	ICON,DTFI,WCR,AITCR,TFIT	*800_____,_____,_____,_____,_____

[illegible]ese lines must be supplied to DOCEWSK as a minimum even if you enter zero.

NOTES

INSTITUTIONAL CONSTRAINTS TO EXPANDED WOOD ENERGY USE

Charles E. Hewett

Resource Policy Center
Dartmouth College
Hanover, N.H. 03755

STRACT

nstraints arising because of the institutional mposition of our society will slow the estab- shment of wood energy systems as a substitute r imported petroleum fuels in stationary source plications. These constraints which occur in l phases of wood energy systems—harvest and ansport of the wood and construction and eration of cogenerating facilities—must be cognized by the nation's energy managers if ployment of wood energy is to take place at a sirable rate. This paper identifies institu- onal constraints to wood energy systems and oposes steps enabling their resolution. rticular attention is given to wood harvesting d construction of wood-fired power plants. The ergy community has failed to recognize that od, while a renewable resource, is a finite source around which interest groups have become rmly established. The Department of Energy ould support and encourage improved Forest rvice resource inventories to address the needs energy systems managers. Similarly, the operation of the pulp and paper industry and vironmental interests must be sought and tention paid to their concerns if wood energy ograms are to be successful. Inability to tain financing is the major constraint to the nstruction of wood-fired cogenerating facili- es. Three programs are proposed to address this oblem. In addition, inaccurate Environmental otection Agency air emissions factors for wood rning facilities may unnecessarily slow the nstruction of wood-fired power plants. Addi- onal research to validate more optimistic ctors in the literature should be conducted and e factors revised where necessary.

TRODUCTION

obviously comes as no surprise to those of you esent this morning that the United States, and deed, the western world, has entered an era of ansition from reliance on petroleum as its pri- ry fuel source to dependence on an assortment of her energy systems. The shortfalls in gasoline pplies in recent weeks and the narrow margin of ves of home heating oil last winter lend fur- redance to earlier signs of impending fuels scarcity reflected by the Arabian oil embargo of the early 1970's. Past experience with major energy transitions in the United States—from wood to coal in the last century and from coal to oil in this century—has indicated that 60 years are required to move the society from primary dependence on one source to another. We must, therefore, search for and develop transitional energy sources—fuels to tide us over while our cloudy visions of energy systems for the next century clear.

In our recent rush to discover "near term" energy sources and match them with their "appropriate end uses", we have stumbled onto wood--a fuel that grows in the forest, and must therefore be clean; a fuel that currently supplies at least one half as much energy nationwide as the nuclear industry and must therefore be successful.

If we overlook domestic use of wood to produce space heat which has grown dramatically in this decade, industrial generation of steam and cogeneration of steam and electricity are the most prevalent uses of wood to produce energy in this country[1]. Because the technology required for such processes is here and now, and many facilities are in place, most of us in roles as energy system analysts or managers have assumed that there are few institutional barriers to increased reliance on wood to produce steam and electricity. Because the technology is commonplace and therefore does not ring bells or blow whistles, we have not evaluated the constraints hindering the rapid deployment of industrial cogeneration systems—a deployment which makes sound economic sense in many cases, today, and which would displace far greater quantities of imported petroleum than would the production of substitute liquid fuels from the same amount of feedstock.

I want to talk this morning about the institutional constraints demanding recognition in four system phases of power cogenerated from wood: the harvest of wood, the transport of wood, construction of cogenerating systems, and operation of cogenerating systems. I will not talk about a fifth phase-electric transmission-where the institutional setting is rapidly evolving as a result of the Federal Energy Regulatory Commission's promulgation of rules under the Public Utilities Regulatory Policy Act.

HARVESTING

Few energy systems analysts or managers have achieved any significant degree of expertise in the production or procurement of wood from the nation's existing forest resource base. This has resulted in a number of cognitive failures among the energy community which have raised significant distrust among concerned onlookers. We have failed to recognize that wood, although a renewable resource, is a finite resource. To provide a sustainable fuel source, careful attention must be given to the rate of harvest of wood and to economically justifiable and environmentally sound methods of enhancing forest regrowth. I was amazed recently in a public meeting when a senior Department of Energy official argued that there would be no constraints on the wood resource base in the foreseeable future. I am continually surprised (though increasingly less so) by the reports coming forth from around the country which subtract the current regional harvest of wood from the net annual growth and assume that the remainder will be available for energy production.

Many factors affect the availablity of wood for energy production. Can it be economically harvested? Are landowners willing to sell it? Does the terrain allow harvesting? Is there a road network of sufficient capacity to enable chip vans to reach the resource once it is cut?[2]. I was delighted, yesterday, to hear Mr. Grosskreutz say that SERI will take a role in biomass resource assessments[3]. Current wood resource inventories conducted by the Forest Service simply are not adequate for sound projection of energy resource availability. Hopefully, SERI will be able to assist the Forest Service financially and perhaps also technically in updating its inventory techniques to meet the needs of the energy community. I hope that SERI will not design and implement new inventory techniques of its own. The establishment of another, separate, forest inventory team would be wasteful of the taxpayer's money.

The energy community has also failed to recognize groups with long standing interest in the management of forest resources and to incorporate these groups into energy production activities. The pulp and paper industry, for example, has been a largely monopolistic consumer of fiber from the nation's forests. I was amused recently when a representative of a company that some of you may know as the Jolly Green Giant spoke up in a wood energy meeting and argued that his company had a proprietry interest in the forest resource; and that it intended to maintain as tight a control on the forests on which it depends as possible.

But why bemoan this fate? The pulp and paper industry is one of the largest consumers of energy in the United States and it is largely dependent on #6 oil. Recent estimates, as I recall, indicate that the pulp and paper industry currently consumes about 5% of the total energy use in this nation. Currently, the industry is roughly 40% self-sufficient in its energy consumption. That is, bark and lignin supply 40% of the industry's fuel[4]. An intensive program to assist this industry in achieving a higher dependence on wood could go a long way toward the goal of doubling the nation's use of biomass announced yesterday by Bob San Martin[5]. Such a program could capitalize on the nation's primary source of expertise in wood harvesting, transport, and handling, and could displace the consumption of millions of barrels of imported oil each year. Wood energy programs capitalizing on the expertise of the pulp and paper industry will be far more successful than programs which pit the established industry against newcomers to the use of wood.

The energy community for the most part has also failed to recognize another important group with an established interest in the use of forests--the environmental community. Unfortunately, the energy market for wood is developing at a time of increased concern over non-point sources of water pollution (particularly erosion and water eutrophication stemming from timber harvesting), preservation of rare and endangered species, and the designation of wilderness. The environmental implications of harvesting wood for any purpose are highly interrelated, often site specific, and certainly management dependent. The energy community has concentrated on this latter point and hoped that environmental concerns over harvesting can be mitigated by the promise of sound management. They will not be. One recent study in Maine indicated that less than five percent of the harvest operations in that state used even the simplest methods of controlling environmental impacts[6]. The environmental community recognizes the merits of cogeneration from renewable resources, however, and could become a strong supporter of wood energy systems. However, rumblings from the environmental community where wood energy programs are relatively far advanced--for example, in Maine and Vermont—indicate that environmentalists intend to insure that the energy community pays more than lip service to sound harvesting techniques[7].

Another concern to this community, and more specifically, to many forest ecologists, are the unanswered questions surrounding the nutrient cycling implications of whole-tree harvesting. Whole-tree chipping is only a recently employed technology and the energy market promises to greatly expand its use. As a Department of Energy conference held in Illinois last year concluded, there is not a significant corpus of research able to be applied to answer questions regarding the nutrient implications of whole-tree removal[8]. Available research, however, does demonstrate that removal of an increasing portion of the tree, harvest of larger percentages of the stand, an[illegible] shorter cutting rotations will lead increasing[illegible] toward problems resulting from nutrient depletion

. The Department of Energy and others ocating wood energy use could achieve much ter support from our environmental institu- is by financially and technically assisting the est Service Experiment Stations and academic titutions with the capabilities to address such blems.

ther institutional problem relating to the vest of wood is the absence of a marketing tem. Problems associated with the establish- of a system to harvest wood from dispersed dholdings by independent contractors and ivery of it to an energy consumer with little no expertise in wood procurement are immense. re exists a chicken and egg phenomeon which may uire government support to break and which cer- nly deserves the attention of those who would wood energy systems established. The invest- required of a contractor desiring to provide d chips to the energy market now ranges between uarter and a half million dollars. It will tainly be impossible for contractors as they st today to obtain financing for such equipment hout guarantees of a sales place for the erial that they hope to produce. As I will cuss in more detail, later, however, the small ustrial concern seeking to institute a wood- ed cogeneration system finds it difficult to ure financing without a guaranteed fuel supply. ernment loan guarantees for one or both parties ld perhaps help overcome some of the problems. itionally, assistance programs to wood vestors that provided them with the organiza- nal skills necessary to operate successfully new, complex, and expensive harvesting systems ld help to ensure the stability of the fuel ply. Such programs might be supported by the but conducted by the Forest Service or the d-grant universties.

imates of the price of wood chips by those in energy community have also failed to provide potentail wood energy user with a reliable and est picture of the price he can expect to pay wood chips. Studies available to date have mined only the costs of harvesting and trans- ting chips and added a 10% profit margin for harvestor and trucker. Such studies overlook microeconomic laws of supply and demand, the lity of a large puchaser or producer to modify se laws, and the commodity cycles which inevi- ly surround a large market for any product. price of wood in the pulp and paper industry typically varied substantially depending upon market for paper, the vagaries of the weather, the ability of harvesting contractors to ply wood. Indeed, the price of energy quality ps in Vermont over the past year has varied m just over $12 per green ton to $26 per ton and k down to $16 per ton[11]. Imagine the dismay an energy producer new to the wood market ing that sort of variation. Attention should paid to this problem-I do not have the solution me today!

t remark regarding the harvest of wood to produce energy relates to competing demands for the resource over the long-term. Trend extrapolations are fraught with dangers and we all recognize the difficulties inherent in such projections. Nonetheless, it is useful to take stock of our current position when sonsidering the future availability of our resources. Over the past 50 years, the Pacific Northwest and California have suppied a disproportional share of the nation's lumber and plywood. In that region, as in the Northeast and Lake States before it, we have mined a resource that was indigenous when we arrived there; but soon the existing old-growth timber will be gone. The increasing rate of closure of Pacific Northwest mills is easily documented, and perhaps is highlighted by Georgia-Pacifics recent move of its corporate headquarters from Portland, Oregon, to Atlanta. This depletion of our old growth timber will cause increasing demand for the production of particleboards, fiber boards, and similar products to substitute for lumber and plywood., Indeed, the production and consumption of such products today is rising rapidly[12,13]. These materials depend for feedstocks on wood fiber very similar to that used for energy production. With current rates of technological development of new processes for the production of these materials, it would be naive to assume that they will not compete increasingly with wood used for energy production during the next thirty years. The growth in recreational use of forests, as reflected by National Park attendance[14,15] and recreational vehicle sales[16,17] and the growth in persons interested in conservation and preservation as evidenced by membership in organizations like the National Audubon Society [18] have been demonstrably exponential over the past two decades. I expect these forces will continue to place demands on forests which will restrict the production of wood from reaching its theoretical potential. Similarly, the increasing parceliazation of land in the east and conversion of woodlands to urban purposes nationwide will affect the availability of the wood resource.

While these projections may dampen the ardor of a group gathered to encourage biomass energy use, they need not do so. In my opinion, sufficient wood will be available to make a significant contribution to our energy needs during the energy transition period that lies ahead. It would be foolish, however to plan to rely on wood as a major energy resource beyond the next thirty years or so without taking substantial steps to secure sole rights to a large portion of the resource. Given the relative socioeconomic benefits of materials production from wood over energy production, I doubt such a policy would be in the best national interest when considered from a broad social perspective. Further, I doubt such a policy will be necessary to assure the United States of adequate energy supplies.

WOOD TRANSPORT

Transportation of wood chips from the harvesting site to the wood-fired power plant typically occurs in 80,000 pound gross vehicle weight chip vans. Problems associated with the transport of wood are highly dependent on the road network characteristics of the biomass harvest region and the plant site. In the harvest region, the presence or absence of roads and bridges of adequate quality to convey chip vans to within one-half mile of the cutting operation can make or break a wood-energy harvest under present economic conditions. The costs of roadbuilding to standards capable of supporting chip van traffic simply cannot be stood by a wood energy harvest unless they can be written off against some other purpose such as a high quality lumber harvest or recreational development.

At the plant site, the road network must be capable of accepting the incremental truck traffic converging from different points in the harvest region to the facility's woodyard. A wood-fired power plant producing the equivalent of 50 megawatts, for example, will require the arrival and departure of approximately 105 trucks per business day. An investigation conducted by the Dartmouth Resource Policy Center of 5 potential sites for such a facility in Maine found 3 with sufficiently adequate transportation networks to support the associated truck traffic. At these sites, the problems of increased noise, air pollution, and bridge and highway maintenance costs were also found to be acceptable. Two of the five sites examined currently experience extreme traffic congestion. At these sites, the addition of 210 one-way truck trips per day—an average of 15 trips per hour or one every four minutes—would have required major road network construction in heavily settled suburban areas[19].

A third problem associated with the transport of wood occurs only in specific regions of the country where roads become particularly fragile during the spring break-up period. Some states post roads to limit the acceptable axle weight of all trucks in this season. Difficulties caused by this practice can be overcome by stockpiling extra wood fuel during the winter harvest period or by accepting higher delivery costs associated with vehicles carrying less than their maximum capacities.

In short, the institutional problems affecting the transportation of wood may dictate the choice of particular locations for wood-using facilities but should not significantly slow the growth in the use of wood as an energy resouce.

CONSTRUCTION

Two major institutional factors may adversely affect the construction of wood-fired cogenerating facilties. The first problem, that of financing proposed facilties, may actually limit the ultimate quantity of energy produced from wood. In a survey of 100 potentail industrial uses of wood-fired boilers conducted by the Resource Policy Center at Dartmouth College, all respondents cited financing as the major constraint to their conversion to a wood energy system[20]. Financing difficulties arise from three main avenues of concern: First, investors simply lack confidence that wood fuel will be available on a sustainable basis to supply even small cogenerating facilities. Bankers, and bond brokers do not understand the marketing of wood fiber and regard it as a high risk activity. Second, wood energy systems, while now in many instances comparing favorably with other sources when the total cost of energy production is considered, do not compare favorabl on a capital cost basis with the financial community's current benchmark—oil-fired systems. This capital intensity, however, in most cases requires small industrial cogenerators to seek support in the financial community. A third problem in financing wood energy systems today is a difficulty facing most new projects-the prohibitively high cost of money.

To date, I am not aware of any programs providing assistance to industries wishing to convert to a wood energy system. Such programs might fall in three categories:

1) Refinement of techniques to inventory biomass on a regional basis. Inventories which considered valid social, institutional and economic constraints—coupled with an interprative assistance program to work with industry and the financial community—could greatly ease concerns which financiers have regarding wood supplies. Such a program would logically be directed from the regiona solar energy centers.

2) This suggestion is closely linked with th first. The regional solar energy centers desperately need experienced members of the financial community on their staffs to assis small industries in the assessment and analysis of the costs and returns of wood energy systems. This assistance should include aid in negotiations with the financial community as necessary. The Northeast Solar Energy Center has such a person on its staff workin with other solar applications. Such a perso is now needed who could devote the substantial portion of his time to biomass applications.

3) Even with successful programs as outlined above on line, I suspect that it may be necessary to develop loan guarantee programs if we are to see the rapid transition to woo energy systems that we desire.

The second major problem inhibiting construction of wood-fired energy systems is the requirement that air quality permits be in hand before construction of such a facility begins. There are a plethora of regulations which have been promulgated by the Environmental Protection Agency as a result of the Clean Air Act. A thorough expl[illegible]tion of these regulations would require sev[illegible] lectures in itself and even then might not be

le. Perhaps the most important of these
ulations are the Prevention of Significant
erioration rules which require a thorough
ient impact review and the utilization of Best
ilable Control Technology for any facility
tting more than 250 tons per year of any cri-
ia pollutant when operating with emissions
ontrolled and more than 50 tons per year when
rating with emissions controls. Strangely
ugh, particulate emissions which common myth
cribes as the major pollution problem from
d-fired boilers do not pose a real delay to the
struction of a wood energy systems. When con-
lled with existing technology, these emissions
m a 50 mw plant can be kept below 40 tons per
r,even when a boiler is fueled on bark alone.
ssions of carbon monoxide, nitrogen oxides and
rocarbons pose a frustrating problem, however.
ng the EPA's Compilation of Air Pollution
tors[21], one finds emissions from a 50 mw
ility as follows:

CO - 781 to 24,000 tons per year
NO_2 - 4000 tons per year
HC - 781 to 24,000 tons per year

iously, emissions of these pollutants appear to
far in excess of the 250 ton review limit.
s, use of these emissions factors in a license
permit application submits the applicant to the
sibility of a 4 to 12 month ambient air moni-
ing program followed by a computer modeling
lysis to determine ambient air impacts.

n in favor of preserving air quality—even at
stantial cost. However, there is now appearing
the literature increasing evidence that the
's volume AP-42 emission factors for wood
rgy systems are perhaps as much as two orders
magnitude high[22]. It is time that the DoE
iew this problem, conduct additional stack
itoring research if necessary, and encourage
EPA to revise its emissions factors where
ropriate. Such an effort could result in a
uction in preconstruction leadtimes of as much
12 to 18 months for cognerating systems
igned to produce more than about 5 megawatts of
rgy equivalent.

hird problem affecting the a potential con-
uction of a wood-fired facility results from
te site location requirements for industrial or
lity facilities. These regulations serve a
irable purpose in protecting the quality of the
ironment but may slow the construciton of wood
rgy systems, particularly by small industrial
sumers who may not be sophisticated in such
ters. The development of expertise to assist
se managers in completing these applications
ld speed the permitting process in many in-
nces. Funding for such a program might best be
ted in state energy or development offices.

RATION

e-construction operating permits are ob-
, there appear to be few additional institutional problems associated with the operation of a wood-energy cogeneration system. Ash diposal, however, may pose such a problem. A 50 Mw plant produces approximately 80 tons of wood ash per day—a quantity that would amount to about 800,000 tons if the facility were operated over a 30 year lifetime at a 90% capacity factor. While the composition of wood ash is benign compared with coal ash, it nonetheless does contain small quantities of several heavy metals which could contaminate water bodies if improperly landfilled. In the state of Maine where we conducted environmental analyses, the disposal of wood ash will create difficulties over the life of a wood-fired system. In that state, landfill space is not readily available because of the characteristics of its surficial geogogy. Additionally, the landfilling of any industrial material is currently a politically-charged issue because of several notable instances of ground water contamination. This situation is complicated by the fact that good data regarding the composition of wood ash from boilers operating under varying conditions with differing fuel mixtures are not readily available. Investigation of wood ash compostion from different combustion situations, particularly where fired with small quantities of back-up fuel such as oil, would be an inexpensive project which could ease the operating license procedures for wood energy systems. Further work also should be done to determine other possibilities for disposing of wood ash. Its economic potential as a soil conditioner, lime substitute, or with additives, as a fertilizer, should be investigated and would require only modest research funding.

SUMMARY

Constraints or hindrances to wood-energy use arise as a result of our society's current institutional composition in all phases of wood energy systems: harvest and transport of wood, and construction and operation of cogenerating systems. None of these constraints, in isolation or combined, will sound the death knell for wood energy systems. Together, however, they will slow industries' transition to increased reliance on wood for the production of cogenerated steam and electricity and will retard society's shift away from dependence on imported petroleum products. The appropriate solution and mitigation of the problems I have raised vary from constraint to constraint. However, they can be summarized in several broad categories: First, energy systems managers must recognize that such problems exist, identify them, and develop industrial assistance programs to smooth the path for companies wishing to install such systems. These problems are regional in nature and should be managed through the regional solar energy centers, the groups charged with commercializing solar alternatives. Second, the Department of Energy must intensify its efforts to develop interagency programs to reduce environmental uncertainties and resource ambiguity. Finally, where additional research is needed, the DoE must recognize that it is its mandate to commercialize the nation's alternative energy systems. Funds must be appropriated where necessary to provide answers to the institutional questions impeding the rapid development of wood as a fuel for stationary source use.

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PRELIMINARY CHARACTERIZATION OF TASE BIOMASS TECHNOLOGIES

J. P. Harper, S. W. Ballou, L. J. Habegger,
A. A. Sobek, and A. A. Antonopoulos
Energy and Environmental Systems Division
Argonne National Laboratory

STRACT

1978, the U.S. Department of Energy initiated
e TASE (Technology Assessment of Solar Energy)
ogram to assess the environmental consequences
increased utilization of solar energy. An
erview of the TASE Program and a discussion of
e biomass technologies characterized in Phase
vill be presented in this paper. Appropriate
omass conversion technologies were selected
r various biomass feedstocks (i.e., agricul-
ral and forestry residues, municipal wastes,
nures, and agricultural and forestry energy
ops). The technology characterization process
volved: description of a model system; input/
tput analysis of material and energy flows
to the system; estimation of the amount of
siduals (e.g., particulates, SO_x, etc.)
nerated during conversion; and estimation of
pital and operating and maintenance costs for
ch system. Results were computed on a per
ad (10^{15} Btu) basis and coded for input into
e SEAS (Strategic Environmental Assessment
stem) model. This discussion of the TASE
omass technology characterizations focuses on
ricultural residues, forestry industry resi-
es, and livestock wastes.

RODUCTION

ar energy technologies have recently become a
cus of intense research and development not
y because of their potential for replacing
nventional nonrenewable energy resources but
so because they are generally perceived to be
atively benign in terms of environmental and socioeconomic impacts. However, the use of solar technologies is not without externalities, especially when construction, material supply, employment requirements, and other secondary or indirect impacts are considered. The potential environmental and socioeconomic impacts, both positive and negative, resulting from a widespread application of solar energy technologies have not been adequately addressed in the past. Accordingly, the Office of the Assistant Secretary for Environment of the Department of Energy has initiated the Technology Assessment of Solar Energy (TASE) Program.

The purposes of the TASE Program are to examine on a national, regional, and community level the nature of the environmental and socioeconomic impacts likely to result from widespread use of solar energy technologies; to identify the potential for substituting solar technologies for conventional sources as a strategy for mitigating energy-related environmental and socioeconomic problems; to identify physical, environmental, and institutional factors that may limit the substitution of solar energy for energy from conventional sources; and to identify regional variations that may facilitiate development and use of particular combinations of solar technologies.

The TASE Program has two phases. Phase I provides a preliminary evaluation of the generic environmental impacts of solar energy technologies. Phase II of the TASE Program will draw upon the results and analysis of the Phase I effort to assess the environmental consequences of a national deployment of solar energy technologies.

One primary objective of the Phase I activity is to evaluate, on a per Quad (10^{15} Btu) basis, the amount of environmental residuals and the costs (capital, operating, and maintenance) of each solar technology system, from raw material extraction to end use. Results of these evaluations are the subject of this paper.

Additional objectives of the Phase I program are to determine, by means of an input/output analysis of selected solar demand scenarios, the raw materials and resources needed to manufacture and operate the solar and any ancillary systems; to examine the ability of communities to assimilate into their physical, social, and institutional structures increasing quantities of solar-derived energy; to assess the environmental and institutional impacts of solar energy use on spatially constrained communities; and to examine the environmental and institutional character of a community, under varying solar growth assumptions.

These objectives provide a basis for moving from generic assessment of impacts to the Phase II assessment that considers specifics of national and regional resource usage, cumulative environmental impacts, socioeconomic impacts, and institutional uncertainties. As a basis for the Phase II analysis, two national scenarios for level of solar deployment in the year 2000 are specified: a scenario with a high level of solar penetration similar to the Domestic Policy Review "maximum practical" scenario and a second scenario with lower penetration, assuming a general continuation of current energy policies. The Phase II assessments are being conducted jointly by the DOE and the Argonne, Brookhaven, Lawrence Berkeley, Los Alamos, Oak Ridge, and Pacific Northwest/Battelle Laboratories, with support from DOE's Division of Technology Assessment and MITRE Corporation.

TASE Phase I Approach

The broad scope of Phase I of the TASE Progr necessitated a division of the tasks required to effect the study. Eight technologies were selected for study, including some non-solar decentralized types. Los Alamos was primarily responsible for heating and cooling, agricultural and process heat, photovoltaics, and wind technologies. SERI, MITRE Corporation, and Lawrence Berkeley Laboratory also developed characterizations for these solar technologies. Argonne and Oak Ridge were assigned biomass technologies. Lawrence Berkeley was responsible for cogeneration and waste combustion. Oak Ridge was additionally responsible for district heating. The biomass technology was divided on the basis of resources; Argonne was to identify appropriate applications for crop and forestry residues, livestock wastes, and sewage while Oak Ridge was to define terrestrial energy crop applications.

Thus, Argonne's assignment for biomass technologies during Phase I was to identify end-use energy applications for various biomass resources Upon selection of a biomass conversion application, the following tasks had to be performed to characterize the chosen system:

> A technical description of the model system and the amount of end-use energy supplied;
>
> An input/output analysis of material and energy requirements of the model system per Quad of end-use energy;
>
> An assessment of the system's capital costs and operating and maintenance costs per Quad of end-use energy; and
>
> An assessment of operating residuals (e.g., pollutants) generated by the system per Quad of end-use energy.

ital' and operating costs were identified ı industrial sector in which expenditures de. This allowed inferences as to the sic raw material requirements for the application.

e integrating factor in the study is SEAS trategic Environmental Assessment System) ouse, 1977). This model permitted the integration of the technology characterization ta being developed at the various national boratories and other institutions, so that e national, regional, environmental, mateal, energy, and economic objectives are nievable.

AS is an extremely intricate system. Basicly, SEAS is a set of interrelated computer ograms that can model energy flows in the S., model the U.S. economy, calculate environntal pollutants, and provide these energy, onomic, and environmental forecasts at the tional and various regional levels. Thus, in SE Phase I the microlevel work on solar energy stems is melded via SEAS into macrolevel recasts.

e remainder of this paper will be on selected omass technology applications studied at gonne National Laboratory. This will provide representative perspective on the microlevel out for biomass technologies into the SEAS del. Because of space limitations, an adeate definition of all the assumptions and rspectives of our work is not possible and e readers are referred to forthcoming project ports; one deals with agricultural and fortry residue applications (Harper et al., 1979) i the other with livestock wastes and sewage llou et al., 1979). These biomass technology plications will be presented for agricultural sidues, livestock wastes, and forest industry sidues in this paper.

ICULTURAL RESIDUE APPLICATIONS

eral Soil Effects

haps one of the most significant environ- effects of using agricultural residues for energy is the deterioration of the soil by the erosive forces of water and wind. It has been reported that runoff almost doubles when crop residues were removed (Brady, 1974). The raindrop effect destroys surface soil structure, causing puddling, decreased water infiltration, increased runoff, reduced porosity, increased soil compaction, and a loss of organic matter and plant nutrients. In the semiarid wheat and cotton growing areas, the wind sorts the dry surface soil material by blowing away the fines, leaving only the semi-sterile skeletal matter (Lyles, 1975). This fine soil material contains a high portion of nutrients that are readily available to plants, and some experiments have shown eroded sediment to contain five times more organic matter and nutrients than the original soil.

Another major loss of soil nutrients is in removal of the residues themselves. Table 1 lists the amounts of major soil nutrients (nitrogen, phosphorus, and potassium) removed with corn, wheat, and cotton residues.

Table 1. Major Soil Nutrients Removed With Crop Residues

Crop	Nitrogen (lb/ton residue)	Phosphorus (lb/ton residue)	Potassium (lb/ton residue)
Corn Residue	22.2	3.6	26.8
Wheat Residue	6.0	0.5	8.9
Cotton Residue	34.8	4.5	29.4

Thus, the nutrients are often removed in excess of those applied, and the difference comes from the reserve previously built up in the soil. These nutrients must be maintained. Therefore, the total nutrients removed in the residue should be accounted for in the amount of fertilizer applied with the next corn crop to mitigate the adverse consequences of soil deterioration. In the analyses of technologies, the land areas temporary affected by residue removal and the additional fertilizer requirement have been incorporated into the analysis of the various crop residue systems.

Combustion of Cotton Ginning Residue

Ginning of cotton produced in the U.S. gives rise to three distinct products: lint (cotton fibers), cottonseed, and ginning residues which consist of leaves, sticks, stems, hulls, soil, and motes (Griffin, 1976)(Oursbourn et al., 1978). Residue production per bale ranges from 98 lb for spindle-harvested mid-south cotton to approximately 750 lb for Texas stripper-harvested cotton (Griffin, 1977).

The seed cotton at the plant must be dried before it can be ginned. Typically 430,000 Btu per bale (Holder and McCaskill, 1963) are required for drying. Recently, drying systems have been developed that use cotton ginning residue as a fuel instead of natural gas for drying the seed cotton. The mean dry-basis fuel value of cotton gin residues is 7032 Btu/lb.

The designs of systems for recovering gin residue and incinerating it to provide energy for drying have been reported by several authors (Lalor et al., 1976) (McCaskill and Wesley, 1976) (McCaskill et al., 1977). Researchers at the U.S. Cotton Ginning Research Laboratory have developed a system with 30% heat exchanger efficiency (McCaskill and Wesley, 1976). This system provided sufficient heat recovery for processing rates from 6 to 30 bales per hour, which spans production capacities for most cotton gins in the U.S.

For this study, the model system proposed for seed cotton drying proposed is simple: it consists of three basic operations, separation, combustion, and heat transfer. The model system does not have a gas scrubber on the incinerator so that a worst-case residual estimate is used to evaluate the conversion technology. The capacity of the system is 20,000 bales per year.

Based upon one-Quad output of useful process heat from the system (i.e., replacing natural gas), the major capital costs are identified in Table 2. The operating and maintenance costs for this application are approximately $3. billion (1972 dollars).

Table 2. Capital Costs Per Quad for Cotton Residue/Combustion System

Industrial Category	Billions of 1972 $/Quad
Industrial Patterns	7.4
Plumbing & Heating Equipment	7.4
Transportation	1.1
New Construction	2.3
Total Capital Cost	18.2

The major operating residuals for the application are found in Table 3. An environmental effect not noted in the table is the waste heat dissipated into the atmosphere. The amount of heat wasted by the system is many times the useful heat produced and accounts for the relatively high capital cost per Quad.

Table 3. Major Operating Residuals Per Quad for Cotton Residue/Combustion System

Residual Category	Estimated Residuals Per Quad (10^3 tons)
Particulates	9,150
SO_2	850
NO_x	1,750
CO	16,150
Hydrocarbons	1,700
Solid Waste/Ash	104,640

Hydrolysis of Corn Residue

Furfural is an aldehyde with the -CHO group in the α position. Its production was commercialized by Quaker Oats in 1922 (Quaker Oats Co., 1974) and it is obtained from pentosan-containing agricultural residues. Furfural serves primarily as a chemical intermediate for the production of furfuryl alcohol, tetrahydrofurfuryl alcohol, furan, tetrahydrofuran, polytetramethylene ether glycol, or as a precur

pyrrole, pyrrolidine, pyridine, piper-
e, lycine methylfuran, and many other
pounds. It is considered in this study as a
mass-derived chemical that can serve as a
stitute for a petroleum-derived chemical,
ereby potentially sparing oil resources.

extraction of furfural from agricultural
idues involves acid digestion under steam
essure followed by a series of distilla-
ns to separate and purify the furfural and
er by-products of the hydrolysis (Faith et
, 1957; Paturau, 1969). The material remain-
after digestion is then separated into a
id and a liquid fraction by a screw press.
solid residue is granular and composed
ncipally of modified cellulose, lignin and
ins (Quaker Oats Co., 1972). The liquid
ction contains the dissolved carbohydrates
spent acid.

furfural is recovered from the vapor by
sing it through a distillation column; the
rhead is condensed and a furfural layer and a
er layer separated by decanting (Faith et
, 1957). The water layer in the decanter
lds highly volatile by-products, namely,
hyl alcohol, methyl acetate, and acetic acid.
production of methyl alcohol and methyl
tate equals roughly one-sixteenth of furfural
duction at the plant while that of acetic
d equals the production of furfural (Lipinsky
al., 1977).

model system for this study is a large
othetical facility for furfural production.
processing rate is 135 tons/day of furfural
uiring 1588 tons/day corn residue (Lipinsky
al., 1977). Major energy and material inputs
o the system are steam (40,000 lb/ton
fural), water (11,000 ft^3/ ton furfural),
ctricity (250 kwh/ton furfural), and sulfuric
l (50 lb conc./ton residue). The capital
ts per Quad of input energy are given in
le 4. The estimated operating and main-
ance costs are approximately $21 billion
2 $) per Quad of end-use energy in the form
urfural.

Table 4. Capital Costs Per Quad for Corn Residue/Furfural System

Industrial Category	1972 $ Per Quad (10^9 $)
Special Industry Machinery	0.71
Fabricated Metal Products	6.15
Fabricated Plate Products	1.92
Plumbing & Heating Equipment	0.44
Structural Metal Products	3.06
Pipes, Valves, & Fittings	3.76
Trucking	1.21
New Construction	13.85
Total Capital Cost	31.10

The major operating residuals are presented in Table 5. It is assumed that the major high energy by-products of the process (e.g., methanol, methyl acetate, and acetic acid) are recovered and sold. This recovery would bolster the overall energy production by the system because these chemicals contain as much energy as the primary product (furfural). The major environmental contaminants are found in the air, water, and land media. A significant temporary land usage is also required to supply the residue.

Table 5. Major Operating Residuals Per Quad for Corn Residue/Furfural System

Residual Category	Estimated Residuals Per Quad (10^3 Tons)
Hydrocarbon	1,012
H_2SO_4 (Vapor)	140
BOD	26,880
H_2SO_4 (Acidity)	2,800
Solid Waste	392,240
Land, Permanent	321 (10^3 acres)
Land, Temporary	140,000 (10^3 acres)

Gasification of Corn and Wheat Residues

The pyrolytic conversion approach - gasification - is examined in this application for the production of a low-Btu gas from corn and wheat residues which can replace natural gas. Gasification is actually a two-stage process. Part of the biomass in the gasifier is combusted with a limited air supply, to provide heat to raise the temperature in the gasifier unit above 1600° F. Under these conditions, the biomass fuel in other parts of the gasifier is pyrolyzed. However, a highly carbonized solid residue, called char, and liquids with high tar contents are also generated. Resultant quantities of these materials depend greatly upon gasifier design and operation.

A commercial unit in operation at Diamond/Sunsweet, Inc., Stockton, California (Goss, 1978) served as a model for this application. The unit has a capacity of 28.6 tons per day and produces enough energy to sustain a steam production rate of 8500 lb/h (15 psi) at a fuel rate of one ton mulled walnut shells per hour. A natural gas pilot flame is required. A laboratory-scale testing model similar to the unit has been built at the University of California at Davis (Williams and Horsfield, 1977).

In the model system, there are only three unit operations: storage, handling, and gasification. Eighty percent of the input fuel is gasified and the principal by-product is a low-Btu gas (150 Btu/ft^3) with a char residue (20% of input material) as a by-product. The major energy flow into the system is natural gas for start-up or occasionally sustaining the process. Table 6 identifies the major capital expenditures for the gasifier system. These costs apply whether corn or wheat residues are utilized. The operating and maintenance costs for the system is $3.6 billion (1972 $) per Quad end-use energy.

The range of operating residuals which has been estimated for this application is given in Table 7. The temporary land use requirement for corn is significantly less than for wheat because of higher per acre yields. If the char from the process could be used as an energy resource, a significant improvement in overall energy generation by the system would result.

Table 6. Capital Costs Per Quad for Wheat-Corn Residue/Gasification System

Industrial Category	1972 $ Per Quad (10^9 $)
Fabricated Metal Products	0.81
Plumbing and Heating Equipment	0.35
Materials and Handling Equipment	0.02
Transportation	0.09
New Construction	0.25
Total Capital Costs	1.52

Table 7. Major Operating Residuals Per Quad for Wheat-Corn Residue/Gasification System

Residual Category	Residuals Per Quad End-Use Energy (10^3 tons)
Particulates	36-38
SO_2	3- 4
NO_x	52-53.7
Combustible Solid Waste	18,000 - 21,099
Land Use, Permanent	3.3 (10^3 acres)
Land Use/Year	49,000 - 56,000 (10^3 acres)

Fermentation of Sugar Processing Residues

Molasses is obtained as a by-product of sugar processing (both cane and beet sugars). U.S.D.A.'s Agricultural Statistics 1977 estimates that 133,676,000 gallons of molasses were produced from mainland cane and 191,700,000 gallons were produced from domestic beet sugars. Cane molasses is preferred by distillers because of the higher invert sugar concentration, although both types of molasses will support fermentation.

This model process developed for the production of ethanol from molasses is similar to other processes for industrial ethanol production in the United States and overseas (Lowenheim [illegible] Moran, 1975; Yang et al., 1977; Paturau, 19[illegible] pinsky et al., 1977; Klostermann et al., ; Wilke et al., 1978). The system has a uction capacity approaching 200,000 gallons ethanol per day and uses approximately 3000 ns/day of molasses. This system is scaled ter the large Battelle system for sugarcane ice (Lipinsky et al., 1977). The capital sts are summarized in Table 8. The operating d maintenance costs are approximately $13.3 llion per Quad of ethanol energy produced.

ble 9 lists the major operating residuals for e system. Air pollution (from distillation d storage) and water contamination are the jor problems. However, it is assumed that -products of the process (e.g., stillage) e marketable.

Table 8. Capital Costs Per Quad For Sugar Processing Residue/Fermentation System

Industrial Category	1972 $ Per Quad (10^9 $)
umbing & Heating	0.75
bricated Metal Products	1.87
mps, Compressors & Blowers	0.002
gineering & Scientific Equipment	0.01
ecial Industry Machinery	0.07
ructural Metal Products	0.10
tal Plate	0.09
pes, Valves & Fittings	1.17
ildings & Auxiliaries	0.90
ansportation	0.37
w Construction	4.29
tal Capital Costs	9.622

Table 9. Major Operating Residuals Per Quad for Sugar Processing/Fermentation System

sidual mponent	Residuals Per Quad End-Use Energy (10^3 tons)
hanol (Vapor)	391
nzene (Vapor)	16
[illegible]	58
Use, Permanent	54.8 (10^3 acres)

LIVESTOCK WASTE APPLICATION

Anaerobic Digestion of Manure

Anaerobic digestion of animal residues has been used in India for many years. Sanghi and Day (1972) report that in 1972 there were about 2,500 anaerobic digesters in rural India producing biogas from cattle dung without destroying the value of the solids as fertilizer. Essentially, only carbon is metabolized into biogas with a heating value of 600 Btu/ft^3. Most nutrients, especially nitrogen, pass through and are discharged in the digested residue.

In the United States, extensive investigations are underway to evaluate process requirements for biogas recovery from beef cattle, dairy cows, swine, poultry, and other animal wastes (Tennessee State University, 1977; Biogas of Colorado, Inc., 1978). One area where anaerobic digestion is close to full-scale production is the Four Corners region of Arizona, Colorado, New Mexico, and Utah, where there are many large beef feedlots.

The process characteristics reported here are based on a model plant designed to process the waste from 25,000 beef cattle. Beef cattle at an average weight of 1,000 lb produce 60 lb of raw manure per head per day, containing 6.9 lbs of total solids, of which 5.9 lbs is volatile solids.

The assumed design of an anaerobic digestion plant for feedlot manure has a process rate of 0.2 lb/day of volatile solids per cubic foot of digester capacity. At 5.9 lb of volatile solids per day per animal, 25,000 head of cattle will produce 147,500 lb of volatile solids, which will require 737,500 ft^3 of digester capacity. This capacity is supplied by three primary digesters 100 ft in diameter by 31 ft high. Three secondary digesters of the same size, with gas-holding covers, are also needed.

The digested sludge (42.5 tons of dry solids per day) is dewatered by 110-hp centrifuges, two on line and one in reserve. The extracted water goes to a controlled algae pond and later is returned to mix with raw manure entering the plant. The algae is harvested and dried as an animal feed supplement; the dewatered, digested residue is used as a fertilizer additive or solar-dried for animal feed.

Assuming a biogas production of 8.5 ft^3/lb of volatile solids, this digestion plant would produce a gross yield of 1.25×10^6 ft^3 of gas per day (Loehr, 1977). The digester gas is purified in an amine absorption step to remove 35% of the CO_2 and all of the ammonia and hydrogen sulfide, resulting in gas with 710 Btu/ft^3.

The capital costs for the process are summarized on a per Quad basis in Table 10.

Table 11 presents the major operating residuals generated on a per Quad basis. The major impact is solid waste from the digesters. This should not be a serious environmental problem if sound land application practices are employed.

Table 10. Capital Costs Per Quad for Manure/Anaerobic Digestion System

Industrial Category	1972 $ Per Quad End-Use Energy (10^9 $)
Cement Cover	2.08
Oil Field Machinery	2.68
Buildings	0.14
Transportation	0.20
New Construction	6.71
Total Capital Costs	11.81

Table 11. Major Operating Residuals Per Quad for Manure/Anaerobic Digestion System

Residual Category	Residuals Per Quad End-Use Energy (10^3 tons)
SO_2	2,570
NO_x	90
Solid Waste	103,700

FOREST INDUSTRY RESIDUE APPLICATION

Cogeneration from Pulp/Paper Residues

The pulp and paper industry is the fourth largest industrial consumer of fuels and electricity in the United States. Paper and allied products industries purchase more than 385 billion kWh equivalent of fuels and electricity per year (Little, Inc., 1976), and consumed (in 1970) 1.5 Quads (Sittig, 1977).

Pulp and/or paper operations are either larger integrated systems (producing both pulp and paper) or smaller nonintegrated systems (producing either pulp or paper). Most of the U.S. paper is manufactured by integrated mills.

The pulping process most prevalent in the United States and selected as the basis of our model forestry residue conversion system is known as the Kraft process. In the Kraft process, under high temperatures (350°F) and pressures (100 psi), the wood chips are cooked for 2-4 hours in an alkaline solution of sodium sulfate and sulfite in order to separate the cellulose fibers from the lignin and other materials (Hall et al., 1977) (Sittig, 1977). The spent cooking liquor, known as black liquor, contains the lignin and the chemicals utilized in the process. The liquor passes through a recovery system, and most of the original chemicals are extended and reused, while the remaining lignin plus other combustible materials are concentrated (55-65% solids). Thereafter, the concentrated liquor is conveyed to the furnaces/boilers and burned to generate heat, steam, and electricity.

It has been estimated that the production of a dry ton of Kraft pulp requires 325 kWh of electric energy as well as 8500 lb of steam at about 80 psi (equivalent to 10.67×10^6 Btu), and that one ton of spent liquor solids burned in the recovery furnace releases 13.2×10^6 Btu (Sittig, 1977).

he model plant studied has one power boiler ning bark and wood residues and two main ecovery boilers burning the organic matter f the black liquor. In addition, a turbogenrator unit is employed for electricity generaion. This unit is envisioned for a plant which nnually processes 200,000 tons pulp and producs 36,000 tons black liquor per year. The major apital costs on a per Quad basis are given in able 12. The operating and maintenance costs or the system are $0.13 billion per Quad.

Table 12. Capital Costs Per Quad for Pulp-Paper Residue/ Cogeneration System

ndustrial Category	1972 $ Per Quad End-Use Energy (10^9 $)
etal Plate	1.41
ngines & Turbines	0.31
ulp Mills	0.07
pecial Industry Machinery	0.20
ransportation	0.15
ew Construction	1.43
otal Capital Cost	3.57

able 13 gives the major operating residuals per uad. Air and land are the two resources most ffected by this technology.

Table 13. Major Operating Residuals Per Quad for Pulp-Paper Residue/Cogeneration System

esidual ategory	Residuals Per Quad End-Use Energy (10^3 tons)
articulates	3,145
O_x	105
O_x	699
O	2,097
ydrocarbons	2,446
olid Waste (Ash)	2,897

CONCLUDING REMARKS

The aforementioned biomass technology applications cover the range of conversion options available from combustion to anaerobic digestion. Typically, the capital costs and operating and maintenance costs for thermal conversion (e.g., combustion and gasification) are significantly less than for biological or chemical conversion (e.g., fermentation or hydrolysis).

Depending upon the technology, impacts to air and water can arise from the use of biomass conversion approaches. However, if by-products of some of the approaches are not utilized these impacts could be more significant.

Developing the right technological mix is also an important factor in assessing the large-scale utilization of biomass. No one application will be able to satisfy the future demand. Thus, identification of the appropriate technology mix is necessary.

The micro-scale analyses provide energy information on a per Quad basis, economic data, and residual data in an interrelated fashion. These data are now being processed through the SEAS model to ascertain the consequences of the large scale application of solar technologies.

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ENERGY FOREST CONVERSION OF MUNICIPAL WASTES

Dr. Geoffrey Stanford, Director
Greenhills Environmental Center
Dallas, Texas

rest systems provide an effective filter and ansing system for sewage waters, raw or treated. y also amass, store, and recycle micronutrients. ir saturation absorbtive capacity for e.g. toxic als, where they store them, under what circum-nces they release them, and in what forms, has n little studied.
pices (energy forests) offer a pleasant multipurpose solution to many of our environmental problems today. One of the processes which we propose is to use them as the ultimate sink for selected fractions of the urban refuse stream. These may be uncontrollably contaminated with variable amounts of toxic and hazardous materials. We consider the implications of this. The principle can be extended to farm and cropland; further benefits follow.

rs is an age of specialization. Yet, although it useful for specialists to identify trees as xonomic species, it is also informative for neralists to take the integrated approach: to udy the forest as an ecosystem. Today, in that irit, I will outline for you an integrated ergy approach to systems management of some of e fundamental resources of our human settlements: el, food, water, and wastes.

ch an approach is overdue, because our management sewage has been near-to-failure for many years, r management of refuse is approaching failure, r management of drinking-water is already failing many places, and our food supply system, both oduction and delivery, is in jeopardy as we try change back from fossil-fuel subsidies to a lf-sustaining solar husbandry. Fortunately the vironmental Protection Agency is gaining momentum its determination to integrate these several tivities, and is now showing that it has not only e will, but also the power, to support this licy [00.01].

wage wastes, sewage sludge, and the biodegradable action of municipal refuse [00.02] each contain a riable spectrum of plant growth adjuvants. ese include beneficial micro-organisms and the bstrates which support their growth, water, sential plant nutrients and micronutrients, tamins, and plant hormones and regulators.

e benefits obtained by using sewage waters and udges to fertilize and irrigate woodlands has en known for a long time [29.32]. Using raw fuse as a litter-fall extender or forest manure s been proposed [60.13], but I have not yet found ports of any forest trials. Here I will rein-rce that idea.

broad terms, I propose that a large fraction of e refuse stream has value as a manure: that it n be mixed with the sewage stream: and that both gether should be used to fertilize coppiced antations. By this program woody fuel can be intensively, even in regions where trees are not common, and drinking-quality water can be derived. I start by considering 'wastes' as virgin resources: then I will quantify their expected benefits as manure and, in particular, their ability to increase coppice growth; next I will summarize the foreseeable disadvantages of these procedures; and finally I will outline how to operate an integrated program on a city and regional scale.

NATURE OF THE RESOURCES

Refuse

Composition Refuse varies widely in overall composition by region, season, income level and local habits. Table 1 shows a representative breakdown. Its main value is that it reveals the very high fraction that is biodegradable, and the surprisingly high fraction (by weight) of metals, most of which are recoverable, and some of which are becoming scarce at the mines.

Value as 'Fertilizer' The value of municipal refuse as a 'fertilizer' in terms of N, P, and K is negligible. The N component is largely from the proteins in the waste foods: assuming a maximum of 50:1 as the C:N ration, and food as only 13% of the biodegradable fraction, N = 0.26%. P is derived mainly from incompletely empty detergent powder packets and animal bones; K is effectively negligible. The real manurial value of refuse is as a substrate for soil organisms: paper is finely comminuted trees, and forest litter is notably fertile. By using this substrate a great number of soil organisms can follow their life programs; some of these fix atmospheric nitrogen for their own needs; at the end of their lifespan this nitrofix joins the soil pool. One year after mixing raw refuse into the soil, it was found to contain twice as much nitrofix as the control [00.03].

TABLE 1. NOTIONAL COMPOSITION OF MUNICIPAL REFUSE(1)

dry weight, %

Paper & board	60	
Garden trash and wood	5	
Food wastes	10	
TOTAL BIODEGRADABLE		75
Iron	6	
Copper and brass	1	
Stainless steel	1	
Mixed alloys, aluminum	2	
TOTAL METALS		10
INERT: glass, pottery, plastics		15

(1) The actual composition varies so widely by region, season, income level, and other variables that these figures should be used with caution: any one contains a margin of some ± 50%.

Toxic and Hazardous Components Municipal refuse contains small amounts of a surprising range of toxic and hazardous materials. These include, for example: -

METALS: Hg, Cd, Ag, Pb. from batteries; Cr, Cd, Cu, Zn, from rust-proofing and plating; Cu, Sn, Pb from water-pipes; Hg, Wo, Cu from light bulbs and T. V. tubes; and even As from rat poison. Organics and their complexes come from biocides: monocot. and dicot. herbicides, pesticides and fungicides, unused pharmaceutical drugs, food preservatives and fly sprays; even explosives from fire-arm shells, spray canisters and liquid fuels. The quantities of any one of these averaged throughout the year may be insignificantly small; but when a shop dumps an outdated consignment the local concentration in the refuse mass can be very high.

Against this we must note that small amounts of nearly every metallic element are essential for one enzyme activity or another, and the refuse stream can play a valuable part in supplying those metals to the soil.

In the longer term both the variety and quantity of these toxic and hazardous materials will diminish steadily as the new Resource Conservation and Recovery Act (P.L. 94-580) comes into effect nationwide. Metal conservation will also be spurred as the costs of mining and of refining rise exponentially with the cost of energy, and recovery/recycle becomes more economic - as it always has been, for example, for gold and platinum.

Sewage

Each of us puts about 100 gallons of water into our home sewer-pipe every day. Table 2 shows the approximate load of body wastes. When kitchen sink garbage grinders are in use, the total load of paper and vegetable fiber is increased. Indirectl through the industries in our city, we use about ten times that volume. In most cities the home an factory piped sewage goes to the sewage works for treatment, while the runoff from the streets goes straight to the river or the sea. This is strange because they each contain about the same load of nutrients as sewage - and often an equivalent load of toxins also. The main value of sewage lies in its nitrofix content - the C:N ratio is around 40: as compared to about 1000:1 for refuse. So when sewage solids and the biodegradable fraction of refuse are mixed together, the mix has much the same composition as had the urban wastes at the end of the last century, before cars replaced horses, and cows were moved out of the city. Those wastes mad enormously fertile manure [B1.25, B1.28]. My thesi is that ours can also, and should be so used. Som people think that this would be bad, because the toxins would get into the food we grew on it. But what would happen if we put all of these wastes onto coppiced plantations and burned the chipwood for fuel?

TABLE 2. PRODUCTION OF SEWAGE SOLIDS, per person, dry weight

		Feces	Urine	Wastewater (detergent)	TOTAL	
TOTAL	gm/day	30	60			
Paper & Fiber	gm/day	15				
Elemental N	mg/day	1.5	15		gm/yr	6
P	mg/day	1.5	1	2500	kg/yr	1
K	mg/day	0.7	2.5		gm/yr	1

Quantities Available and Yields Expected

I will start from the assumption that a well-managed coppice, twelve miles square, can provide enough wood on a continual seven-year rotation cycle to fuel the generator for the present-day electricity consumption of 100,00 people [AC.37].

Refuse People produce about one ton of refuse pe year, of which 75% is biodegradable. If that is put onto each section of the coppice for the first two years of a seven-year cycle, five tons/acre is available. The expected harvest yield of chipwood is on average of the order of seven tons/acre/year or fifty tons/acre for each cycle. This figure contains two implications. Firstly, we expect fift tons output from five tons input - an energy gain of 10 x; this justifies the description of this program as a 'breeder reaction'. Secondly, we wil be replacing only 5% of the fuelwood export by a biological equivalent in compensatory input. This may prove in the long term to be insufficient to maintain productivity.

Sewage Sewage will be distributed onto the coppice during the first five years of the seven-year cycle. That provides a two-year interval befo harvesting, which will reduce any possibility

hogens being distributed on the harvested chip-
People produce 100 gallons of wastewater/
ll of which can be used for irrigation.
ting this on 5/7ths of their coppiced forests
five consecutive years would provide one and
half inches of water for each of those years.
t is negligible. But the contributions from in-
try may raise that by a factor of up to ten-
d, which is significant.

ld from the Waters If we assume that 1000
nds of water are evaporated from the soil and
nspired by the trees for every one pound of bio-
s net gain (annual increment), then the waste-
er produced by each person will provide 303
nds of biomass fuel each year, out of the total
seven tons that each of them needs to generate
ir electricity [AC.37]. That is negligible.
when we apply the 10 x factor, that becomes one
one half tons, which is a significant fraction.

ocedure

we accept that the program is sound in theory,
it be done in practice? I believe it can.
re is extensive experience already available on
merits of large-scale application of greenwater
land [00.04] for irrigating and fertilizing.
there is, as yet, very little experience with
refuse. We already know that it is impracti-
le to mix the raw waste directly into agricul-
al topsoil even after the metals have been re-
ed [00.03], and converting it to compost is
ll uneconomic for commercial farmland or for-
ry. I have been proposing for some years an
ernative technology, and hopefully a trial will
n be funded.

lraulic Sorting

increasing cost of finding and operating large-
le landfills, together with the increasing eco-
ic attractiveness of recovering metals from re-
e, are making it obligatory to install a refuse
edder at any new facility. Separating out the
als before shredding is done by hand and by
net. The shredded mix is again sorted by re-
ted passes through a series of air classifiers.
lighter fractions, which are blown further away
the airstream, are collected to be landfilled or
ned, while the heavier fractions are recovered
the metals separated out. Now the burnable
ction is, but for the plastics, also the biode-
dable fraction which we need for manure. It is
ally uneconomic to truck that from the recovery
ility to the fields, to spread it, and to till
under [00.03]. Piping it as a slurry is much
aper, and if we make the slurry with greenwater
will be of the order of 1:100 solids, which
ps easily. Now the proper program suddenly be-
es obvious.

THE LANDMIX PROGRAM ON COPPICE

Take the refuse to the shredder, and dump the entire output into a lagoon full of greenwater. The fraction which floats is mainly wood, plastic and fats. Recover the fats and oils, and return the residue to the shredder for further breakdown. The fraction which sinks is mainly metals, ceramics and glass. Dredge that out, separate and recover the metals, and use the residue for roadbed or concrete aggregate. The suspended fraction, mixed with the balance of the treated sewage stream, is pumped out to the coppice to irrigate and manure it. The waters percolate through the litter and topsoil, and are clean when they reach the aquifer. Any excess runs onto neighboring coppice land, with the same result. The waters that stay in the topsoil and percolate to the woodland streams are similarly cleansed. The area of land needed in coppice to fuel a conventional electricity generator has an ample safety reserve to absorb and process these waste streams continually.

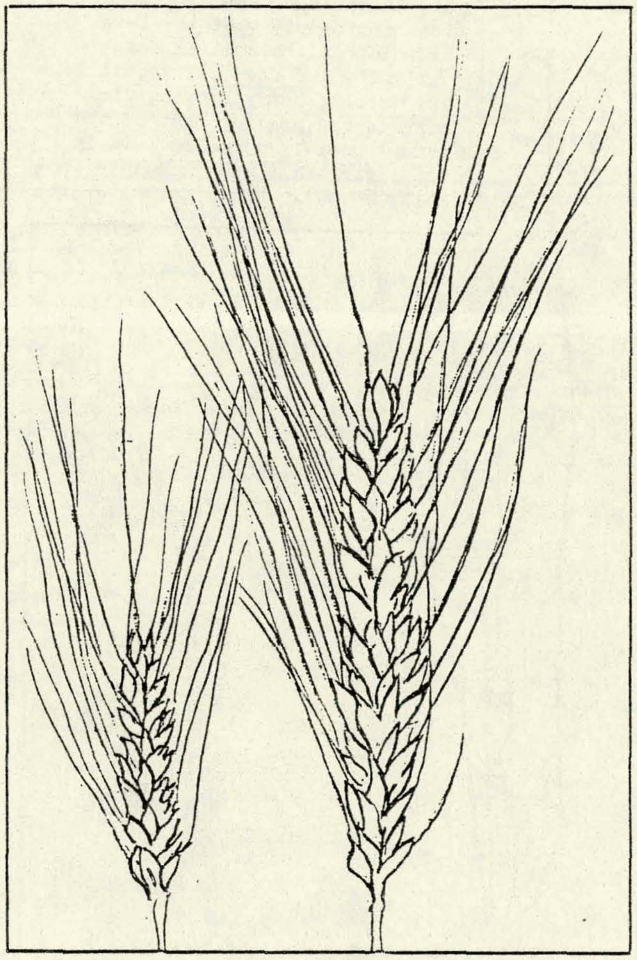

Fig. 1

Both these specimens are Turkey Red wheat. The smaller is grown with normal chemical fertilizer, the larger in soil into which 20 tons of shredded waste paper was ploughed into each acre, without fertilizer, three years earlier. Scale: x 3/4

Problems

It would be unreasonable to expect that this program would not meet some problems. The quantities involved are not too great - in fact, we do not have enough water or refuse to get anywhere near saturation (say 50 inches and 50 tons dry weight/acre). Spread of pathogens is unlikely, for most viruses and bacteria die in the open air, and the total elapsed time-scale attenuates spores and cysts. Metals may pass from the mix through the soil and into the trees, and then escape in the stack gases. I do not know of any published measurements about that possibility [00.05]. In alkaline soils most metals are precipitated and are taken up by plants only according to need. In acid soils they stay mainly in the microbial mass, and are taken in excess of need into the plants only when the level of organics in the soil is dangerously low and chemical fertilizers are forcing heavy growth. In forests the greater part of the metals pool is normally held in the litter-fall and immediate topsoil [80.44]. Under normal growth trees hold very little metals, although they need traces of about 30 mineral elements for their healthy growth [35.01]. So the toxic and hazardo materials presently found in the waste streams do not pose a significant threat to the environment in this program. From an engineering standpoint, the slurry may contain particles of glass too small to settle. They can be expected to abrade the pipelines severely; but we know from composting powdered glass in refuse that the spicules become smoothed by chemical action, so the problem may be less severe than one fears. The output nozzles may become blocked, but that is a well understood problem. Irrigation may not be possible every day of the year: holding-ponds of sufficient reserve capacity will take care of that.

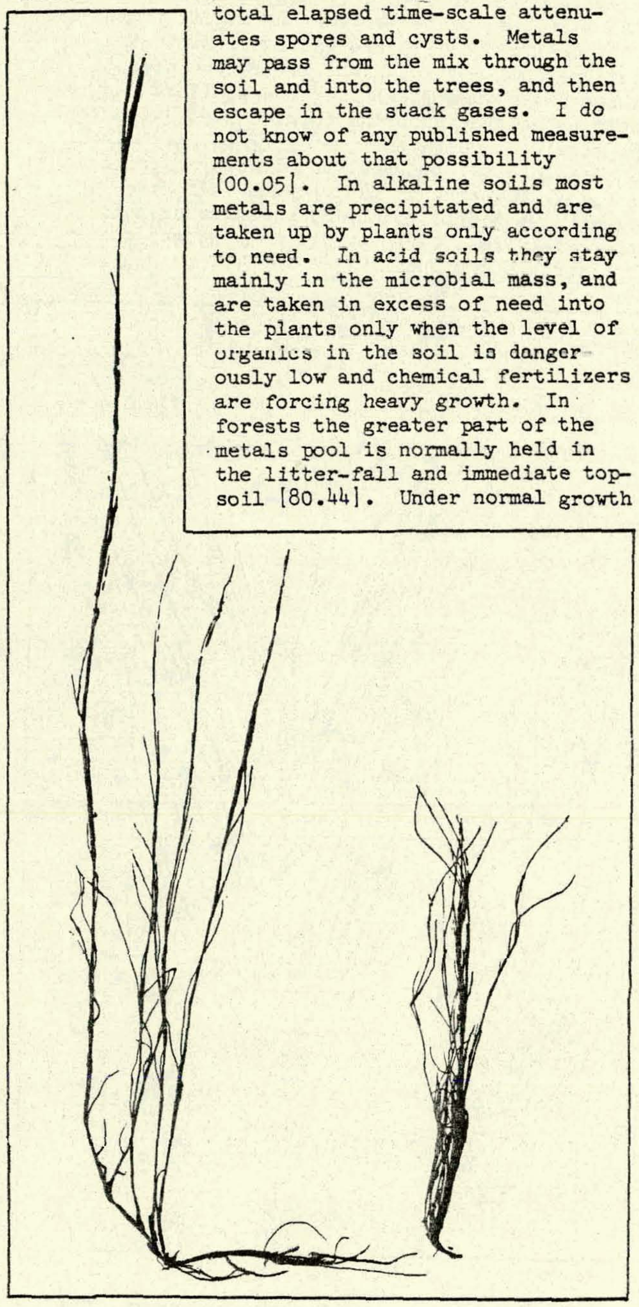

Fig. 2

These two specimens come from the same field. The smaller is not worth cropping for hay. The larger is grown in the same soil, into which 5 tons of unsorted municipal waste was ploughed the previous year into each acre. Scale: x 1/5

All-in-all, it seems likely that the combined benefits - social, economic, environmental, energetic - are so great as compared to present day costs and disadvantages that the few problems that arise will be small by comparison.

THE AGRO-INDUSTRIAL PARK

Once the benefits that can be obtained from using the total municipal waste stream to manure and irrigate the coppice-woods are appreciated, it is a logical step to apply them to agriculture in general.

The slurry is pumped out onto the fields, allowed to settle and thicken, and plowed or disced into the topsoil. The resulting high crop yields will encourage food-processing factories, feedlots, dairies, and similar enterprises to establish locally. Where better than near to this resource management complex? Their wastes will go straight into the same system hammermill and sewage plant. Back on the land, these will increase the area made fertile each year. Now what used to be the refuse disposal system and the sewage treatment plant together take on a new image: they are the hub of a thriving agro-industrial park (Fig. 3).

BENEFITS

Marginal land can be managed productively as coppice-wood for fuel, paper pulp, or chipboard. Manuring it with selected fractions of the municipal solid waste stream, and irrigating it with municipal wastewaters, increases its productivity, provide an environmentally safe system for ultimate disposal and recycling, and supplies clear run-off water.

The city which adopts the agro-industrial park program gains some welcome spin-off benefits. The combined costs of sewage treatment, landfill management, and drinking-water supply fall dramatically. There is an actively growing demand for labor, both from the factories and the fields. Th land which the city owns around the wastes management facility yields high rents, and the factorie pay taxes on top of that.

What more can you ask from municipal wastes?

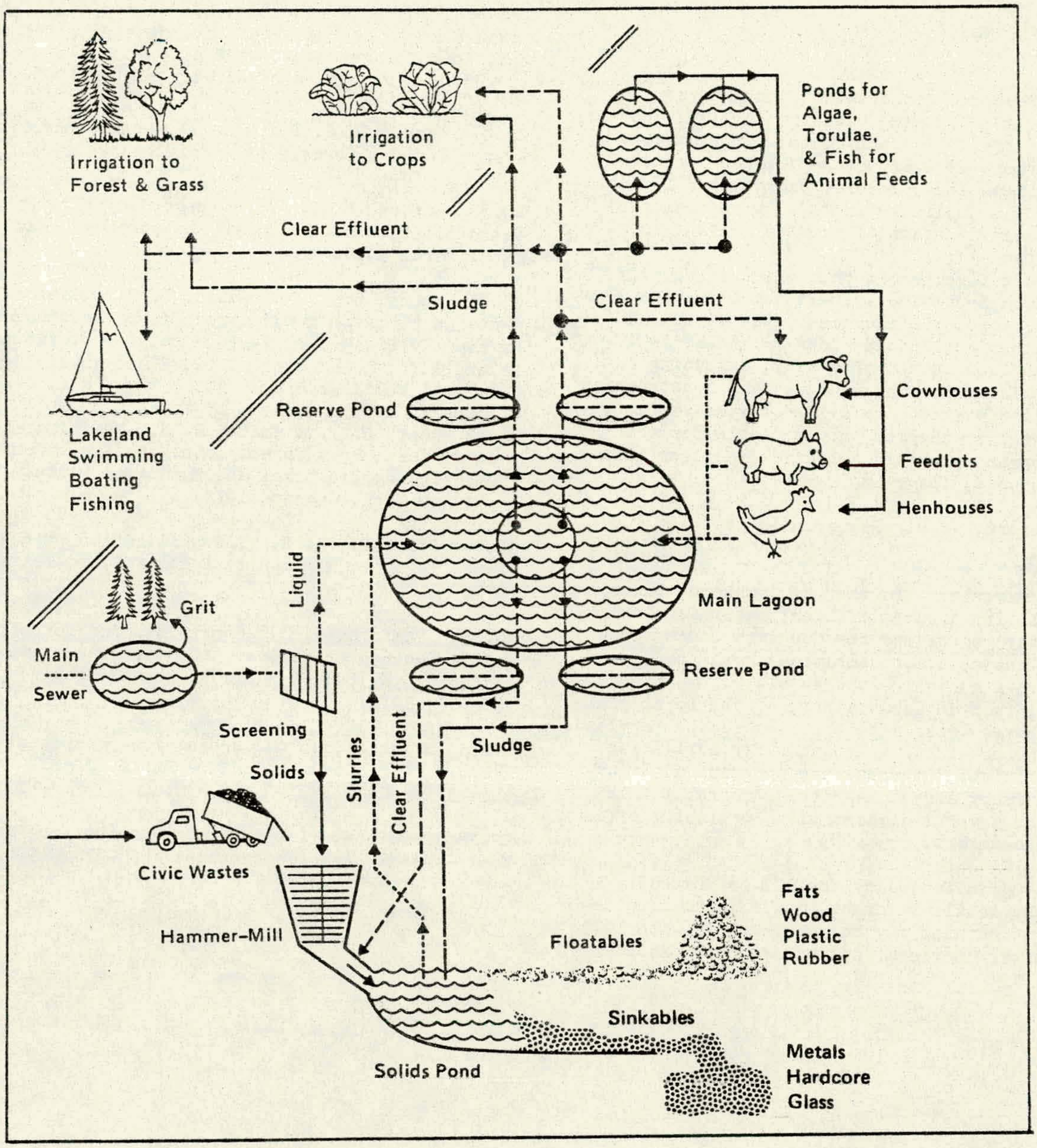

Fig. 3

THE AGRO-INDUSTRIAL PARK

This scheme shows a modified conventional sewage treatment plant in the center, the hydraulic refuse separating lagoon below, and the intensive husbandry facilities above right. Together these, with the associated factories (not shown), constitute the agro-industrial park. In the top left-hand area are the agricultural croplands which support, and are supported by, the park.

REFERENCES AND NOTES

00.01 The Resource Conservation and Recovery Act (1972), The Safe Drinking Water Act (1974), and The Clean Water Act (1977) each contain implications for integration. A draft concept paper for integration of these three federal acts with relevant state codes was published in 1978, with implementation foreseen as starting October 1979.

00.02 Municipal refuse is that which is usually collected from housewives, apartment complexes, offices and shops. It includes restaurants and cafeterias, but not institutions or industries. The total surface wastes of a city, including industrial solid wastes and sludges, and construction and demolition wastes, are called 'urban wastes'. For the most part they are not suitable for the landmix program, although selected fractions - from slaughter houses, for example, are.

00.03 Stanford, G. B. Research in progress funded by the EPA.

00.04 For example: - In the USA, Lubbock, Texas has used all its treated effluent water for irrigation for many years; and the Muskegon County program in Illinois, a more ambitiously planned program, is proving highly successful. It may come as a surprise, that in Texas alone, at the local scale, over 250 sewage works send their sludge to farmland, and over 150 use their greenwater for irrigating parks and fields. Overseas the programs at Braunschweig, Germany and at Werribee, Australia have operated for several decades with continuing efficiency and benefits.

00.05 A program funded by the DOE is presently studying the metals movement from soil to coppicewood. First results may be published in late 1979 [Wm. E. Sopper, personal communication]. Another project, funded by the EPA, has as one of its obj tives a study of the metals movement from refuse soil into forage crops, with and without sludge [00.03]. A report is expected early in 1980.

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MISSION ANALYSIS - MARKET PENETRATION MODELING

A Study of Biomass-Derived Fuels and Chemicals
Likely to Achieve U.S. Market Penetration in
the Years 1985, 2000, and 2020 (Contract No.
EY-76 C-0115PA131, Summary of Final Report Volumes #1 through #7

Prepared by:
Dr. Fred A. Schooley, Project Leader (415) 326-6200
Mr. Ronald L. Dickenson, Project Supervisor
Dr. Stephen M. Kohan, Thermochemical Analysis
Mr. Jerry L. Jones, Biochemical Analysis
Mr. Paul C. Meagher, Modeling
Mr. Kent R. Ernest, Feedstock Availability
Ms. Gwen Crooks, Process Comparisons
Ms. Katherine A. Miller, Product Demand Analysis
Mr. Wing S. Fong, Biochemical Analysis

study, conducted by SRI International for U.S. Department of Energy (DOE), Fuels from ass Systems Branch, was designed to examine rous technological processes for producing ul fuels and chemicals from agricultural s and residues in order to assist DOE in deining the level and type of federal support anted.

ass offers a significant potential for reng national dependence on imported fossil s through the conversion of a renewable en- source to useful liquid and gaseous fuels, tric power, process steam and chemicals. ral previous studies have indicated that feae national goals could be the production of t 5 quadrillion Btu (quads) of energy by year and 10 quads of energy by the year 2020. results of this study indicate that these are istic and achievable goals provided federal ram funding levels for biomass development increased and federal incentives are sucfully applied to increase biomass feedstock lability.

study involved the identification of over 0 possible missions (specific conversion es from biomass feedstock to useful fuel and ical products to end-use markets) prior to selection of 15 missions for detailed anis.

ectives

specific purposes of the study were:

1) To determine those biomass missions most likely to result in energy market penetration in the years 1985, 2000, and 2020;

2) To quantify the level of market penetration expected in those years;

3) To provide R&D Program recommendations for the U.S. DOE Fuels from Biomass Systems Branch.

Method of Approach

The method of approach used by the study team involved the projection of biomass feedstock availability by market price within U.S. census region and the development of a computerized model to estimate regional biomass fuel product market penetration in five year increments. This effort required the regional projection of market prices and demands for ten conventional fuels and chemicals as well as the derivation of biomass fuel product production data on 53 missions. Fifteen missions were examined in detail with the development of process flow diagrams, descriptions, economics, and energy and material balances. The costs of energy production for the 15 missions under regulated utility financing and a 65% to 35% debt to equity ratio are shown in Tables I and II. The optimistic estimates reflect high by-product values and product yields for the biochemical missions and a 20 percent reduction in base case capital costs for the thermochemical missions.

Results

Using the base case assumptions for feedstock availability (without federal incentives) 15 of the 53 missions penetrate the market by the year 2020, producing approximately 5.4 quadrillion Btus of fuel and chemical products, including electricity and steam. Assuming federal incentives and optimistic but achievable feedstock availability, 17 of the 53 missions penetrate the market by year 2020, producing approximately 10.3 quadrillion Btus of fuel and chemical products. The penetration expected by type of fuel for the "base case" and "optimistic" scenarios are shown in Table III.

The levels of market penetration shown are the result of computer model routines which simulate the competitive fuels environment and project usable energy demand as a function of energy price and historical market relationships. The model formulations allow a consistent comparison of projected market prices (marginal costs) with biomass derived product revenue requirements and compute product demand levels considering numerous factors, including mission commercialization dates, conversion process thermal

Table I. DETAILED MISSION ANALYSIS RESULTS: LARGE THERMOCHEMICAL FACILITIES

Route	Conversion Process[b]	Revenue Required ($/MM Btus)[a] Base Case $	Revenue Required ($/MM Btus)[a] Optimistic[c] $
Wood to:			
Heavy fuel oil	CL	5.37	4.76
Methanol	GOB	7.77	6.72
Ammonia ($/short ton)	GOB	164.00	141.00
SNG	GOB	6.41	5.56
Steam	DC	3.00	2.73
Electricity	DC	16.38	14.40
Steam and electricity	DC	3.42	3.06
Oil and char[d]	P	4.50	4.00

[a] Key: CL = catalytic liquefaction; GOB = gasification--oxygen blown;

DC = direct combustion; P = pyrolysis.

[b] 1977 dollars in year 1985. Data source, SRI Detailed Analysis--Regulated Utility Financing.

[c] Capital cost = 80% of base case.

[d] Char valued at $1.25 per million Btu.

Table II. DETAILED MISSION ANALYSIS RESULTS: LARGE BIOCHEMICAL FACILITIES

Route	Conversion[b] Process	Revenue Required ($/MM Btu)[a] Base Case $	Optimistic[c] $
Cattle manure to IBG	AD	4.87	4.37
Cattle manure to SNG			
100,000 head environmental feedlot	AD	7.02	2.75
10,000 head environmental feedlot	AD	14.44	7.50
Wheat straw to ethanol	F	52.60	29.20
Sugar cane to ethanol	F	26.95	20.00
Kelp to SNG	AD	20.70	10.70
Algae to ethanol	F	26.90	17.40
Wheat straw to IBG (40% conversion)	AD	23.76	9.00

[a] Key: AD = anaerobic digestion; F = fermentation

[b] 1977 dollars in year 1985. Data source: SRI Detailed Analysis — Regulated Utility Financing.

[c] High by-product values and product yields.

Table III. MARKET PENETRATION- -BIOMASS PRODUCTS

Product	Estimated Biomass Derived Products Quadrillion Btu (Excludes Existing) 1985	2000	2020
<u>Base Case Scenario</u>			
Gaseous products (SNG, IBG LBG)[a]	0	.13	.29
Methanol/Ethanol	0	0	0
Ammonia[b]	.02	.40	.56
Process steam or steam/electric	.68	2.32	4.01
Pyrolytic fuel oils	.04	.61	.53
Total Quads	.74	3.46	5.39
<u>Optimistic Scenario</u>			
Gaseous products (SNG, IBG, LBG)[a]	0	.21	.45
Methanol/Ethanol	0	0	0
Ammonia[b]	.13	.41	.58
Process steam or steam/electric	1.10	4.17	0.39
Pyrolytic fuel oils	.04	.87	.83
Total Quads	1.27	5.66	10.25

[a] SNG = Synthetic natural gas; IBG = Intermediate Btu gas; LBG = Low Btu gas.

[b] Assumes 18.3 million Btu/ton of ammonia

:ies, and the time lag between technology
tion and widespread technology implemen-
However, the relationship between ex-
biomass product price (revenue required)
d alternative fuel and chemical market price is
e most important factor in determining annual
rket penetration levels. See Appendix for the
sults of the sensitivity analyses for each of
e 15 missions, showing the effects on costs of
anges in operating factors, feedstock prices,
cility life and facility size.

rket Penetration Model

e methodology underlying this study's compara-
ve biomass assessment is based in part on pre-
ous SRI work in the field of energy market
alysis. In particular, this investigation has
awn on previously developed analytical ap-
oaches to the problem of forecasting the ex-
cted market potential of newly introduced
ergy technologies and commodities. To aid in
ese past analyses, a computer model was
veloped and applied to the evaluation of market
tentials for various solar technologies and
nthetic fuels. However, the level of model de-
il necessary to investigate specific biomass
ssions has required further modeling effort.
is work has focused on the development of a me-
odology that uses an iterative process to con-
rge toward equilibrium biomass supply/demand/
ice conditions. The following paragraphs
scribe model inputs, summarize the methodology,
d discuss details of the market penetration
rmulas contained within the model.

put Data

e data required for operation of the model con-
st of the resource availabilities of the
rious types of biomass feedstocks, the process
onomics of biomass conversion options, a frame-
rk of energy demands and market prices, and a
t of three parameters that are used to describe
e interaction of the biomass-derived products
th the markets in which they compete.

omass Resources

cause of the unique character of biomass re-
urce availability, computer modeling requires a
re sophisticated approach than that suitable
r other types of resources. Unlike other
ergy sources, biomass is both a renewable and
pletable feedstock at the same time. It is re-
wable over long periods and depletable in the
ort term because of restrictions such as length
the growing season and rate of residue genera-
on.

omass availability is most readily described in
manner similar to that typically used for
ssil fuels by the use of a curve that considers
e resource quantity available as a function of
ce. Unlike fossil fuel resource curves, how-
the biomass curves vary in time, reflecting changes in the expected future availability of biomass feedstocks from residues and energy farms.

Five major biomass feedstocks are considered in this analysis--low moisture plants, high moisture plants, woody crops, manure, and marine crops. For each of these categories, project final report Volume III describes a set of resource curves, corresponding to the years 1975, 1985, 2000, and 2020--the time frame of the analysis. Prices for the intervening years are found by interpolation within the computer model.

Conversion Economics

The process economic data required for biomass conversion describe the costs and efficiencies required in producing an energy product (such as SNG) from a biomass feedstock. This information determines the production price for any biomass product at any specific biomass feedstock cost.

Specifically, the computer model cost inputs consist of a specific capital cost (SCC) in the units of dollars per million Btu of biomass-derived product per year and a operating and maintenance cost (M) in dollars per million Btu. The efficiency input parameter, e, specifies the amount of product energy obtainable per unit of energy contained in the biomass feedstock. These parameters are supplied for each feedstock-product mission under consideration, as described in final report Volume II, Section V.

In determining the overall product price, a capital recovery factor (CRF) is first applied to the SCC to obtain a capital charge in dollars per million Btu of product energy. The maintenance and feedstock costs are added to this, resulting in the following product cost equation:

Biomass product production price =

$$\text{CRF} \times \text{SCC} + \text{M} + \frac{\text{feedstock cost}}{e}$$

where

CRF = capital recovery factor
SCC = specific capital cost
M = operating and maintenance cost
e = efficiency factor.

Expected Energy Demands and Prices

In order to perform a market penetration analysis, a framework of energy product demands and market price projections in which the biomass-derived products compete must be assumed. For each of the markets under consideration, an overall energy demand sets the size of the available market open to the biomass products. A projection of the alternative (nonbiomass) fuel market price also is given as an estimate of the competitive environment that the biomass-derived product will face. Both the product demand and the market price projections are input as a function of time and region, as described in final report Volume II, Section IV.

Market Penetration Parameters

The market penetration parameters are input to the model to characterize marketplace behavior. Three different parameters are used—market share, the behavioral lag half-life, and the behavioral lag response. Their definitions and uses are described later in this section.

Methodology

This market penetration analysis uses an iterative process that converges toward equilibrium biomass supply/demand/price conditions. Supply/demand/price equilibrium is defined here as the situation in which the production price of the biomass feedstocks is at the levels necessary to match the supplies of feedstocks with the demands for them. Figure #1 is a flow diagram of the procedure.

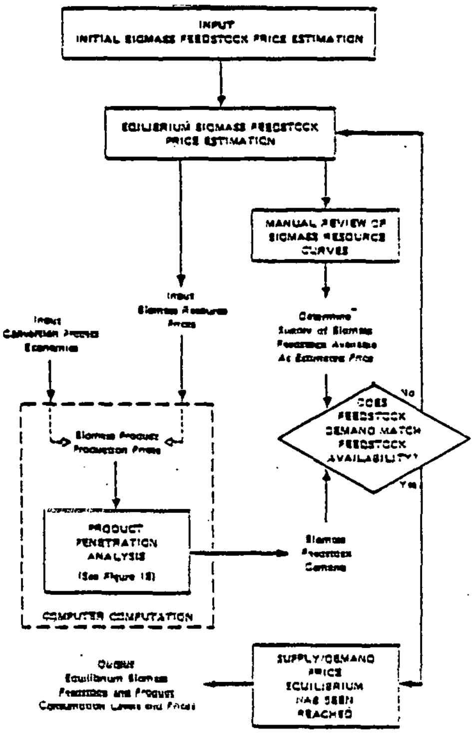

FIGURE 1 OVERALL METHODOLOGY OF BIOMASS COMPARATIVE ANALYSIS

Estimates are first made for the equilibrium marginal production prices of each major biomass feedstock as a function of time and region. Because of the relationships between feedstock price and quantity of feedstock available (as described by a set of biomass resource curves), doing this also determines estimates for the equilibrium supply quantities of each major feedstock.

Using the initial feedstock production pric the model calculates the equilibrium margi production prices of each biomass-derived product as a function of time. (The equation shown earlier is used to find these prices). Product penetration estimates can then be made for each biomass product and the level of demand for each of the feedstocks can be evaluated for the estimated feedstock prices.

To test for equilibrium conditions, the demand estimates for each major feedstock are compared with the supply quantity estimate obtained earlier. On the first iteration through this procedure, supply-demand mismatches possibly will occur for various feedstocks and at various time periods. These discrepancies indicate that the equilibrium feedstock prices have not been accurately estimated and that adjustments must be made.

If the calculated demand for a particular feedstock exceeds the quantity available at the estimated feedstock price level, for example, then the estimated price is too low and should be adjusted upward. On the next iteration through the procedure, one of the effects of this adjustment will be to make a larger feedstock supply available, as determined by the appropriate biomass resource curve. Another effect will be to push the biomass product prices upward, causing a reduction in demand as calculated by the product penetration analysis. These two effects will push the biomass feedstock supply and demand levels toward closer agreement. At this point, the direction and size of the mismatch is agair evaluated, feedstock prices are again adjusted, and another iteration of the process is performed. After several such iterations, the equilibrium supply/demand/price condition will be reached for each feedstock and for each time period chosen.

The results obtained from the final equilibriu iteration will be those that are useful to th overall biomass mission analysis. Most importantly, they will provide estimates of regiona demand for each major biomass feedstock and eac major biomass-derived product.

Product Penetration Analysis

The basic steps followed in the product penetration assessment are presented in Figure #2. Thi procedure first entails a static economic analysis based on the competition between the biomass-derived product price and the market price. Th result is a steady-state market share, reflectin a situation that would be expected to exist afte a period during which the competitive economi forces remained constant. Steady-state conditions do not hold soon after the introduction o a new, cost-competitive mission, technology, o product. The marketplace will be in a stat flux as the newcomer gains wider recognitio acceptance.

is dynamic market behavior. is modeled as a ioral lag" constraint, which reduces the at which a new mission or technology may be troduced. With this information, a dynamic omass product market share can be generated and plied to the product demand forecast to obtain estimate of the potential biomass mission netration as a function of time.

this methodology, the results produced are not teractive with the basic framework of energy mands and alternative prices in which the technologies compete. Thus, the assumed market dends and prices are not directly perturbed by e biomass mission market penetration, a valid sumption as long as the potential biomass ssion demand does not become too large a share the total demand.

eady-State Market Share*

an idealization, the share of a particular rket that a single new technology or product n attain at any particular time under steadyate conditions can be represented by the curve own in Figure 3 and given by:

eady-state market share to biomass-derived

$$\text{oduct} = \frac{1}{1 + (\frac{P_b}{P_m})^{\gamma}},$$

ere P_b and P_m are the marginal price of the omass product and the marginal market price, spectively.** This static representation says t when P_b and P_m are equal and under steadyte conditions, the market will be shared

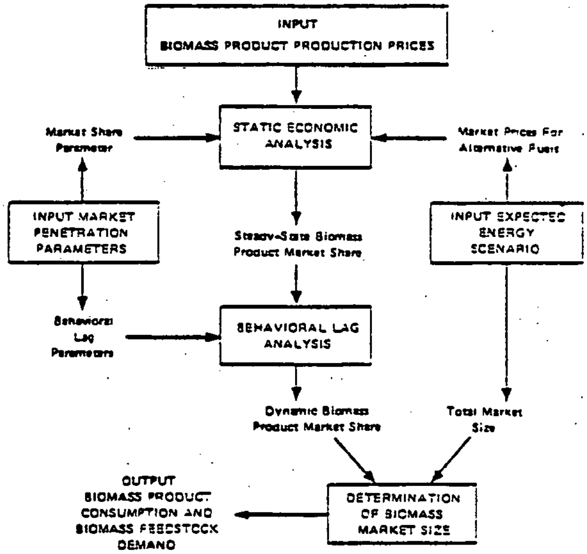

FIGURE 2 LOGIC USED IN BIOMASS PRODUCT PENETRATION ANALYSIS

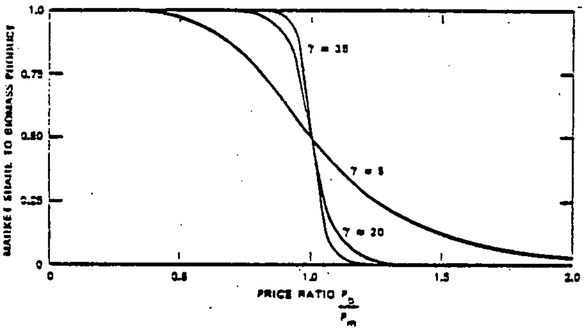

FIGURE 3 STEADY-STATE MARKET SHARE

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efinitions for this and other terms are given at the end of this section.

he market penetration model converts the average market prices presented in Section IV into arginal prices before use in the steady-state market share analysis. The equation usef for this urpose is:

$$_{m_t} = \frac{0.25(P_t + P_{t+1})(D_t + D_{t+1}) - 0.25(P_t + P_{t-1})(D_t + D_{t-1})(1 - \frac{\Delta T}{X})}{0.5[D_{t+1} - D_{t-1} + (D_t + D_{t-1})(\frac{\Delta T}{X})]}$$

where

P_m is the marginal market price

t denotes the time period under consideration

P is the average market price

D is the total demand for the market

ΔT is the size of the time intervals used in the analysis (5 years)

X is the lifetime of the energy facility or production unit used in the arket.

equally. The market share parameter (Y) is a measure of market imperfections, price variations, and consumer preferences.

When two or more new technologies are competing for a share of the same market, a more general market share formula must be used. For example, if N different biomass missions all produce the same product (such as SNG), then their respective market shares would be represented by the following equation:

$$\text{Steady-state share to } N^{th} \text{ Biomass Product} = \frac{1}{\left(\frac{P_{B_N}}{P_{B_1}}\right)^{Y} + \left(\frac{P_{B_N}}{P_{B_2}}\right)^{Y} + \ldots + \left(\frac{P_{B_N}}{P_{B_N}}\right)^{Y} + \left(\frac{P_{B_N}}{P_M}\right)^{Y}}$$

where P_{B_1} through P_{B_N} represents the prices of the first through the N^{th} biomass products and P_M represents the expected market price. If all of the prices P_{B_1} through P_{B_N} and P_M were equal, each biomass product would receive 1/(N + 1) of the market.

In this representation, a single representative marginal price is used for each biomass product and alternative fuel. Actually, significant individual variations from these representative prices do exist. For example, continued governmental regulations might introduce significant disparities into the future marginal prices of pipeline natural gas. Also, other sources such as LNG imports and various high-Btu synthetic products will be introduced at yet different prices, causing an even wider distribution of prices about our assumed representative marginal price.

The market share parameter is used to model such price variations.

Decision-makers also influence market share. Even if a new technology is somewhat more expensive than the alternative, some fraction of purchasers will choose it, perhaps because of environmental or "energy independence" considerations. Alternatively, some fraction of purchasers will continue to use their familiar fuel source even if economic considerations dictate a change to a new one. Imperfect price information also affects the market share curve. These various factors are aggregated into the one market share parameter. In a perfect market with high price elasticity and none of these real world effects, Y would be infinite, and the product with even a very slight economic advantage would obtain a 100 percent steady-state market share.

Based on previous work with energy commodities, SRI has found that the characteristic response patterns of various markets can be modeled by a suitable choice of Y. Large industrial and utility markets, for example, would generally be modeled with a Y value in the range of 25 to 35. These high values reflect the strong response to price variations by industrial and utility consumers who deal with large quantities of energy and are acutely aware of economic c siderations. The Y value used for this mark penetration analysis was 20, reflecting the somewhat larger degree of market imperfection that would be expected to obtain in markets that are available to biomass-derived products. Even lower Y values would be applicable for smaller scale energy consumers, who typically behave in a less strictly economic fashion. Factors such as aesthetics and novelty, for example, would be expected to play a much larger role in the decision-making process of a residential consumer than an industrial buyer.

Dynamic Market Response

The market share curve in Figure #3 is only a static representation. To assess the dynamics of market penetration, a "dynamic market response curve" can be used to describe how fast the current market will move toward the static price-determined market share curve as a result of real-world behavioral response. This is called the behavioral lag effect.

The dynamic market response curve, given by:

$$\frac{1}{1 + \left(\frac{h}{n}\right)^{\alpha}}$$

is shown in Figure #4, where

h = behavioral lag half-life (time required for one-half of the market to respond to the entrance of a new product)

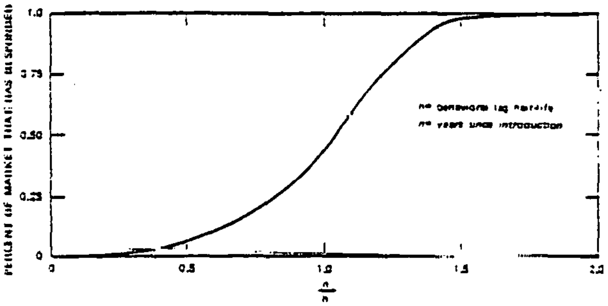

FIGURE 4 DYNAMIC MARKET RESPONSE

n = years since new product introduction
α = behavioral lag response parameter.

This curve tends to slow the introduction of new technology based on the time that it take for market decision makers to accept and switc to the new product.

The behavioral lag parameters, h and α, provide means of quantifying the dynamic marke response. A half-life of 10 years was chosen fo this analysis, allowing for a maximum of 50 percent market penetration at a point 10 years the date assumed for commercialization.

d behavioral lag parameter, α, fixes the re- shape (curvature) of the dynamic market ponse curve once the half life parameter has :n chosen; a value of 4 was chosen for this :ameter.

find the share of the open market captured by new fuel in any particular year, the equilib- im market share and dynamic market response :ve are multiplied. This is done on an annual sis, resulting in a dynamic market share for : new product that varies with time. The :ual biomass product demand is then found as a iction of time by applying the dynamic market ire to estimates of the size of the market that available to the new product.

tputs

e market penetration calculations are performed five-year intervals over the time frame of the alysis. However, particular emphasis is placed the analysis of three periods--the year term 985), the intermediate term (2000), and the ng term (2020).

e results of the penetration analysis provide timates of the expected market penetration (in ads) of each mission by region and time inter- l. The expected demands for each type of feed- ock and the equilibrium marginal prices for omass feedstocks and products are also deter- ned.

commendations

e current programs of the Fuels from Biomass anch are directed at specific research develop- nt and demonstration (RD&D) goals in both the omass production and conversion areas. In both eas, processes, techniques, and technologies ist in which there is near-term potential to celerate commercialization. Therefore, it is kely that increased program emphasis as well as expansion of the current RD&D program would sult in increased energy production over a latively short time period.

e general approach for achieving program goals d near-term energy increases requires the rallel development of several biomass produc- on and conversion technologies and the use of adily available resources to test or demon- rate those missions that are relatively near mmercialization. Unless the program effort is gnificantly expanded, less effort should be aced on options with low market penetration po- ntial and distant commercialization dates.

sed only on market penetration projections, ocesses that appear to offer minor future con- ibutions include marine crop production, talytic liquefaction, and fermentation to pro- ce ethanol. Missions that appear to have er near term commercialization potential in- :

- Gasification of wood and low moisture plants to produce IBG, SNG, LBG, and ammonia.
- Pyrolysis of wood and low moisture plants to produce SNG, fuel oil and char.
- Combustion of wood and low moisture plants to produce steam and steam with electricity as a by-product.
- Anaerobic digestion of manure and high moisture terrestrial crops to produce IBG and SNG.

Increased RD&D emphasis and federal commercialization incentives appear to be justified for all of the above missions. However, only limited additional RD&D support is recommended for the direct combustion mission because its commercialization potential is firmly established.

The remainder of this Executive Summary discusses specific and general recommendations based upon the foregoing analysis as well as prior biomass related research conducted at SRI and elsewhere.

Overall biomass program success (as measured by the levels of future quad production) can best be achieved by emphasis on production methods, procedures and policies designed to increase feedstock availability. Biomass fuel product market penetration is highly sensitive to feedstock scenario and costs. A doubling of feedstock availability at $30 per ton allows biomass energy product production to approximately double. Therefore, financial incentive systems, of various types, possibly including loan, tax credit, and subsidy systems, should be developed to: (a) increase the use of available crop and timber lands for the production of energy crops and combination energy, food/fiber crops, and (b) encourage construction and operation of biomass fuel and chemical product production facilities.

Additional program recommendations resulting from the study are contained in the final report volumes.

Definitions

Biomass product production price--the total of all costs necessary for the manufacture of a biomass-derived product, including manufacturer's profit.

Market price--the price against which the biomass-derived products must compete for market share.

Steady-state market share--the fraction of the product market that a biomass-derived product would supply at a point long after the biomass product production price and market price had become unchanging with time.

Market-share parameter (γ)--a variable used to characterize the behavior of an energy marketplace. This variable estimates the steady-state market share of a biomass product once the comparative economics are determined.

Behavioral lag--a means of describing the resistance of market decision-makers to the introduction of new energy products and technologies. This concept forms the basis for the dynamic market response curve.

Behavioral lag half-life (h)--the time required for one-half of a given market to respond to the availability of a new energy product or technology.

Behavioral lag response parameter (α)--this variable fixes the relative shape of the dynamic market response curve for a particular market once the behavioral lag half-life has been chosen.

Mission--a specific biomass feedstock to product conversion process or procedure directed toward a designated fuel or petrochemical market.

Appendix

SUMMARY OF COST SENSITIVITY DATA ON 15 MISSIONS

Mission

1 Wood to oil
2 Wood to methanol
3 Wood to ammonia
4 SNG from wood
5 Steam from wood
6 Electricity and steam from wood
7 IBtu gas from cattle manure
8 SNG from cattle manure
9 Wheat straw to IBtu gas
10 Wheat straw to ethanol
11 High sugar content plant to ethanol
12 Wood to oil via pyrolsis (maximum liquid yield)
13 Kelp to SNG via anaerobic digestion
14 Kelp to alcohol via fermentation
15 Cogeneration of electricity and steam from wood

MISSION I--SELECTED SUMMARY DATA

WOOD TO OIL VIA CATALYTIC LIQUEFACTION

	Base Case	Sensitivity To:							
		Feedstock Price			Operating Percent		Capital Investment		Project Life
A. Product									
Output/Day Bbls of oil	5268	5268	5268	5268	5268	5268	5268	5268	5268
Equivalent Btus/day (10^9)	30.6	30.6	30.6	30.6	30.6	30.6	30.6	30.6	30.6
B. Feedstock									
ODT per day*	3000	3000	3000	3000	3000	3000	3000	3000	3000
Btus/day (10^9)	57.6	57.6	57.6	57.6	57.6	57.6	57.6	57.6	57.6
Cost-Dollars/million Btus	$1.0	2.0	1.5	0	1.0	1.0	1.0	1.0	1.0
C. Total Capital Investment							-20%	+30%	
Millions of dollars***	144.9	146.5	145.7	143.4	144.7	144.5	116.5	189.6	144.9
Dollars/million Btus**	$1.95	1.98	1.97	1.93	2.22	2.52	1.57	2.52	1.72
D. Annual Cost of Feedstock									
Millions of dollars	18.8	37.7	28.3	0	16.8	14.7	18.8	18.8	18.8
Dollars/million Btus	$1.88	3.76	2.83	0	1.87	1.87	1.88	1.88	1.88
E. Annual Operating Cost†									
Millions of dollars	15.5	15.5	15.5	15.5	15.3	15.1	13.2	18.9	15.5
Dollars/million Btus	1.54	1.54	1.54	1.54	1.70	1.93	1.31	1.88	1.54
Operating percent	90%	90%	90%	90%	80%	70%	90%	90%	90%
F. Revenue Requirements* **									
Dollars/Bbl of oil	31.2	42.2	36.7	20.1	33.6	36.6	27.6	38.5	29.8
Regulated utility Dollars/million Btus††	5.37	7.28	6.33	3.47	5.79	6.32	4.76	6.29	5.14
G. Plant Life Years	20	20	20	20	20	20	20	20	30

---

* Assumes a 50% moisture content feedstock and 19.2 million Btus/dry ton - process efficiency = 53%

† Excludes feedstock cost and plant depreciation

** Capital component of product cost

*** Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% return on debt (65% debt and 35% equity). Income tax = 52%

†† Assumes 5.8 x 10^6 Btus/barrel

MISSION 2--SELECTED SUMMARY DATA

WOOD TO METHANOL VIA GASIFICATION (OXYGEN BLOWN REACTOR)

	Base Case:			Sensitivity To:							
				Feedstock Price			Operating Percent		Capital Investment		Project Life
A. Product											
Gallons of Methanol/day†† (10^3)	100	200	600	600	600	600	600	600	600	600	600
Btus/day (10^9)	5.5	10.06	33.2	33.2	33.2	33.2	33.2	33.2	33.2	33.2	33.2
B. Feedstock											
ODT per day*	500	1000	3000	3000	3000	3000	3000	3000	3000	3000	3000
Btus/day (10^9)	9.6	19.2	57.6	57.6	57.6	57.6	57.6	57.6	57.6	57.6	57.8
Cost-Dollars/million Btu	$1.0	1.0	1.0	2.0	1.5	0	1.0	1.0	1.0	1.0	1.0
C. Total Capital Investment									-20%	+30%	
Millions of dollars***	58.13	100.76	268.71	270.26	269.48	267.16	268.50	268.28	215.61	348.36	268.71
Dollars/million Btu **	4.33	3.51	3.35	3.37	3.36	3.32	3.73	4.31	2.69	4.34	2.95
D. Annual Cost of Feedstock†											
Millions of dollars	3.14	6.28	18.84	37.69	28.26	0	16.75	14.66	18.84	18.84	18.84
Dollars/million Btu	1.73	1.73	1.73	3.47	2.60	0	1.73	1.73	1.73	1.73	1.73
E. Annual Operating Cost											
Millions of dollars	6.88	8.98	29.38	29.38	29.38	29.38	28.88	28.38	25.10	35.80	29.39
Dollars/million Btu	3.81	2.72	2.69	2.69	2.69	2.69	2.98	3.33	2.30	3.28	2.69
Operating percent	90%	90%	90%	90%	90%	90%	80%	70%	90%	90%	90%
F. Revenue Requirements											
Dollars/million Btu Regulated utility	9.87	7.96	7.77	9.53	8.65	6.01	8.47	9.37	6.72	9.35	7.37
G. Plant Life Years	20	20	20	20	20	20	20	20	20	20	30

---

* Assumes a 50% moisture content feedstock and 19.2 MMBtus/dry ton - process efficiency = 58%

† Excludes feedstock cost and plant depreciation

** Capital component of product cost

*** Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% return on debt (65% debt and 35% equity) Income tax = 52%

†† Assumes 55,610. Btus/gallon

MISSION 3--SELECTED SUMMARY DATA

AMMONIA FROM WOOD VIA GASIFICATION WITH AN OXYGEN BLOWN REACTOR

	Base Case			Sensitivity To: Feedstock Price			Operating Percent		Capital Investment		Project Life
A. Product											
Tons of ammonia/day	250	500	1542	1542	1542	1542	1542	1542	1542	1542	1542
Btus/day (10^9)[*]	4.6	9.2	28.2	28.2	28.2	28.2	28.2	28.2	28.2	28.2	28.2
B. Feedstock											
ODT per day	500	1000	3000	3000	3000	3000	3000	3000	3000	3000	3000
Btus/day (10^9)	$9.6	$19.1	$57.3	$57.3	$57.3	$57.3	$57.3	$57.3	$57.3	$57.3	$57.3
Cost-dollars/million Btu	1.00	1.00	1.00	2.00	1.50	0	1.00	1.00	1.00	1.00	1.00
C. Total Capital Investment									- 20%	+ 30%	
Millions of dollars	$65.0	$110.1	$267.3	$268.9	$268.0	$265.7	$267.1	$266.9	$214.4	$346.5	$267.3
Dollars/million Btu[**]	$6.95	$5.96	$4.71	$4.74	$4.73	$4.68	$5.52	$6.57	$3.86	$6.00	$4.24
Dollars/ton	$127.40	$109.02	$86.24	$86.75	$86.49	$85.73	$101.17	$120.38	$70.53	$109.77	$77.64
D. Annual Cost of Feedstock											
Millions of dollars	$3.1	$6.3	$18.8	$37.7	$28.3	0	$16.8	$14.7	$18.8	$18.8	$18.8
Dollars/million Btu	$2.09	$2.09	$2.03	$4.06	$3.04	0	$2.03	$2.03	$2.03	$2.03	$2.03
Dollars/ton	$38.27	$38.27	$37.14	$74.28	$55.71	0	$37.14	$37.14	$37.14	$37.14	$37.14
E. Annual Operating Cost[†]											
Millions of dollars	$5.7	$9.6	$20.6	$20.6	$20.6	$20.6	$20.6	$20.6	$17.2	$25.7	$20.6
Dollars/million Btu	$3.78	$3.18	$2.22	$2.22	$2.22	$2.22	$2.22	$2.22	$1.85	$2.76	$2.22
Dollars/ton	$69.08	$58.18	$40.55	$40.55	$40.55	$40.55	$40.55	$40.55	$33.88	$50.57	$40.55
Operating percent	90	90	90	90	90	90	80	70	90	90	90
F. Revenue Requirement											
Regulated utility:[***]											
Dollars/million Btu	$12.82	$11.23	$8.96	$11.02	$9.99	$6.90	$9.77	$10.82	$7.74	$10.79	$8.49
Dollars/ton	$235.	$205.	$164.	$202.	$183.	$126.	$179.	$198.	$142.	$197.	$155.
DCF:											
Dollars/million Btu	$18.93	$16.39	$13.60	$15.67	$14.64	$11.52	$15.00	$16.79	$11.46	$16.81	$13.60
Dollars/ton	$346.	$300.	$249.	$287.	$268.	$211.	$274.	$307.	$210.	$308.	$249.
G. Plant Life--Years	20	20	20	20	20	20	20	20	20	20	30

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[*] Assumes 18.3 million Btus/ton - process efficiency = 49%.

[**] Capital component of product cost.

[***] Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% rate of return on debt (65% debt and 35% equity - income tax = 52%.

[†] Excludes feedstock cost and plant depreciation.

MISSION 4--SELECTED SUMMARY DATA

SNG PRODUCTION FROM WOOD VIA GASIFICATION (OXYGEN BLOWN REACTOR)

				Sensitivity To							
	Base Cases			30 Year Facility Life	Plant Investment Cost		Feedstock Cost			Operating % of Capacity	
A. Product											
Output per day (SCF) 10^6	6.4	12.7	38.2	38.2	38.2	38.2	38.2	38.2	38.2	38.2	38.2
Equivalent Btus (10^9)	6.0	12.0	36.1	36.1	36.1	36.1	36.1	36.1	36.1	36.1	36.1
B. Feedstock											
ODT per day[*]	500	1,000	3,000	3,000	3,000	3,000	3,000	3,000	3,000	3,000	3,000
Btus In per day[*] (10^9)	9.6	19.2	57.4	57.4	57.4	57.4	57.4	57.4	57.4	57.4	57.4
Cost- Dollars per Million Btus	$1.00	$1.00	$1.00	$1.00	$1.00	$1.00	$2.00	$1.50	$1.00	$1.00	$1.00
C. Total Capital Investment- Millions of dollars[***]	$50.1	$88.5	$238.5	$238.5	$191.4 - 20 %	$309.2 + 30 %	$240.1	$239.3	$237.0	$238.3	$238.1
Dollars per million Btu[**]	(3.40)	(3.01)	(2.70)	(2.70)	(2.16)	(3.49)	(2.71)	(2.71)	(2.68)	(3.03)	(3.46)
D. Annual Cost of Feedstock Millions of dollars	$ 3.14	$ 6.3	$18.8	$18.8	$18.8	$18.8	$37.7	$28.3	-0-	$16.8	$14.7
Dollars per million Btus	(1.59)	(1.59)	(1.59)	(1.59)	(1.59)	(1.59)	(3.18)	(2.39)	---	(1.59)	(1.59)
E. Annual Operating Cost[†] Millions of dollars	$ 5.9	$10.3	$25.2	$21.2	$21.4	$30.9	$25.2	$25.1	$25.1	$24.9	$24.5
Dollars per million Btus	(2.97)	(2.61)	(2.12)	(1.79)	(1.81)	(2.61)	(2.13)	(2.12)	(2.12)	(2.36)	(2.66)
Operating percent capacity	90%	90%	90%	90%	90%	90%	90%	90%	90%	80%	70%
F. Revenue Requirements Dollars per million Btus (regulated utility)	$7.96	$7.21	$6.41	$6.08	$5.56	$7.69	$8.02	$7.22	$4.80	$6.98	$7.71

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[*] Assumes a 50% moisture content feedstock and 19.2 million Btus per dry ton - process efficiency = 63%

[†] Excludes feedstock cost and plant depreciation.

[**] Capital component of product cost.

[***] Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and on 9% return on debt (65% debt - 35% equity). Income tax = 52% Facility life = 20 years.

MISSION 5--SELECTED SUMMARY DATA

STEAM PRODUCTION FROM WOOD VIA DIRECT COMBUSTION

				Sensitivity To					
	Base Cases			Plant Investment Cost		Feedstock Cost		Operating % of Capacity	
A. Product - Steam									
Output per day (000 lbs/Hr)	239	478	1,434	1,434	1,434	1,434	1,434	1,434	1,434
Equivalent Btus per Day[††]	7.4	14.8	44.4	44.4	44.4	44.4	44.4	44.4	44.4
B. Feedstock									
ODT per Day[*]	500	1,000	3,000	3,000	3,000	3,000	3,000	3,000	3,000
Btus In.per day (10^9)	9.6	19,2	57.4	57.4	57.4	57.4	57.4	57.4	57.4
Cost-dollars per million Btus	1.00	1.00	1.00	1.00	1.00	1.50	-0-	1.00	1.00
C. Total Capital Investment (million dollars)[***]	$17.4	$32.3	$94.1	$76.0 - 20%	$121.3 + 30%	$94.9	$92.6	$93.7	$93.3
Dollars per million Btus[**]	(0.98)	(0.90)	(0.89)	(0.71)	(1.13)	(0.89)	(0.86)	(1.12)	(1.56)
D. Annual cost of feedstock (millions of dollars)	$3.1	$6.3	$18.9	$18.9	$18.9	$28.3	-0-	$14.7	$10.5
Dollars per million Btus	(1.30	(1.30)	(1.30)	(1.30)	(1.30)	(1.95)	-0-	(1.30)	(1.30)
E. Annual operating costs (millions of dollars)[†]	$ 2.6	$ 4.5	$11.8	$10.2	$14.1	$11.8	$11.8	$11.2	$10.6
Dollars per million Btus	(1.08)	(0.95)	(0.81)	(0.72)	(0.98)	(0.82)	(0.82)	(1.00)	(1.33)
Operating Capacity	90%	90%	90%	90%	90%	90%	90%	70%	50%
F. Revenue Requirements dollars per million Btus (regulated utility)	$ 3.36	$ 3.15	$ 3.00	$ 2.73	$ 3.41	$ 3.66	$ 1.68	$ 3.42	$ 4.19

[†] Excludes feedstock cost and plant depreciation.

[*] Assumes a 50% moisture feedstock - 19.2 million Btu per dry ton - process efficiency = 77%

[**] Capital component of product cost.

[***] 15% return on equity (35% of total) and 9% return on debt (65% of total) - facility life = 20 years.

[††] 450 lbs psia - 810°F

MISSION 6--SELECTED SUMMARY DATA

ELECTRICITY PRODUCTION FROM WOOD VIA DIRECT COMBUSTION

				Sensitivity To						
	Base Cases			Plant Investment Cost		Feedstock Cost		Operating % of Capacity		
A. Product Plant size MW =	25	50	150	150	150	150	150	150	150	150
Output per day MWh	600	1200	3600	3600	3600	3600	3600	3600	3600	3600
Equivalent Btus per day	2.04	4.08	12.24	12.24	12.24	12.24	12.24	12.24	12.24	12.24
B. Feedstock										
ODT per day*	500	1000	3000	3000	3000	3000	3000	3000	3000	3000
Btus In. per day (10^9)	9.6	19.2	57.4	57.4	57.4	57.4	57.4	57.4	57.4	57.4
Cost-dollars per million Btus	1.00	1.00	1.00	1.00	1.00	1.50	-0-	1.00	1.00	1.00
				- 20%	30%					
C. Total Capital Investment (million dollars)***	$30.7	$58.2	$165.6	$133.3	$214.2	$166.3	$164.3	$165.4	$165.1	$164.8
Dollars per million Btu**	(7.00)	(6.61)	(6.25)	(5.03)	(8.09)	(6.29)	(6.19)	(7.15)	(9.08)	(12.44)
D. Annual Cost of Feedstock (million dollars)	$ 2.8	$ 5.6	$16.8	$16.8	$16.8	$25.1	-0-	$14.7	$11.5	$ 8.4
Dollars per million Btus	(4.70)	(4.70)	(4.70)	(4.70)	(4.70)	(7.05)	—	(4.70)	(4.70)	(4.70)
E. Annual Operating Cost (million dollars)†	$ 4.1	$ 7.6	$19.3	$16.6	$23.4	$19.3	$19.3	$18.8	$18.1	$17.4
Dollars per million Btu	(6.93)	(6.43)	(5.43)	(4.67)	(6.57)	(5.43)	(5.43)	(6.04)	(7.39)	(9.77)
Operating % Capacity	80%	80%	80%	80%	80%	80%	80%	70%	55%	40%
F. Revenue Requirements dollars per million Btu (regulated utility)	$18.63	$17.74	$16.38	$14.40	$19.36	$18.77	$11.62	$17.89	$21.17	$26.91

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†Excludes feedstock cost and plant depreciation

*Assumes a 50% moisture content feedstock - 19.2 MM Btu/dry ton - plant efficiency = 21%

**Capital component of product cost.

***15% return on equity (35% of total) and 9% return on debt (65% of total) - facility life = 20 years.

MISSION 7--SELECTED SUMMARY DATA

IBTU GAS PRODUCTION FROM CATTLE MANURE VIA ANAEROBIC DIGESTION

				Sensitivity To						
	Base Cases			Plant Investment Cost		Feedstock Cost			Operating % of Capacity	
A. **Product** (No. of Head of Cattle) =	10,000	100,000	250,000	10,000	10,000	100,000	100,000	100,000	10,000	10,000
Output per day (10^6 scf)	.226	2.26	5.65	.226	.226	2.26	2.26	2.26	.226	.226
Equivalent Btus per day (10^6)*	226	2260	5650	226	226	2260	2260	2260	226	226
B. **Feedstock**										
ODT per day	45	450	1125	45	45	450	450	450	45	45
Btus In per day (10^6)	675	6750	16,875	675	675	6750	6750	6750	675	675
Cost- Dollars per Million Btus	.33	.33	.33	.33	.33	.165	0.67	0	.33	.33
C. **Total Capital Investment-**				- 20 %	+ 30 %					
Millions of dollars***	1.83	10.82	23.91	1.48	2.34	10.82	10.82	10.82	1.83	1.83
Dollars per Million Btus**	(2.68)	(1.35)	(1.13)	(2.01)	(3.60)	(1.35)	(1.35)	(1.35)	(3.19)	(3.66)
D. **Annual Cost of Feedstock**										
Millions of dollars	0.07	.74	1.86	0.07	0.07	.37	1.48	0	0.07	0.06
Dollars per Million Btus	(1.00)	(1.00)	(1.00)	(1.00)	(1.00)	(0.50)	(1.99)	(0)	(1.06)	(1.04)
E. **Annual Operating Cost**†										
Millions of dollars	0.39	1.87	3.45	0.36	0.44	1.87	1.87	1.87	0.38	0.38
Dollars per Million Btus	(5.26)	(2.52)	(1.86)	(4.86)	(5.94)	(2.52)	(2.52)	(2.52)	(5.75)	(6.62)
Operating percent capacity	90%	90%	90%	90%	90%	90%	90%	90%	80%	70%
F. **Revenue Requirements**										
Dollars per Million Btus										
Regulated utility	$ 8.94	$ 4.87	$ 3.99	$ 7.87	$ 10.54	$ 4.37	$ 5.86	$ 3.87	$ 10.00	$ 11.32

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† Excludes feedstock costs and plant depreciation

* Plant efficiency = 33.5%

** Capital component of product costs

*** Calculated to yield a 15% discounted cash flow (dcf) rate of return on equity and 9% return on debt (65% debt - 35% equity) facility life = 20 years

MISSION 8--SELECTED SUMMARY DATA

SNG PRODUCTION FROM CATTLE MANURE VIA ANAEROBIC DIGESTION

				Sensitivity To						
	Base Cases			Plant Investment Cost		Feedstock Cost			Operating % of Capacity	
No of Head of Cattle	10,000	100,000	250,000	10,000	10,000	100,000	100,000	100,000	10,000	10,000
A. **Product**										
Output per day (million scf)	.204	2.04	5.10	.204	.204	2.04	2.04	2.04	.204	.204
Equivalent Btus per day (millions)*	204	2040	5100	204	204	2040	2040	2040	204	204
B. **Feedstock**										
ODT per day	45	450	1125	45	45	450	450	450	45	45
Btus In per day	675	6750	16,875	675	675	6750	6750	6750	675	675
Cost- Dollars per Million Btus	0.33	0.33	0.33	0.33	0.33	0.165	0.67	0	0.33	0.33
C. **Total Capital Investment-**				- 20 %	+ 30 %					
Millions of Dollars***	$ 2.4	$ 13.6	$ 29.3	$ 1.9	$ 3.1	$ 23.8	$ 23.8	$ 23.8	$ 2.4	$ 2.4
Dollars per Million Btus**	(4.17)	(2.11)	(2.01)	(3.28)	(5.43)	(2.26)	(2.18)	(2.05)	(4.82)	(5.51)
D. **Annual Cost of Feedstock** Millions of Dollars	.07	.70	1.75	.07	.07	.35	1.40	0	.06	.05
Dollars per Million Btus	(1.04)	(1.04)	(1.04)	(1.04)	(1.04)	(0.52)	(2.08)	-	(1.04)	(1.04)
E. **Annual Operating Cost**† Millions of Dollars	$ 0.6	$ 2.6	$ 4.9	$ 0.6	$ 0.7	$ 2.6	$ 2.6	$ 2.6	$ 0.6	$ 0.6
Dollars/Million Btus	(9.23)	(3.87)	(2.90)	(8.49)	(10.42)	(3.87)	(3.87)	(3.87)	(10.24)	(11.70)
Operating % Capacity	90 %	90 %	90 %	90%	90 %	90 %	90 %	90 %	80 %	70 %
F. **Revenue Requirements** Dollars per Million Btus Regulated Utility	$ 14.44	$ 7.02	$ 5.95	$ 12.81	$ 16.89	$ 6.65	$ 8.13	$ 5.92	$ 16.10	$ 18.25

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† Excludes feedstock cost and plant depreciation

* Plant efficiency - 30.2%

** Capital component of product cost

*** Calculated to yield an 9% return on debt and 15% dcf return on equity (65% debt - 35% equity) - Plant life 20 years

MISSION 9--SELECTED SUMMARY DATA

WHEAT STRAW TO IBTU GAS VIA ANAEROBIC DIGESTION

	60% Conversion											
		Sensitivity To:								Sensitivity To:		
	Base Case	Feedstock Prices			Capital Costs		Operating Percent		Base Case	Feedstock Prices		
A. Product												
Cubic ft. gas/day (500 Btu/cu ft)	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	30.0	30.0	30.0	30.0
Btus/day (10^9)*	2.5	2.5	2.5	2.5	2.5	2.5	2.5	2.5	15.0	15.0	15.0	15.0
B. Feedstock												
ODT per day	500	500	500	500	500	500	500	500	3000	3000	3000	3000
Btus /day (10^9)	7.5	7.5	7.5	7.5	7.5	7.5	7.5	7.5	45.0	45.0	45.0	45.0
Cost per ton	$25	15	35	0	25	25	25	25	$25	15	35	0
C. Total Capital Investment					+30%	-20%						
Millions of dollars***	12.6	12.6	12.6	12.6	15.7	10.6	12.6	12.6	65.7	65.6	65.7	65.5
Dollars/million Btu **	2.10	2.10	2.10	2.10	2.61	1.78	2.38	2.70	1.82	1.82	1.82	1.81
D. Annual Cost of Feedstock												
Millions of dollars	$4.1	2.5	5.8	0	4.1	4.1	3.6	3.2	$24.7	14.8	35.6	0
Dollars/million Btu	4.99	2.99	6.99	0	4.99	4.99	4.99	4.99	4.99	2.99	6.99	0
E. Annual Operating Cost†												
Millions of dollars	$5.1	5.1	5.1	5.1	5.4	5.0	4.7	4.3	$27.4	27.4	27.4	27.4
Dollars/million Btu	6.21	6.21	6.21	6.20	6.52	6.01	6.48	6.80	5.55	5.55	5.55	5.55
Operating percent	90%	90%	90%	90%	90%	90%	80%	70%	90%	90%	90%	90%
F. Revenue Requirements												
Dollars/million Btu Regulated utility	$13.30	11.30	15.30	8.30	14.12	12.76	13.83	14.49	$12.36	10.36	14.36	7.36

	40% Conversion											
		Sensitivity To:								Sensitivity To:		
	Base Case	Feedstock Prices			Capital Costs		Operating Percent		Base Case	Feedstock Prices		
A. Product												
Cubic ft. gas/day (500 Btu/cu ft)	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	16.8	16.8	16.8	16.8
Btus/day (10^9)*	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4	8.4	8.4	8.4	8.4
B. Feedstock												
ODT per day	500	500	500	500	500	500	500	500	3000	3000	3000	3000
Btus /day (10^9)	7.5	7.5	7.5	7.5	7.5	7.5	7.5	7.5	45.0	45.0	45.0	45.0
Cost per ton	$25	15	35	0	25	25	25	25	$25	15	35	0
C. Total Capital Investment					+30%	-20%						
Millions of dollars***	$12.6	12.6	12.6	12.6	16.7	10.6	12.6	12.6	65.66	65.63	65.74	65.55
Dollars/million Btu **	3.74	3.74	3.74	3.74	4.56	3.05	4.23	4.84	3.25	3.25	3.25	3.25
D. Annual Cost of Feedstock												
Millions of dollars	$4.1	2.5	5.8	0	4.1	4.1	3.7	3.2	$24.75	14.85	34.65	.01
Dollars/million Btu	8.92	5.35	12.50	0	8.92	8.92	8.92	8.92	8.92	5.35	12.50	0
E. Annual Operating Cost†												
Millions of dollars	$5.1	5.1	5.1	5.1	5.4	5.0	4.7	4.3	$27.4	27.4	27.4	27.4
Dollars/million Btu	11.10	11.10	11.10	11.10	11.74	10.73	11.50	12.13	9.90	9.90	9.90	9.90
Operating percent	90%	90%	90%	90%	90%	90%	80%	70%	90%	90%	90%	90%
F. Revenue Requirements												
Dollars/million Btu Regulated utility	$23.76	20.19	27.34	14.84	25.22	22.79	24.71	25.89	$22.07	18.50	25.65	13.15

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* Process efficiency = 34%

† Excludes feedstock costs and plant depreciation

** Capital component of product cost

*** Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% return on debt (65% debt and 35% equity). Income tax = 52%

MISSION 10--SELECTED SUMMARY DATA

WHEAT STRAW TO ETHANOL (4% SUGAR SOLUTION) VIA ENZYMATIC HYDROLYSIS AND FERMENTATION

25 MM Gallons Per Year of Ethanol

(Facility Daily Outputs: 500 Tons of Sugar and 75,768 Gallons of Ethanol)

	Base Case	Sensitivity To:							
		Feedstock Prices			Capital Costs		Operating Percent		
A. **Product**									
Gallons of ethanol/day (000)	75.8	75.8	75.8	75.8	75.8	75.8	75.8	75.8	75.8
Btu/day (10^9)*	5.5	5.5	5.5	5.5	5.5	5.5	5.5	5.5	5.5
B. **Feedstock**									
ODT per day-Wheat straw	3270	3270	3270	3270	3270	3270	3270	3270	3270
Btu/day (10^9)	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0
Cost-$/dry ton-Wheat straw	$15.0	9.0	0	3.0	15.0	15.0	15.0	15.0	15.0
C. **Total Capital Investment*** **					+30%	-20%			
Sugar plant ($ 10^6)	$94.9	94.6	94.2	95.5	121.8	76.9	94.3	94.6	94.7
Ethanol plant ($ 10^6)	32.9	32.6	32.2	33.6	43.2	26.2	33.2	33.0	33.0
Total Capital Investment	127.8	127.2	126.2	129.1	165.0	103.1	127.5	127.5	127.7
Dollars/million Btu**	$8.90	8.86	8.79	9.00	11.50	7.18	11.38	9.98	9.52
D. **Annual Cost of Feedstock**									
Millions of dollars	$16.2	9.7	0	32.4	16.2	16.2	12.5	14.3	15.2
Dollars/million Btu	9.00	5.40	0	18.00	9.00	9.00	6.94	7.95	8.44
E. **Annual Operating Costs**††									
Sugar plant ($ 10^6)	54.4	54.4	54.4	54.4	56.6	53.0	44.2	49.3	51.7
Ethanol plant ($ 10^6)	8.2	8.2	8.2	8.2	8.2	8.2	6.7	7.4	7.8
Total Annual Operating Costs ($ 10^6)	62.6	62.6	62.6	62.6	64.8	61.2	50.9	56.7	59.5
Operating percent	90%	90%	90%	90%	90%	90%	70%	80%	84%
Dollars/million Btu	$34.70	34.70	34.70	34.70	36.00	33.92	36.15	35.31	35.29
F. **Revenue Requirements*** **									
Sugar Costs-$/Lb	$0.25	0.23	0.20	0.30	0.27	0.24	0.27	0.26	0.26
Regulated utility Dollars/million Btu-Ethanol	52.60	48.96	43.49	61.70	56.50	50.10	54.47	53.24	53.25

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* Process efficiency = 11%

** Capital component of product cost - assumes 30-year plant operation

†† Excludes feedstock cost and plant depreciation

return on debt

MISSION II--SELECTED SUMMARY DATA

SUGAR CANE MILL

High Sugar Content Plant to Ethanol (10.7% Sugar Solution) Via Fermentation
(165 Day/Year Sugar Plant - 330 Day/Year Ethanol Plant)

	Base Case	Sensitivity To: Feedstock Prices			Sensitivity To: Capital Cost	
A. Product						
Daily Output						
Tons of sugar	500	500	500	500	500	500
Gallons of ethanol (000)	75.8	75.8	75.8	75.8	75.8	75.8
Btu/day (10^9)*	5.5	5.5	5.5	5.5	5.5	5.5
B. Feedstock						
ODT per day (cane)	2756	2756	2756	2756	2756	2756
Btu/day (10^9)	41.3	41.3	41.3	41.3	41.3	41.3
Cost/dry ton	$65.0	50.0	100.	0	65.0	65.0
C. Total Capital Investment					+30%	-20%
Sugar plant ($ 10^6)	49.2	49.2	49.2	49.1	63.8	39.5
Ethanol plant ($ 10^6)	21.0	20.7	21.7	19.8	21.2	20.8
Total Capital Investment	60.2	69.9	70.9	68.9	84.0	60.3
Dollars/million Btu	4.17	4.84	4.91	4.77	5.82	4.18
D. Annual Cost of Feedstock						
Millions of dollars	$29.56	22.74	45.47	0	29.56	29.56
Dollars/million Btu	16.28	12.53	25.05	0	16.29	16.29
E. Annual Operating Costs						
Sugar plant ($ 10^6)	6.8	6.8	6.8	6.8	8.4	5.8
Ethanol plant ($ 10^6)	5.0	5.0	5.0	5.0	5.0	5.0
	11.8	11.8	11.8	11.8	13.4	10.8
Dollars/million Btu	$6.50	6.50	6.50	6.50	7.38	5.95
F. Revenue Requirements						
Dollars per LB of sugar	.13	.11	.18	.04	.14	.12
Regulated utility Dollars/million Btu	$. 2695	23.87	36.46	11.27	29.49	26.42

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* Sugar plant operates 165 day/year at 1000 T/day annual average = 500T/D
Operating percent sugar plant = 45%; Ethanol plant = 90%

MISSION 12--SELECTED SUMMARY DATA

WOOD TO OIL FOR DIRECT COMBUSTION AND CHAR VIA PYROLYSIS (MAXIMUM LIQUID YIELD)

	Base Case			Plant Size and Feedstock Price				Sensitivity To: Operating Percent	Sensitivity To: Capital Investment		Sensitivity To: Project Life
A. Product											
Gallons of oil/day†† (10^3)	26	52	156	26	52	156	156	156	156	156	156
Char (tons/day) *†	7.7	15.4	46.2	7.7	15.4	46.2	46.2	46.2	46.2	46.2	46.2
Btu/day (10^9)	151	302	918	151	302	918	918	918	918	918	918
B. Feedstock											
GDT per day*	500	1000	3000	500	1000	3000	3000	3000	3000	3000	3000
Btu/day (10^9)	9.5	19.1	57.3	9.5	19.1	57.3	57.3	57.3	57.3	57.3	57.3
Cost-Dollars/MM Btu	1.0	1.0	1.00	2.0	2.0	2.0	0	1.0	1.0	1.0	1.0
C. Total Capital Investment									-20%	+30%	
Millions of dollars***	12.3	22.2	61.4	12.6	22.7	63.0	59.9	61.2	49.7	79.0	61.4
Dollars/MM Btu**	0.7	0.7	0.6	0.7	0.7	0.7	0.6	0.7	0.5	0.7	0.5
D. Annual Cost of Feedstock†											
Millions of dollars	3.1	6.3	18.8	6.3	12.6	37.7	0	16.8	18.8	18.8	18.8
Dollars/MM Btu	1.4	1.4	1.4	2.7	2.7	2.7	0	1.4	1.4	1.4	1.4
E. Annual Operating Cost											
Millions of dollars	3.0	5.3	10.3	3.0	5.3	10.3	10.3	10.3	9.0	12.4	10.3
Dollars/MM Btu	1.3	1.1	0.7	1.3	1.1	0.7	0.7	0.8	0.6	0.9	0.7
Operating percent	90%	90%	90%	90%	90%	90%	90%	80%	90%	90%	90%
F. Revenue Requirements (Total Product Basis)											
Dollars/MM Btu	3.4	3.2	2.7	4.7	4.5	4.1	1.3	2.9	2.5	3.0	2.6
Dollars/MM Btu (Oil only)†††	6.1	5.6	4.5	8.9	8.5	2.6	1.4	4.9	4.0	5.1	4.2
G. Project Life (Years)	20	20	20	20	20	20	20	20	20	20	30

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* Assumes a 50% moisture content feedstock and 19 MMBtu/dry ton - process efficiency = 73.6%

† Excludes feedstock cost and plant depreciation

*† Assumes .302 lbs/1 lb dry wood

** Capital component of product cost

*** Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% return on debt (65% debt & 35% equity)

†† Assumes .250 lbs/1 lb dry wood or 210 gals of oil/dry ton Income Tax = 52%

††† Assumes char valued at 1.25/MM Btu and representing 55% of total output

MISSION 13--SELECTED SUMMARY DATA

KELP TO SNG VIA ANAEROBIC DIGESTION

		Plant Size			Sensitivity to Feedstock Prices					
A.	**Product**									
	Cubic ft. gas/day (10^6) (1000 Btu/cu ft.)	5.6	16.8	33.6	33.6	33.6	33.6	16.8	16.8	16.8
	Btu/day (10^9)	5.6	16.8	33.6	33.6	33.6	33.6	16.8	16.8	16.8
B.	**Feedstock**									
	DAFT/day ††	1000	3000	6000	6000	6000	6000	3000	3000	3000
	Btu/day (10^9)	16	16	16	16	16	16	16	16	16
	Cost/ton (DAF)	$100	100	100	$ 25	200	0	$ 25	200	0
C.	**Total Capital Investment**									
	Millions of dollars***	$30.1	68.9	115.7	$113.6	118.4	113.0	$67.9	70.3	67.6
	Dollars/MM Btu**	$ 2.2	1.6	1.3	$ 1.3	1.4	1.3	$ 1.6	1.7	1.6
D.	**Anuual Cost of Feedstock**									
	Millions of dollars	$33.0	99.0	198.0	$ 49.5	396.0	0	$24.7	198.0	0
	Dollars/MM Btu	$17.9	17.9	17.9	$ 4.5	35.7	0	$ 4.5	35.7	0
E.	**Annual Operating Cost†**									
	Millions of dollars	$ 4.3	9.3	16.0	$ 16.0	16.0	16.0	$ 9.3	9.3	9.3
	Dollars/MM Btu	$ 2.3	1.7	1.5	$ 1.5	1.5	1.5	$ 1.7	1.7	1.7
	Operating Percent	90%	90%	90%	90%	90%	90%	90%	90%	90%
F.	**Revenue Requirements**									
	Dollars/MM Btu Regulated Utility	$22.4	21.2	20.7	$ 7.3	38.6	2.8	$ 7.8	39.1	3.3

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* Process efficiency = 35%

† Excludes feedstock costs and plant depreciation; assumes 16 MM Btu/Dry Ash Free Ton of Feedstock

** Capital component of product cost

*** Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% return on debt (65% debt and 35% equity). Income tax = 52%

†† Dry Ash Free Tons

MISSION 14--SELECTED SUMMARY DATA

ALGAE TO ETHANOL VIA ACID HYDROLYSIS AND FERMENTATION
25 Million Gallons per Year Ethanol Plant

	50% Sugar Conversion							80% Sugar Conversion
	Base Case	Sensitivity To: Feedstock Prices		Capital Investment		Operating Percent		Base Case
A. **Product**								
Daily Output								
Tons of sugar	500	500	500	500	500	500	500	500
Gallons of ethanol (000)	76	76	76	76	76	76	76	76
Btu/day (10^9) *	5.7	5.7	5.7	5.7	5.7	5.7	5.7	5.7
B. **Feedstock**								
DAF tons per day	1126	1126	1126	1126	1126	1126	1126	703
Btu/day (10^9)	18.0	18.0	18.0	18.0	18.0	18.0	18.0	11.25
Cost/dry ton✓	75	100	0	75	75	75	75	75
C. **Total Capital Investment**				+30%	-20%			
Sugar plant ($ 10^6)	27.9	28.0	27.7	36.3	22.3	27.9	27.8	19.5
Ethanol plant ($ 10^6)	28.0	28.0	28.0	36.4	22.4	28.0	28.0	26.5
Total Capital Investment***	55.9	56.0	55.7	72.7	44.7	55.9	55.8	46.0
Dollars/MM Btu**	4.5	4.5	4.5	5.9	3.6	5.1	5.8	3.7
D. **Annual Cost of Feedstock** ✓✓								
Millions of dollars	27.9	37.2	0	27.9	27.9	24.7	21.6	17.4
Dollars/MM Btu	14.9	19.8	0	14.9	14.9	14.9	14.9	9.2
E. **Annual Operating Costs**								
Sugar plant ($ 10^6)	5.2	5.2	5.2	5.9	4.7	5.0	4.7	3.4
Ethanol plant ($ 10^6)	7.0	7.0	7.0	7.9	6.4	6.4	6.3	7.0
Total Annual Operating Costs (10^6)	12.2	12.2	12.2	13.8	11.1	11.4	11.0	10.4
Dollars/MM Btu	6.5	6.5	6.5	7.4	5.9	6.8	7.5	5.5
Operating %	90%	90%	90%	90%	90%	80%	70%	90%
F. **Revenue Requirements**								
Dollars per lb of sugar	0.11	0.14	0.03	0.12	0.11	0.12	0.11	0.08
Dollars/MM Btu of Ethanol	26.9	30.8	11.0	28.2	29.6	26.8	28.2	19.0

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* Process efficiency = 32%; assumes 75,600 Btu/gallon of ethanol

** Capital component of product cost

✓✓ Excludes feedstock cost and plant depreciation

*** Calculated to yield a 15% discounted cash flow (DCF) rate of return on equity and a 9% return on debt (65% debt and 35% equity). Income tax = 52%

✓ Assumes 16 MM Btu/DAF ton of feedstock, 80% carbohydrate

MISSION 15--SELECTED SUMMARY DATA

CO-GENERATION OF STEAM AND ELECTRICITY FROM WOOD VIA DIRECT COMBUSTION

	Base Cases			Sensitivity To						
				Plant Investment Cost		Feedstock Cost			Operating % of Capacity	
A. Product										
Plant capacity MW	3.65	7.3	21.9	21.9	21.9	21.9	21.9	21.9	21.9	21.9
Electricity MWh per day	87.6	175.	525	525	525	525	525	525	525	525
Steam pounds per hour (000)	210.	420.	1260.	1260.	1260	1260	1260	1260	1260	1260
Equivalent Btus per day (10^9)	7.2	14.5	43.5	43.5	43.5	43.5	43.5	43.5	43.5	43.5
B. Feedstock										
ODT per day*	500	1000	3000	3000	3000	3000	3000	3000	3000	3000
Btus in per day (10^9)	9.6	192	57.4	57.4	57.4	57.4	57.4	57.4	57.4	57.4
Cost- Dollars per Million Btus	1.00	1.00	1.00	1.00	1.00	1.50	0	1.00	1.00	1.00
C. Plant Life (years)	20	20	20	20	20	20	20	20	20	20
D. Total Capital Investment-				-20 %	+ 30 %					
Millions of dollars***	$ 21.6	$ 40.3	$ 109.1	$ 87.9	$ 140.8	$ 109.8	$ 107.7	$ 108.9	$ 108.6	$108.3
Dollars per Million Btus**	(1.37)	(1.29)	(1.16)	(0.94)	(1.49)	(1.17)	(1.15)	(1.33)	(1.68)	(2.30)
E. Annual Cost of Feedstock										
Millions of dollars	2.8	5.6	16.8	16.8	16.8	25.1	0	14.7	11.5	8.4
Dollars per Million Btus	(1.32)	(1.32)	(1.32)	(1.32)	(1.32)	(1.98)	-	(1.32)	(1.32)	(1.32)
F. Annual Operating Cost† ($ Millions)	2.9	5.1	12.0	10.2	14.6	12.0	12.0	11.8	11.5	11.3
Dollars/Million Btus	(1.36)	(1.19)	(0.94)	(0.80)	(1.16)	(0.94)	(0.94)	(1.06)	(1.32)	(1.78)
Operating percent	80%	80%	80%	80%	80%	80%	80%	70%	55%	40%
G. Revenue Requirements										
Dollars per Million Btus Regulated utility	$ 4.05	$ 3.80	$ 3.42	$ 3.06	$ 3.97	$ 4.09	$ 2.09	$ 3.71	$ 4.32	$ 5.40

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† Excludes feedstock cost and plant depreciation

* Assumes a 50% moisture content feedstock - 19.2 Million Btus per dry ton - Process efficiency = 75.7%

** The capital portion of product cost (cents per Million Btu of total product steam and electricity)

*** 15% return on equity (35% of total) and 9% return on debt (15% of total) - Plant life = 20 years

NOTES

Session IV A

SOLAR HYDROGEN PRODUCTION

William Hoagland
Solar Energy Research Institute
1536 Cole Blvd., Golden, Colorado 80401

ABSTRACT

The Solar Hydrogen Program at the Solar Energy Research Institute is currently investigating several methods of producing hydrogen from renewable resources. The major emphasis is the photobiological approach, specifically biophotolysis, defined as the formation of hydrogen gas from water using the photosynthetic apparatus of green plants and algae. Another technique being investigated is photoelectrolysis of water using sunlight. Currently, efforts are being directed toward the development of electrode materials with suitable absorption characteristics and hence, the ultimate conversion efficiency possible in the electrolytic cells, and also attaining the maximum photovoltage to reduce the external bias requirements.

The Solar Hydrogen Program is addressing very basic research to support as many new and novel approaches as possible until developmental efforts are warranted.

INTRODUCTION

The production of Hydrogen from non-fossil energy sources for use as a fuel, chemical reagent, or energy storage and transport medium has been gaining increased attention over the past several years. Although hydrogen was used extensively as a fuel for heating and lighting in the days of gas produced from coal, the principal application today is as a chemical intermediate in petroleum processing and for the synthesis of ammonia and methanol. Other industrial processes use hydrogen as a unique chemical for Hydrogenations, reducing atmospheres, etc. [1] It is currently estimated that as much as 1.4 Quads (10^{15} Btu) per year are consumed to produce hydrogen as a chemical feedstock, primarily from natural gas (76%) and petroleum (23%) [2] Worldwide demand for hydrogen (≈3 quads) has more than tripled in the last forty years and increased by a factor of 40 in the United States during the same period. [3]

Why is hydrogen being seriously considered as an energy transport and storage medium? Some of the advantages which hydrogen affords are: [4]

- o Hydrogen is versatile. It can be burned directly, catalytically oxidized at lower temperature, or used in a fuel cell to generate electricity.
- o Hydrogen is clean. The product of combustion is water only.
- Hydrogen is renewable. The primary source of hydrogen is also water or biomass.
- o Hydrogen is transportable. It may be transported via pipeline or truck as a primary fuel or blended with natural gas.
- o Hydrogen is storable. It may be stored as a liquid or a solid metal hydride.

The Department of Energy Hydrogen Energy Coordinating Committee (HECC) has given much attention to specific applications for hydrogen in a hydrogen economy. The major applications include: off-peak power storage, vehicular fuel, energy transport medium for nuclear energy, and energy storage medium for intermittent sources of energy (e.g., solar).

Projections of future hydrogen demand have been published in several recent reports, and show that as many as 22 quads of hydrogen could be required by the year 2000 to meet the needs of an accelerated synthetic fuels industry and new uses in the industrial utility and transportation areas. [5] It is rapidly becoming obvious that the nation must eventually change its production of hydrogen from natural gas and oil to a non-fossil energy source. Much basic research is needed before Solar Energy can become a viable, renewable energy source for hydrogen production. This paper will review some of the solar approaches to hydrogen production being investigated by the SERI program and their current status.

The overall objective of the Solar Hydrogen Production program at SERI is to develop economical, commercial processes to produce hydrogen from an abundant and renewable resource, specifically, solar energy. The evolution of hydrogen as a universal fuel has not come to pass, because there exists no viable process for large scale production which does not ultimately depend on a fossil fuel. Such processes must be cost competitive with other sources of energy for easy acceptance, because environmental or social forces are not likely to produce equal results before the end of the century.

Since the technical barriers to solar hydrogen production will require much basic research, the program goals are relatively long term and the budget is commensurately modest. The Solar Energy Research Institute is currently developing a recommended program plan for Solar Hydrogen Production. This plan will assess the current program and identify which technical problems should be addressed and their relative priority for a balanced program. Table 1 shows the current level of funding for the last two years and the number of subcontracted research efforts supported.

TABLE 1

Subcontractor	Approach	Description
Univ. of California (XH-8-1314)	Photo-biological	Research to develop practical processes for H_2 production using photosynthetic microorganisms as catalysts.
Battelle-Columbus (XH-9-1357)	Photoelectrolysis	Continued evaluation of commercial feasibility of H_2 production by the solar photoelectrolysis of water.
M.I.T. (XJ-9-1358)	Photoelectrolysis	Investigation of composite structures, solid solutions, and modified surfaces to develop efficient and stable electrodes.
Rockwell International (XH-9-8005)	Bio-conversion	Investigation of bromination process for production of H_2 from biomass.

Total funding: FY78 - $425K
FY79 - $450

SOLAR APPROACHES TO HYDROGEN PRODUCTION

Solar approaches for producing hydrogen may be put into these broad categories: photoconversion, bioconversion and thermochemical conversion processes. Photoconversion includes three processes which produce hydrogen at the direct expense of light energy:

- photobiological - the utilization of phototrophic organisms (and their components) to produce hydrogen at the expense of light energy.
- photoelectrochemical - the utilization of photoelectric properties of certain semiconducting materials to convert light energy into either electrical energy or the chemical potential of high energy products.
- photochemical - that part of chemistry which deals with the effects of light in producing chemical change (e.g., hydrogen).

Bioconversion (as a means of producing hydrogen) includes any scheme that uses a form of biomass as a source of hydrogen. Thermochemical processes which produce hydrogen mainly dissociate water into its elements at high temperature. This review will deal with photobiological, photoelectrochemical and bioconversion processes which produce hydrogen or other gaseous fuels.

PHOTOBIOLOGICAL PRODUCTION OF HYDROGEN

Since the phenomenon of hydrogen evolution or uptake in a biological system was first documented, by Roelofsen in 1934, many additional photobiological systems have been identified. [6] None, however, have developed to the point of engineering systems development due to the poor conversion efficiencies attained thus far. Three categories of biological systems identified as hydrogen producers are: 1) photosynthetic bacteria, 2) green algae and 3) cyanobacteria (blue-green algae). Progress in these areas in recent years give hope that such systems may have practical application for hydrogen production, although current system conversion efficiencies rarely exceed a few tenths of a percent, falling far short of the theoretical maximum calculated at around 5%. [7]

Short duration efficiencies of 2-3% have been achieved, however. [8] The SERI subcontracted Hydrogen Production Program is currently supporting only one effort in photobiology (See Table 1), but has solicited basic research proposals for new and novel approaches for photobiological hydrogen.

PHOTOCHEMICAL PRODUCTION OF HYDROGEN

There has been a great amount of increased interest in the field of photoelectrochemistry, due to the current energy crisis. The first ph[illegible]lectrochemical experiment was performed by Becquerel in 1839, [9] but current interest focuses on two aspects of photoelectrochemistry: Photovoltaic cells which produce electricity from sunlight, and photoelectrolysis. Photoelectrolysis of water using sunlight is of particular interest to the SERI Solar Hydrogen Production program, and we are currently funding two research efforts in this area (Figure 1). Electrode stability is a major problem for photoelectrolysis devices. For photoelectrolysis cells, n-type semiconductors that are stable against oxidative decomposition during O_2 evolution are limited thus far to wide band gap materials which results in conversion efficiencies 1%. Smaller band gap p-type semiconductors are also stable but require large external bias voltages. [10]

The SERI internal program also includes photoelectrochemical research, and three areas of interest have been proposed for next year's program. The first is to develop theory and models or a better basic understanding of the photo-induced charge transfer process at semiconductor-electrolyte interface and establish the thermodynamic limits on the conversion efficiency of these devices. The second is to develop new electrode materials and structures to find systems which exhibit high conversion efficiency and stability. The third is to identify and study other important chemical reactions beside water splitting which can be driven by photoelectrochemical cells. [11] It is planned that the external Solar Hydrogen Production program and the in-house activities will complement each other.

CONVERSION OF BIOMASS TO PRODUCE HYDROGEN

Although there are many approaches for utilizing biomass as a source of hydrogen, only one is being supported by the SERI program. This particular effort in progress at Rockwell International is looking at the feasibility and stoichiometry of reacting bromine with waste cellulose to produce hydrobromic acid which is then electrolyzed to produce hydrogen. The bromine is recycled. The current work is concerned with studying the bromination reaction with a variety of biomass and determining reaction rates. The results to date indicate the bromination reaction is adaptable to a variety of biomass feedstocks, and residence times of 10 minutes or less is possible. [12] Many other processes are being investigated, but at present are not a part of the SERI program.

FUTURE DIRECTIONS

The SERI Solar Hydrogen Production program is developing a research strategy which will optimize the program balance and allocation of resources. At present, the program is supported solely by [illegible]mass Energy Systems Branch at the Depart[illegible] Energy, so the program must necessarily emphasize the biological/biomass aspects of hydrogen production. Toward this end, a Request for Proposal was issued during FY79 for "Studies to Investigate Photobiological Production of Hydrogen or Other Gaseous Fuels," and several new efforts are being initiated as a result. Additional support for the photoelectrolysis and other approaches will be sought when the recommended program plan is available (around 9/1/79). It is also anticipated that the activities will emphasize basic research until the technical and economic barriers are resolved to the point where pilot scale systems studies would be justified.

SUMMARY

The availability of hydrogen produced from renewable energy source will provide a tremendous opportunity for displacement of our natural gas usage by a clean fuel. This could have many economic and environmental advantages in addition to finding a replacement for a resource which is being rapidly depleted. Hydrogen has many special advantages over other alternatives, and is a virtually unlimited resource.

While many of the cited approaches hold much promise for mid- to long-term applications, much basic research needs to be accomplished before any can be selected for development. Thus, the immediate objective of the SERI program is to provide the framework within which new and novel approaches for solar hydrogen production can be proposed, researched, assessed and developed as appropriate. This will be accomplished through close interaction with other organizations (both in and outside of government) and through a well balanced, appropriately funded program.

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FOLLOW-ON STUDIES OF THE PHOTOSYNTHESIS ENERGY FACTORY

CONTRACT NO. EY-76-C-01-2548: 7 JULY 1978

Malcolm D. Fraser, John F. Henry, Norman J. Barbera, and Louis C. Borghi
InterTechnology/Solar Corporation
Warrenton, Virginia 22186

ABSTRACT

A Photosynthesis Energy Factory (PEF) is an integrated bioconversion system consisting of a dry-land Energy Plantation^{R}, a wood-fired power plant, and a wetlands biological wastewater treatment system, such as an algae pond. Products of a PEF are electricity from the power plant, synthetic natural gas from digestion of the wetlands biomass, and reclaimed wastewater. Effluents and by-products from one system part can be beneficially used by other parts, leading to increased energy conversion and resource recovery at lower costs. In the initial study a general technoeconomic model was used to investigate possible interactions between the various subsystems. In a second project the PEF model has been expanded and generalized by analyzing possible model improvements in the areas of materials transportation, water and nutrient balances, other types of wetlands biological systems, and improvements in wood-fired combustion systems. Direct application of municipal wastewater to a dry-land Energy Plantation has been analyzed, also. Potential sites for PEF systems will be evaluated with the improved technoeconomic model.

INTRODUCTION

The Energy Plantation^{R*} concept developed at InterTechnology/Solar Corporation [1**,2,3] is an innovative, systematic approach to producing a practical and economic fuel from plant matter on a large scale. The plantation is designed and operated to minimize the cost of the plant material produced by carefully choosing the appropriate plant species, planting density, and harvest schedule for each site. A discussion of the design, operation, and economics of the Energy Plantation has already appeared in the literature [4]. In-depth engineering studies by InterTechnology/Solar [5,6] have established that, depending upon the plantation location, the cost of such perpetually renewable clean fuel in its solid as-fired form can be competitive with conventional fossil fuels.

Another alternate source of energy, which also offers the additional advantage of decreasing the environmental impact associated with the disposal of wastewater and residues, is the concept of growing algae in shallow ponds. Algae ponds are open shallow ponds in which algae and bacterial populations work together to utilize sunlight and nutrients to produce cell mass. In an algae pond system for recovering energy from various residues, the algae would be digested anaerobically to yield a methane-containing gas, which would be processed into SNG.

As long as 20 years ago, work was undertaken at the University of California at Berkeley to develop a system utilizing the algae pond concept to treat waste and to produce fuels. The technical problems were essentially concerned with integrating the components and optimizing their operation. The components themselves--the algae pond, the digester, and the sedimentation, separation, and finishing stages--were already being used in waste treatment. The work over the past 20 years has concentrated in three areas: (1) identifying and quantifying algal growth-limiting parameters; (2) maximizing gas production from anaerobic fermentation; and (3) optimizing the system with respect to gas production, residue uptake, land utilization, and cost-effectiveness.

Both the Energy Plantation and the algae pond can contribute to the solution of the population, resources, and energy problems facing us. Recently, it became apparent that they could better accomplish these missions when the two are integrated to form one composite system. In short, each produces a by-product that can be used to advantage by the other. The carbon content of sewage limits the production of the algae pond, but carbon dioxide, a by-product of combustion of solid Energy Plantation fuel (which currently appears to be the best way of using plant matter as fuel), can be supplied to increase productivity of the algae pond. The waste heat from the boiler can be used to control the temperature of the algae digester. The sludge generated as a by-product of the algae pond provides a source of inorganic nutrients and water for the Energy Plantation, which thus provides an ideal disposal site for the sludge.

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[*] Energy Plantation is a registered trademark of InterTechnology/Solar Corporation.

[*] bers in parentheses designate References at e paper.

The Photosynthesis Energy Factory (PEF) is thus a synergistic combination of the dry-land Energy Plantation and the algae pond which can produce on a perpetually renewable basis nonpolluting and totally domestic fuels from marginally useful land, solar energy, and various residues. Simultaneously, from different parts of the PEF, chipped solid fuel is produced, from which electricity is generated, and methane or SNG is recovered from waste CO_2 and municipal or industrial wastewater. Incidental economic benefits--which are significant--include secondary or tertiary treatment of municipal and industrial effluents, and the complete elimination of the need for a sanitary landfill for disposal of the resultant sludge. The PEF is not merely a combination of convenience, but a truly interactive utilization of materials and energy.

PREVIOUS WORK

An initial project was undertaken to study the concept of the PEF and its characteristics and apparent benefits, and to develop preliminary plans for a demonstration system. The project was divided into three parts or tasks. The objective of one task was to analyze the concept of the PEF, with particular emphasis on the complementary and synergistic aspects of the system. A second task was concerned with the analysis and selection of potential sites. The final task was to develop preliminary designs and associated cost estimates for potential demonstration systems at the best sites.

The PEF comprises three major subsystems--the dry-land Energy Plantation, a wood-fired power plant, and an algae production system. Various streams of energy and materials flow between these subsystems. To analyze the resultant interactions between the subsystems, a comprehensive technoeconomic model was developed to describe the performance and the cost of the PEF. Models of the three subsystems were developed with the aid of information and data that were already available as the result of previous studies. New data and new concepts were introduced into the models wherever possible. The University of California at Berkeley, as a subcontractor on this initial project, was responsible for supplying state-of-the-art data on algae pond performance and costs. These subsystem models were engineering models developed in sufficient detail to represent the important variables and variable-parameter interactions influencing subsystem performance and costs. Of the three subsystem models, the most comprehensive and the most complex was the Energy Plantation model, which is a complete design model.

For the twin purposes of defining the applicability of the PEF concept and selecting the best site for a demonstration PEF project, data were gathered on the characteristics of land, the availability of municipal and industrial effluents and residues, and the supply and demand for energy at a wide variety and number of potential sites. A format was developed for handling this data base, and a number of suitability indexes were defined for evaluating the site data. Dat were obtained from a number of sources in the literature as well as from state energy offices on a large number of potential sites. As the result of this site selection procedure, a number of sites were chosen for analysis by means of the technoeconomic model.

The model was then used to design a demonstration PEF system at each of the selected potential sites. This preliminary design illustrated for a specific site the benefits and the impact to be expected from a demonstration PEF project. Estimated costs were provided also for each demonstration PEF. Comparing these preliminary designs and their costs was then done to show where and under what conditions a PEF would be expected to be cost-effective in recycling wastes and producing fuels from biomass which would be competitive with presently used fuels. The results of this initial project have been published [7,8].

Results

The results of the site analyses illustrated the magnitude of the interactions within a PEF at a site. Some of these results are shown in Table I. In the analysis of the algae pond subsystem, it was found that at the present time, it is not cost-effective to design the algae pond to produce algae up to the phosphorus limit, with carbonation. The algae pond in the PEF at each site was sized to handle the wastewater flow from each population center, and the Energy Plantation was sized to produce solid fuel at an optimally low cost. This ratio of algae pond area to Energy Plantation area, or rather wastewater flow to production of biomass, is an important quantity which determines the extent of the influence of the algae pond upon the Energy Plantation.

The amount of SNG which can be produced by the algae pond system at each potential site is relatively small compared to the total market for gas at each site. The economic optimum for the size of an Energy Plantation unit, in terms of the cheapest fuel produced, is from 24,000 to 36,000 acres. The estimated productivities of these demonstration Energy Plantations varied from 4.1 to over 9 dry tons per acre-year.

The economics of a demonstration PEF at the potential sites were analyzed under the assumption of municipal financing. The annualized cash flow from the algae pond is a positive cash flow resulting from the applicable wastewater treatment credit rather than a cost--the wastewater treatment credit being by far the most significant part of the credit as compared to the credit for the gas produced. The excess credit or positive cash flow from the algae pond can be applied to the annualized cost of the power plant to enable the cost of electricity to be reduced. This reduction in cost amounts to 1.6 to 2.5 mills/kWh for the electricity produced from 100,000 tons of biomass per MGD of municipal wastewater flow. For the more efficient power plant, the cost of electricity, with the credit applied, of 40 m per kilowatt-hour or so for a demonstration P should be attractive to rural areas, which are

TABLE I

CHARACTERISTICS OF POTENTIAL DEMONSTRATION PEF'S
Carbon-Limited Design of Algae Pond System, Municipal Financing

	Algae Pond			Energy Plantation				Power Plant	
Site	Wastewater Flow MGD	Acres	Gas Produced 10^6 SCF / Yr	Acres	Tons* / Acre-Yr	10^3 Tons** / Yr	Cost of Biomass $ / O.D. Ton	Annual Prod. 10^6 KWH / Yr	Cost of Elect. Mills / KWH
Kissimmee, FL	1.3	20	4.15	36,000	7.06	262	17.63	456	52
Pensacola, FL	6.0	101	19.1	24,000	8.00	198	20.04	341	36
Chanute, KS	1.1	22	2.65	36,000	7.45	277	18.43	419	42
Maysville, KY	0.7	13	1.67	36,000	4.12	153	23.59	264	41
Natchitoches, LA	1.8	40	5.75	36,000	7.55	279	18.58	469	42
Minden, LA	1.5	34	4.8	36,000	7.04	261	19.17	446	42
Traverse City, MI	1.8	38	2.87	36,000	6.32	232	22.96	360	40
Bemidji, MN	1.1	21	2.04	27,000	9.05	250	16.90	381	40
Yazoo City, MS	1.1	27	3.51	36,000	5.62	209	19.01	360	43
Hammond, NY	1.5	54	3.18	36,000	6.37	237	18.95	411	39
Jamestown, NY	4.0	144	8.5	33,000	6.59	225	19.45	389	34
Greenwood, SC	2.1	46	6.71	30,000	6.50	201	18.80	362	40
Knoxville, TN	17.0	487	54.2	33,000	4.41	150	23.53	267	33
Caldwell, TX	0.4	6	1.15	36,000	7.38	275	19.99	415	42
Martinsville, VA	2.5	62	7.98	30,000	7.06	218	18.65	394	37

* Average productivity based on above-ground biomass, oven-dry tons.
** Average annual sustained production including contribution of root mass produced upon replanting, oven-dry tons.

now paying the highest rates for electricity transmitted over a large distance from a large central power plant operated by a large utility.

One important interaction between the algae pond and the Energy Plantation is the contribution of fertilizer from the algae pond. With ordinary municipal wastewater, a carbon-limited algae pond system operating the year around can provide about 2.9 percent (32.8 tons nitrogen per year) of the total annual nitrogen fertilizer requirement for 100,000 oven-dry tons of annual biomass production per MGD of wastewater flow, and this fertilizer contribution lowers the cost per ton of biomass about $0.15. Phosphorus and potassium are also produced from the algae pond.

The analysis which was performed in this initial project indicated that some interactions between the PEF subsystems are generally cost-effective while others are probably site-specific or can be improved upon. However, it was also concluded that various improvements should be made in the model to reflect different modes of operating the PEF to make it a more flexible concept.

DESCRIPTIONS OF TASKS IN CURRENT PROJECT

From the results of the initial project, it was concluded that certain refinements in the design of a PEF should be analyzed to give the PEF greater applicability as well as to improve its economics. Thus, it was decided to investigate in more detail certain aspects of the design and operation of the dry-land Energy Plantation subsystem. In addition, alternative ways of i[illegible]orating the wastewater treatment function i[illegible] PEF were to be studied.

One possible interaction within a PEF which was not considered in the initial project is the contribution of water from the wastewater treatment subsystem to the dry-land Energy Plantation. One of the original assumptions used was that PEF's would be sited in locations having an adequate natural supply of water for the woody biomass. However, it is possible that the wastewater treatment subsystem can augment the available natural rainfall or even supply the entire water requirement of a PEF. Thus, it might be possible to site a PEF in semi-arid or arid locations to expand its applicability. In the current project work is being done to study the water balance of a PEF and to determine the feasibility and costs of implementing an irrigation system to distribute the water.

The results from the initial project indicated that supplying the necessary nutrients to maintain the productivity of the land--particularly nitrogen--was a significant cost item in the economics of producing woody biomass. Because of the importance of nutrients, it was decided that the nutrient balance in the Energy Plantation model needed to be refined to predict the required amount of nutrients more precisely. In particular, because the leaves contain a high percentage of nitrogen compared to the wood, work is being done to include in the nutrient balance the effect of nutrient recycling via leaf fall and a more precise accounting of nutrient leaching.

Transportation was another significant cost item which appeared to have potential for cost savings through a more detailed analysis of alternative system designs. Thus, alternative methods for handling and transporting the woody biomass are being analyzed, such as pneumatic tube transport, chip baling, and alternative methods for drying the chips. In addition the transportation system is being analyzed in greater detail to see where cost savings might be achieved through optimization.

The initial results from studying the PEF concept indicated that the most significant credit resulting from the wetlands biological wastewater treatment subsystem was the wastewater treatment credit itself rather than the credit for the value of the gas produced. Thus, it became of interest to look for better ways of incorporating the wastewater treatment function within the PEF than via an algae pond. One way that this might be done is to apply the wastewater directly to the Energy Plantation. However, this process has limitations, with respect to both the particular location and local soil quality, and the composition of the wastewater. Work is therefore required to collect the necessary data on technical limitations and EPA regulations and to develop a model for this process. In addition other wetlands biological species besides algae have been suggested for wastewater treatment, and it was decided to investigate the possible use of these other plants to perform this function.

Finally, additional work is being done in this second project to look for new and improved technology to include in the power plant subsystem model. Additional potential sites are also to be identified where the new modes of operating a PEF--e.g., with irrigation or direct application of wastewater--would be applicable. The new complete PEF model will then be exercised to examine and compare the various modes of operation, and to determine the economic viability of PEF systems for sites displaying widely different local climatic and site-specific constraints.

ACCOMPLISHMENTS AND RESULTS

Additions to Energy Plantation Model

Water Balance and Irrigation - Even in areas in the United States where natural rainfall is sufficient for hardwood growth (25 inches or more), periods of water stress may occur during the growing season [9,10]. Lack of adequate moisture may have disastrous consequences on the survival and establishment of clones or seedlings. It has also been shown that yields of hardwood plantations can be increased by reducing water stress during the growing season. An irrigation subroutine is therefore being included in the PEF model to evaluate the cost-effectiveness of irrigation for site-specific conditions.

Irrigation requirements for a site are determined through a month-by-month water-balance analysis. The Blaney-Criddle method as adapted by the Soil Conservation Service [11] is used in this analysis. The method first determines the water consumptive needs of deciduous plantations for local climatic conditions. The irrigation requirements are then estimated by comparing these needs to local water inputs from rainfall. The monthly irrigation requirements are inputs to the irrigation subroutine. The peak monthly requirement is used to determi the peak capacity and capital cost of the irrig tion system. It is assumed that the irrigation required during the month of highest demand will be supplied through four weekly applications.

The operation costs are estimated on the basis of the total irrigation needs for the growing season. Self-propelled traveling sprinklers fed by underground mains are assumed in the model. Water can be supplied from wells, river or lake water, or effluents from a wastewater treatment plant. The response of the plantation is described by a relation of the form $y = ax + b$ where y = yield, x = number of growth days with sufficient moisture, and a and b = constants. The use of this relation is suggested by the work of Zahner [12].

Nutrient Balance - In the original nutrient balance model, the annual fertilizer requirement was calculated on the basis of replacing the nutrients removed in the biomass harvested less nutrient credits obtained from recycling the sludge from the algae pond and the ash from the boiler to the Energy Plantation. Credits were taken only for nitrogen and phosphorus in the sludge and calcium in the ash. Allowance was made for loss of nutrients by leaching, and application of fertilizer was assumed to occur only once per rotation.

The nutrients present in the biomass harvested were calculated on the basis of tables developed from data on (1) biomass distribution in leaves, stems, and branches of young trees as a function of age and (2) the nutrient composition of leaves, stems, and branches for young trees. Although the biomass-harvested term in the original model represented only woody biomass, the nutrient content calculated from the tables was for a weighted average of wood and leaf material.

Expansion and modification of the model have proceeded along the following lines: (1) more explicit definition of the amounts of wood and leaf material harvested; (2) inclusion of potassium and phosphorus in the ash recycle and calcium and potassium in the sludge recycle credits; (3) the use of both organic and inorganic leaching terms and updating of the values used in the original model; (4) addition of leaf recycle and nitrogen fixation parameters to the model; and (5) modifications corresponding to the specific cases of steady-state and transition conditions, as well as application of fertilizer on a yearly and on a rotation basis.

The nutrient balance model estimates the annual fertilizer requirement over the entire plantation once steady-state conditions have been attained. Prior to that, transition period conditions prevail. The model calculates the fertilizer requirements for the entire plantation over the length of the transition period. In each case, application of fertilizer may occur yearly if irrigation is to be used, or once per rotation, following harvesting, when irrigation is not practiced.

These different cases require modification of the model in the following ways: (1) during the transition period, the leaf recycle parameter includes an exponential decay function for decomposition of leaf material in any particular year, while in steady-state, the total amount of leaf material recycled in any particular year is assumed to be decomposed that year; (2) in the case of yearly application of fertilizer, simple organic leaching loss terms for leaf recycle and sludge parameters, and inorganic leaching loss terms for ash and fertilizer parameters are used, while these leaching losses are compounded for fertilizer application once per rotation.

The leaf recycle parameter is calculated in the growth model as the difference between leaf production and leaf harvest. The number of harvesting days in the growing season (May-October) will influence the amount of leaf material harvested. As the number of harvesting days in the growing season increases, more leaf material is removed from the site and less is recycled, increasing the requirement of inorganic fertilizer. To simplify calculations, it was assumed that the leaf biomass present at any time during the growing season is the same as that found at the end of the growing season. Leaf material harvested early in the growing season is therefore less than accounted for in the model; the model is conservative in this regard.

The exponential decay function used with the leaf recycle parameter during the transition period assumes a constant rate of decay [13]. In reality, decay will not be constant as the easily leachable components will decompose after leaf fall more quickly than predicted, and the more resistant materials will decompose more slowly.

The rate of breakdown of organic material is dependent upon environmental conditions and the chemical nature of the substrate being decomposed. Temperature and moisture have been found to be the most important environmental parameters, and C/N ratio and lignin content the most important chemical parameters [14,15]. Rather than assume one value for the rate parameter in the decay function to be operative for all regions and all species for Energy Plantation applications, a relationship will be used to predict the rate parameter from evapotranspiration and leaf lignin content data [16].

As the plantation design calls for establishment over a specified number of years, steady-state conditions for biomass production will exist at the end of this establishment period. However, steady-state conditions regarding leaf recycling will not be attained until a number of years later, depending upon the value of the rate parameter in the decay function; as the rate parameter increases, steady-state leaf recycling (annual decomposition equals annual input) will be reached more quickly. The requirement for supplemental fertilizer will be greater until steady-state leaf recycling occurs.

Organic and inorganic leaching parameters for each nutrient will be used in the model equations. In the case of yearly fertilizer application, these parameters will be entered as simple losses. In the case of fertilizer application once per rotation, the losses are compounded as power functions of the duration of the growth period. Thus, both the leaching values themselves and the length of the growth period will have a significant effect on the fertilizer requirement with application on a once-per-rotation basis.

A parameter for nitrogen fixation was added to the nitrogen equation in the model to account for input by nitrogen-fixing species possibly interplanted with Energy Plantation species [17], or used as cover crops. The expression used in the model allows this input only for planting units not harvested in any particular year. This parameter is replaced in the equations for the other nutrients by a term for recycling of nutrients in ash from the power plant.

Biomass Handling and Transportation

Biomass handling and transportation account for a large part of the operating costs and energy requirements for producing woody biomass on Energy Plantations. Transportation of wood from the growing site to the power conversion site, chipping of the wood, and drying prior to combustion can comprise as much as 70-80% of the total energy requirements, excluding energy conversion losses [18,19]. The cost of these operations ranges from 12-50% of the total operating costs, not including power generation costs [19,20]. Consequently, a good deal of the present work on the Energy Plantation subsystem of the PEF has been directed toward improvements in plantation transportation and handling systems.

Before improvements can be analyzed, the material to be handled must be clearly defined. The trees on the Energy Plantation will be a variety of hardwood species. The trees will be grown on short rotations of 2 to 4 years and harvested by prototype harvester-chippers. Previous studies show the use of such harvesters to be more economic than conventional feller-buncher-type harvesters. The resultant green whole-tree chips will have densities of 19 to 26 lb/ft^3 and moisture contents of 33 to 53% on a green basis. The amount of material handled ranges from 150,000 to 280,000 oven-dry tons/year or 270,000 to 590,000 green tons/year.

Improvements considered for the transportation and handling system include outdoor pile drying, woodchip baling, and the use of pneumatic tubes as an alternative to truck transportation. Outdoor piles of green woodchips are known to generate internal heat spontaneously which may dry woodchips. Drier chips would increase power conversion efficiencies and reduce transportation costs. Information in the current literature shows the internal heat to have an insignificant effect on moisture content and to result in a significant amount of chip deterioration, suggesting pile drying to be an inappropriate system alternative. However, the young, whole-tree, hardwood chips present on the plantations have larger amounts of bark and living wood cells per unit area than chips previously studied, which may result in spontaneous pile drying. While this alternative has presently been discarded for lack of proof, future research may show it to be a viable system alternative.

Harris Press & Shear Co. has developed a baling machine capable of compacting woodchips into bales resulting in a 2-to-1 volume reduction, a typical decrease in moisture content from 45 to 35% (green basis), and a weight reduction of approximately 18% [21,22]. Since transportation of woodchips is limited by weight, volume reduction would not benefit the transportation system, although weight reduction would. The decrease in moisture content would increase net power output by reducing drying requirements prior to or during wood combustion. However, baling equipment is expensive, labor requirements for the machine are high, and the cost of baling outweighs any apparent advantages.

Pneumatic tube systems capable of handling large amounts of bulk material have been developed by, and are available from, Tubexpress, Houston, Texas. The systems are closed-loop systems with automated loading and unloading stations and are designed to run 24 hrs/day 365 days/yr. Available information shows loading, unloading, and transporting of loose chips with large tube systems to be less costly than trucks; small tube systems are more costly [23]. However, due to the scattered distribution of plantation land at plantation sites, the cost of gathering the woodchips to the tube for transportation to the power conversion facilities raises tube costs above truck costs regardless of tube size.

Since tube capacity is a function of weight and volume, both of which can be reduced by baling, smaller tubes are required for the same amount of oven-dried wood when baled rather than loose. Pneumatic tube transportation of baled chips is less costly than loose chips. Unfortunately, the decrease in tube transportation costs is surpassed by the increase in baling costs, and transportation of baled chips by pneumatic tube is economically competitive with truck transportation of loose chips only at large distances. Pneumatic tube transportation of woodchips, whether loose or baled, has been eliminated from consideration in Energy Plantation transportation systems.

The original transportation system with loose, green woodchips and trucks has been shown to be more economic than the alternatives considered. Changes in the original system have been made to allow more efficient use of the equipment and to reduce overall costs. Previously farm tractor-dumpwagon combinations (small hauling units) collected chips from harvesters and conveyed the chips to the edge of the individual plantation parcels. From there the wood was transferred into semitractor-trailer units (large hauling units) and taken to the power conversion facility. Recognizing that small units are more economic on short hauls due to loading and unloading involved with large units, the small units are now used outside the parcels, generally when the parcels are within 2-4 miles from the centrally located power conversion site.

The actual maximum distance is a function of plantation size, the ratio of plantation land to geographic land, and land productivity as well as site-specific equipment and labor costs. Transportation from beyond that distance is more economic with large hauling units; the small units are restricted to the parcels. A minimum area for the outer area serviced by large units has also been defined. With less than 4 or 5 large units, investment in loading and unloading equipment for the units is not warranted and small units should be used throughout the plantation.

At the power conversion facilities, the woodchips are taken from the hauling units and fed to pneumatic pile builders which construct temporary storage piles. Chips are taken from these piles and dried in rotary dryers using the flue gas from the conversion facilities. Drying is done to minimize chip deterioration and to increase conversion efficiencies. From the dryers chips are directed to a second pile builder which constructs the primary storage piles. These piles contain up to 90 days worth of fuel. From there the chips can be retrieved for burning in the power plant boilers.

Direct Application of Wastewater

Direct application of municipal wastewater or sludge to the Energy Plantation is one of the synergisms considered in the PEF concept. Specific problems must be dealt with when using land application as a method of disposal of wastes. State regulations and guidelines generally require secondary treatment before land application (30 mg/l BOD_5 and suspended solids and no more than 200 fecal coliform organisms per 100 ml)[24]. The quality of surface and groundwater must also be preserved. This requirement imposes a limit on the amount of wastewater or sludge which can safely be applied to the land and therefore on the potential benefits resulting from wastewater application to the plantation. Some states also impose limits on the slope of the land on which wastes are applied. This regulation could reduce the amount of land available for Energy Plantation use.

To be fully effective, wastewater application on the plantation should take place during the growing season when irrigation is needed. Substantial seasonal wastewater storage capacity is therefore required. Another important consideration is the relationship between transportation costs of the wastes to the plantation and the cost of the land. Transportation costs for wastes increase rapidly with distance from the point of generation (the urban center) while land cost decreases sharply as the distance from the city increases. An optimum location resulting in the minimum combined costs of land and transportation must therefore be chosen for the PEF [25].

A subroutine was developed to analyze the cost effectiveness of wastes utilization on the plantation. The irrigation needs are determined as scribed before. If irrigation is supplied by st ɜr, the amount of nitrogen percolating nn ‹ceed EPA limits, or $(NW) \leq U + D + 2.7$ C_p where NW = nitrogen applied through wastewater, = nitrogen uptake by plants, D = denitrification, = percolating water and C_p = percolate nitrogen 10 mg/liter.

ch a nitrogen balance is performed for each te using median values for the components of e wastes (solids, nutrients, heavy metals) as ggested by Sommers [26]. The method described Lofty [27] which takes into account the resi- al nitrogen available from previous waste plications is used to determine the nitrogen ant uptake. If nitrogen percolation exceeds e EPA limits, the rate of application of waste- ters must be reduced accordingly, thereby miting the usefulness of wastes as a source of trients and irrigation water. A qualitative alysis indicated that sites such as Phoenix, , having high irrigation requirements, could t rely on wastewater only to supply their water eds.

tal application of heavy metals through sludge wastewater application cannot exceed the EPA mits if the plantation land is ever to be nverted back to farming. At rates of applica- on envisioned for most sites considered in the alysis, the potential lifetime of the planta- on before EPA limits are reached will be from to 80 years [25]. After that period, the land uld still qualify for farming.

e model for irrigation with wastewater com- ises the same elements as the original irriga- on model but includes also transportation of stewater by pipelines and storage in ponds. etreatment of the wastes before application is sumed to be performed by the municipality. The F is credited for the disposal of the wastes ter treatment.

alysis of Improved Combustion Equipment

evious PEF work concerning power generation uipment defined the conversion facility capa- lities in terms of equipment available at that me. Feedstock was limited to woodchips. erall conversion efficiencies were 17.0 to .5%, steam quality was 650 PSIA @ 750°F, load ctors were 20-25%, equipment size was based on 0,000-1,700,000 lb/hour of steam or 25- to 5-MW electric plants. Present work will consider e use of new equipment to improve operating nditions, thereby increasing overall efficiency.

e mix of new equipment considered was wood sifiers and gas turbines. While the combina- on of these two pieces of hardware is capable overall efficiencies as high as 70 or 80%, monstrated reliability, durability, and economy e lacking. The major problem lies in prevent- g harmful material from traveling out of the sifier and damaging the turbine.

d Bed Combustion (FBC) boilers and in- rate boilers were also considered. The ers have capabilities of simultaneously burning a variety of fuel, wet or dry, and can remove sulfur from stack gases without expensive air pollution equipment. Disadvantages include high energy requirements, difficulty with fuel distribution in combustion chambers, and particulate carryover. Available fluidized-bed boilers operate under the conditions defined in the previous PEF study, although manufacturers are willing to build equipment to required specifications.

Inclined-grate designs can also handle green fuels, should be able to handle a variety of fuels, allow more complete combustion of the fuel, and have demonstrated high reliability while in service. Inclined-grate systems in existence operate with a much higher steam quality (900-1400 PSIA, 900°F), and appropriate sizes are available from a number of manufacturers. Unless sulfur in the fuel feedstock is of major concern, inclined grates appear to be a better choice of boiler systems. The improved capabilities of this equipment over equipment previously available should increase overall conversion efficiency some 5 or 10%.

Alternative Wetland Biosystems

Algae were selected as the prime focus for an integrated subsystem for PEF because the technology of algal ponds is reasonably well developed and is based on years of experience with ponds for treatment of municipal and industrial wastes. However, algae present some major engineering problems in mass cultivation and harvesting.

For example, productivities are limited by the amount of carbon in the wastewater. Carbon dioxide or some other carbon source must be introduced if growth to the N or P limit is to be accomplished, which is desirable because this would allow the most effective use of waste nutrients for recycling to the Energy Plantation. Introduction of CO_2 to the pond system in the amounts required appears to be cost-prohibitive.

Construction and operation costs for ponds are high. For maximum biomass production and yield, the ponds must be shallow, requiring relatively flat land and extensive site preparation. Availability of flat land in marginal land areas applicable to Energy Plantation usage will be restricted. Mixing is required to keep the microalgae suspended and maintain productivity, so operating costs are high.

Harvesting of microalgae is a major technical and economic constraint. Effective methods resulting in high yields, such as centrifugation, are high in cost or require addition of unwanted chemicals which interfere with anaerobic digestion of the harvested algal biomass. Microstraining is less costly but less effective and requires species control to some degree, which has not yet been effectively demonstrated in open systems for long periods of time and results in reduced yields.

Aquatic plants not subject to these constraints include the floating and emergent freshwater macrophytes, such as water hyacinth (Eichhornia *crassipes*), duckweed (*Lemna sps.*) and cattail (*Typha latifolia*). Because they are floating or emergent, pond depth and turbidity are not important design criteria. Carbon dioxide can be obtained directly from the atmosphere, and harvesting should be easier than for microalgae because of the macroscopic size of the plants. To compare the relative merits of using these plants for the wetland biosystem in PEF, data were obtained from the literature on productivity and conversion to methane by anaerobic digestion.

Water hyacinth is a floating plant with large leaves which extend upward from the water's surface from clusters of roots beneath the surface. It thrives in warm sluggish waters in the southern United States and Central America, but is not winter-hardy in temperate regions. Although water hyacinth seems clearly superior to other aquatic plants both in productivity and in conversion to methane for the PEF subsystem, its geographic range, which is restricted to semi-tropical areas, is a disadvantage.

Duckweed is a minute floating plant with a wide natural geographic range. The fronds attain a maximum of 0.4 inches in length and width; several elongated roots hang from the underside of each frond. However, the plants are susceptible to damage by wind in unprotected open water areas, and productivity even under favorable conditions is far inferior to that of water hyacinth.

Cattail abounds in and near shallow ponds and marshes throughout the United States, growing even in northernmost states with severe winters. The plant is rooted to bottom mud and a portion of the foliage may be submerged. Dense clusters of tall spikes rise above the surface. Productivity is high. However, cultivation, harvesting, and conversion problems offset these advantages. Cultivation requires a fairly level bottom as shallow depth is necessary. Harvesting would probably require draining of the ponds, and conversion to methane efficiently is not possible due to the high fibrous component of cattail biomass.

Only water hyacinth appears to have significant promise as a PEF subsystem component. In warm climates, water hyacinth is potentially superior to algae as an aquatic component of a PEF system.

COMPLETION OF THE PROJECT

Work remaining to complete the project includes finishing the programming of the model additions, verifying the model, gathering data for a few additional sites, and running the model program to analyze and compare the different modes of designing and operating the PEF. The work product will be a comprehensive analysis of the economic viability of the PEF.

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NOTES

POTENTIAL WATER QUALITY IMPACTS FROM LARGE SCALE CROP RESIDUE HARVESTING

Michael F. Torpy, Loren J. Habegger
Shen-Yann Chiu and Susan Barisas

ABSTRACT

The design methodology and some of the results of a study on the location in the contiguous U.S. and effects of substituting solar for coal energy are described. A scenario similar to a maximum solar penetration scenario is addressed and the relative increases in sheet erosion with and without biomass residue harvesting are evaluated. Changes in erosion are projected by using the Universal Soil-Loss Equation. Some results are described in relation to the continuing studies which will address comparisons between coal and solar erosion and water quality impacts.

INTRODUCTION

Most of the aspects of solar technology implementation are regarded with favor in relation to environmental quality. One of the more significant adverse environmental impacts of solar technology is the increased erosion and the subsequent increases in non-point source pollution from biomass residue harvesting. The Energy and Environmental Systems Division of Argonne National Laboratory is assessing the possible extent of solar technology impacts as described in two scenarios developed by the Domestic Policy Review. The scenario discussed in this paper is similar to the maximum solar penetration scenario (high solar scenario) developed by the Department of Energy. In that scenario, 14.2 Quads of solar energy would provide 12% of the 114 Quads required by the year 2000. Of the total solar energy projected in that scenario, 5.2 Quads are projected from all biomass products, 1.5 Quads of which are from biomass residue harvesting. This accounts for 1.3% of the total energy demand in the scenario.

This study is designed to provide an overview of the crop lands available for biomass residue harvesting, the characteristics of the land affected by the harvesting, and the severity of the resulting impacts of erosion including water quality impacts. In addition, the methodology is designed to compare the differences in erosion within regions where biomass residue harvesting may be substituted for coal mining in a scenario. The study is ongoing and this paper discusses study and some preliminary results.

METHODOLOGY

The study relies on a national crop inventory of the U.S. Department of Agriculture [1] to provide a basis for locating the likely crop production. This inventory is coupled with a Stanford Research Institute Study [2], which describes the residue yield ratios and is used to define the amount of crop residue available for harvesting.

The crops considered in the analysis are:

1. Barley
2. Oats
3. Peanuts
4. Rye
5. Soybeans
6. Cotton
7. Wheat
8. Hay
9. Rice
10. Flax
11. Corn
12. Irish Potatoes
13. Sorghum
14. Vegetables
15. Sugar cane
16. Sugar beets

The location and acreage of each crop are defined at the county level and aggregated to the Land Resource Area (LRA) as defined by the U.S. Department of Agriculture, Soil Conservation Service [3]. The 156 LRAs in the contiguous U.S. are grouped according to similarities in land use, elevation, climate, water availability, and soil types.

The Universal Soil-Loss Equation, developed by Wischmeier and Smith [4], is coupled with the scenario siting data to describe the relative increases in soil loss projected for biomass residue harvesting. The relationships of the variables affecting soil erosion were derived from many controlled studies on experimental plots and small watersheds. The equation has the form

$$A = R \cdot K \cdot L \cdot S \cdot C \cdot P$$

where A = soil loss in tons per acre
R = rainfall factor (erosion index)
K = soil erodibility factor
L = slope-length factor

S = slope-gradient factor
C = cover and management factor
P = erosion control supporting practice factor

The equation predicts soil erosion and was developed to provide specific and reliable guides to selection of adequate soil and water conservation practices for farm fields.

Sheet erosion is the movement of soil resulting from raindrop splash and surface runoff. Other forms of erosion include channel erosion, in which soil is transported as sediment by flowing streams of water across the land; and streambank erosion, in which riparian soil is transferred by the energy of moving water. The degree of severity of sheet erosion is proportional to the intensity of rainfall and influenced by soil and crop-related factors that affect the subsequent movement of the dislodged soil. Thus, sheet erosion removes the lighter soil particles, organic matter, and soluble nutrients from the land and is a serious detriment to soil fertility and productivity, as well as to resulting stream quality.

The Soil-Loss Equation is relatively accurate for estimating the degree of sheet erosion that may occur on small watersheds. The applicability of the equation suffers as the size of a watershed increases because variables such as slope gradient, soil type, and cover factors must be averaged over much larger areas. As a result, the developers of the equation have strongly recommended that large watersheds be divided into small areas for computation of soil loss and that the resulting data be summed for the larger area [6]. This form of analysis, i.e., averaging the results to the LRA level, has been used by other researchers to estimate soil loss and sediment loading [7,8]. However, such results must be treated with caution. The technique is most useful in the comparative analysis of the effects of soil management practices rather than in determining the actual magnitude of soil loss at the LRA level.

The Soil-Loss Equation describes the relative change in sheet erosion during the impact period, i.e., the period from crop harvest to vegetative canopy reestablishment. To identify these impacts, specific C factors for each crop in each LRA were developed in this study to reflect residual harvesting conditions during the impact period. The C value for a specific crop varies among regions of the county depending on the erosive rainfall conditions and the period and degree of plant cover. To provide the basis for comparison, two C factor values were identified for each case: one was developed on the assumption that crop residues are left on the field and tilled into the soil, and the second was developed on the assumption that the residues are completely removed at the time of harvest.

The derivation of C values that accurately reflect existing crop management variables require site-specific data. In this study, typical good and average crop-management cases were specified and do not necessarily reflect the actual practices in a particular state of the LRA. In general, rotation of row crops and close-grown crops was assumed to be a good crop management and continuous production of one crop was assumed to be average management.

The study period for each crop, starting from the average harvest date through the following average planting date, was obtained from USDA Handbook 283, *Usual Planting and Harvesting Dates*.

In addition to the C factors, R factors describing the characteristics of the rainfall and snowfall during the study period were also developed. These factors, when used in the soil-loss equation, provide a basis for eventually estimating non-point source sheet erosion and subsequent sediment delivery to the stream.

The sediment delivery ratio, a factor that describes the portion of the eroded soil reaching the stream at a specified rate, can be applied to the Soil-Loss Equation. Use of the sediment delivery ratio is dependent on watershed size and soil type, and the ratio is usually applied only to small watersheds. To date, we have not found an acceptable data base for applying the sediment delivery ratio in studies that are national in scope. Instead, the ratio will be applied in case analyses for areas in which results of the study indicate that impacts related to residual harvesting could constrain energy development. In any event, it is important to identify increased sheet erosion because it represents the potential for the loss of a nonrenewable resource, as well as the potential for water quality deterioration in certain areas.

RESULTS AND DISCUSSION

In a case where good management practices were assumed, the results indicate that some geographic areas potentially have relatively high-impact areas, and these will be analyzed more carefully with regard to the environmental quality trade-offs from coal mining. Specifically, the states of Montana, North Dakota, and Wyoming have been identified in the study as areas with potential for high environmental impacts from sheet erosion after residue harvesting (see Table 1, Column A). Although these values are relatively high, comparison with similar values of the scenario surface-mining projections in the same region may indicate that residue harvesting is preferable because of environmental considerations.

Table 1. Results of Good Management Case

	% Increase In Soil Loss During Study Period	% Of Soil Loss Annually	% Increase During The Study Period In Soil Loss Per Acre Impacted
la.	64	15	20.9
riz.	36	8	12.0
rk.	61	17	20.7
alif.	34	25	22.2
olo.	48	13	17.5
onn.	30	8	0.2
el.	56	10	6.7
l.	32	4	0.5
eorg.	69	17	12.2
D.	35	18	14.8
ll.	80	14	4.0
nd.	75	17	4.5
owa	103	9	2.4
an.	73	22	13.7
ent.	65	17	9.9
ou.	50	14	14.1
aine	43	8	1.2
aryl.	52	11	8.1
ass.	31	8	0.3
ich.	48	6	8.9
inn.	68	8	0.6
iss.	63	18	31.7
o.	76	16	9.8
ont.	56	27	29.2
eb.	92	13	5.3
ev.	30	16	17.5
.H.	27	7	0.6
.J.	39	8	4.0
.M.	51	10	16.0
.Y.	44	8	6.9
.C.	64	14	10.0
.D.	67	43	42.6
H.	70	14	9.2
K.	55	22	24.0
reg.	30	18	16.0
enn.	61	14	10.3
.I.	34	8	0.1
.C.	59	14	12.6
.D.	73	19	15.9
N.	62	17	23.0
X.	65	17	45.8
T.	25	7	8.6
	47	9	21.4
ig.	62	16	10.7
ash.	24	13	12.1
.V.	63	14	7.7
isc.	62	7	3.3
y.	47	14	22.0

Columns B and C of Table 1 also indicate some results that will also receive a closer analysis. The study indicates that Iowa, Nebraska, Illinois and Indiana would experience large relative increases in sheet erosion as the result of harvesting large residue crops of primarily corn and soybeans. Many areas within these states are already subjected to severe non-point source pollution, and the projected large additional increases would indicate an essential need for strict erosion-control practices. Other results from the study indicate:

- soil losses by State range from 4 to 25% increase per year.
- soil losses per acre increase by 0.01 to 1.56 tons.
- soil losses per ton of residue range from 0.01 to 0.92 tons.
- greatest per cent increases may occur in Northern Plains.
- greatest actual increases may occur in Appalachian Regions and California.
- impact of crops varies with region and cropping practices.

WORK REMAINING AND FUTURE STUDY

The model for sheet erosion resulting from biomass harvesting and from coal mining is developed to the point of quantification for case analysis. When the coal portion of the scenario is complete, erosion from both sources will be compared within each scenario.

Some characterization of sediment delivery ratios will be coupled with estimation of the amounts of attached phosphorus and nitrogen nutrients. By considering the nutrients associated with increased sheet erosion, the aquatic effects and likelihood of eutrophication, as well as the effects from particulate sedimentation can be considered.

In this study, the biomass inventory in the scenario is based on historical crop and land-use data. The economics of crop production strongly determine the crops finally used for biomass harvesting. Because of the economics, it is expected that some crops will be replaced by others with a greater residue harvesting value and that marginal land will be brought into production. In the former case, the pattern of sheet erosion from residue harvesting would change and perhaps become exaggerated. In the latter case, the severity of erosion would greatly increase, causing extreme sediment and nutrient loading to the stream. These factors are relevant to an accurate analysis and will be addressed in continuing studies.

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CARBOHYDRATE CROPS AS SOURCES OF FUELS

W7405 eng 92

Starting Date: April 1, 1978

Edward S. Lipinsky
Battelle Columbus Laboratories
Columbus, Ohio 43201

ABSTRACT

Evaluation of yields and costs of sweet sorghum and sugarcane production using narrow interrow spacing yielded favorable results leading to a tentative "go" decision for development of sweet sorghum as an energy crop. Sugarcane's limited geographical range relegates it to a secondary position for now. Experimental production of sweet sorghum is being expanded into North Dakota (long day length and short growing season) and into Kansas and Nebraska drylands. Farm Bureau trials in Illinois, Iowa, and Ohio will test farmer acceptance. Grain sorghum-sweet sorghum hybrids to stretch the fermentation season are being evaluated. The benefits of narrow interrow spacing necessitate the development of integrated planting, harvesting, processing, storage, and conversion systems. Developments outside the scope of this paper are underway in these areas.

This paper provides a report on selected aspects of a program to develop carbohydrate crops as sources of fuels, in which approximately a dozen organizations are participating. Battelle Columbus Division acts as the program manager and spokesorganization for this diverse team. The team's diversity arises from its many technical disciplines, range of organizational types (research institute, USDA, universities, engineering companies, and consultants), and its geographical breadth (Fig. 1). Due to space limitations and the biomass production orientation of this session, the project team's activities in stalk processing, harvesting, sugar solution preservation, ethanol process development, corn products, and policy studies can only be mentioned. Separate reports on these activities are forthcoming.

DESCRIPTION OF TASK

The development of carbohydrate crops as fuel sources involves the development of a flexible system for solving near-term technology problems and for transferring the solutions to users. Difficult questions to be answered include: which specific crops shall be grown, where, which process and conversion systems, how much fuel should be produced, at what cost, who should develop and who should implement, and when? Although this task description appears infinite, economic and technical realities serve to limit the objectives.

OBJECTIVES: COST AND PERFORMANCE TARGETS

Although carbohydrate crops could be employed to provide fuels for cooking, drying grain, or home heating, top priority has been assigned to liquid motor fuels. It is these fuels that are subject to arbitrary curtailment and price changes.

A tentative goal of one billion to ten billion gallons of replacement for gasoline or diesel fuel has been selected. Although this will solve only approximately 10 percent of the problem, it is a worthwhile contribution. Ultimately, other renewable resources (especially trees) can be expected to contribute significantly more to the solution of the imported motor fuel problem, but both time and technological risks are involved. Ethanol and methanol from carbohydrate crops need to be ready for sale at approximately $1 per gallon and $0.65 per gallon, respectively. These prices are not competitive with $0.56 per gallon gasoline from petroleum at $15.50 per barrel; but, by the time significant numbers of alcohol plants have been built, $15.50 petroleum will be a fond memory. In setting performance targets, ethanol and methanol have a preferred position, because the 1978 Energy Act specified alcohol fuels for exclusion from federal excise taxes. However, any alternatives (such as ketone fuels) must be available for an equivalent price, considering what they contribute both in mileage and in octane rating. It is not sufficient to have significant quantities of ethanol and methanol at relatively reasonable prices. Those quantities and prices must be attained without undue adverse impacts on food prices.

These goals need to be achieved on a timely basis. The fuel system has enough lags in it to make two billion gallons per year a very ambitious goal for the year 1985. Attainment of 10 billion gallons of fuel per year by 1990 and 20 billion gallons by the year 2000 from carbohydrate crops would require an all-out effort that could be achieved only by a clear mandate to industry, solid financial backing, and a rapid deterioration in the

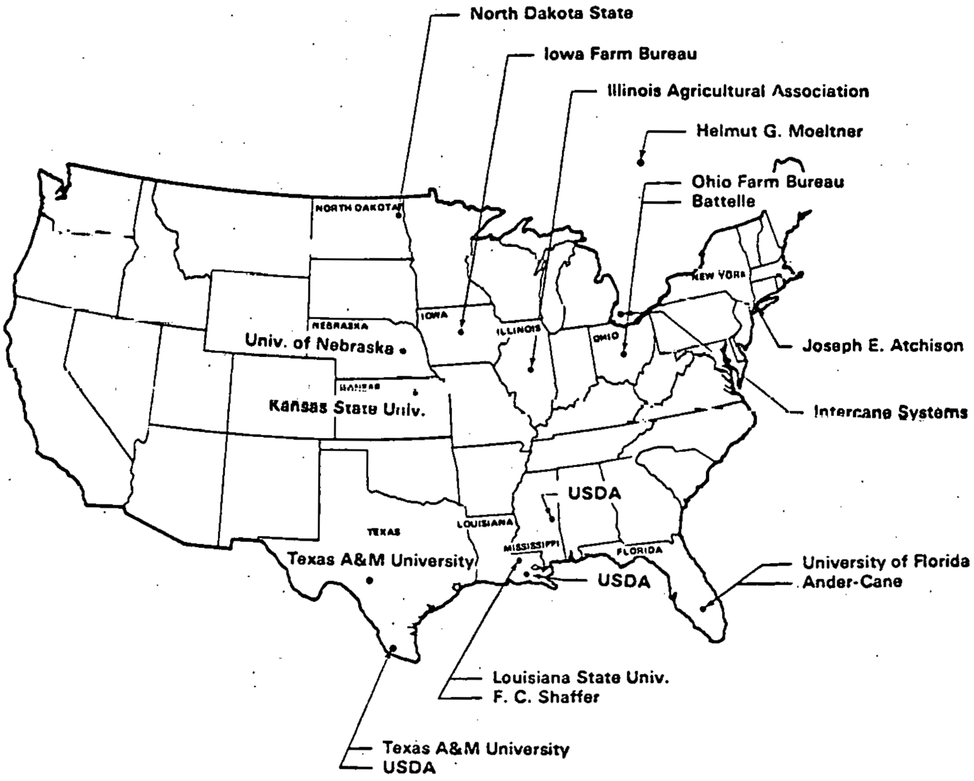

Fig. 1. Geographical Breadth of Team

availability of petroleum. These are not pleasant circumstances, but the DOE Biomass Energy Systems program would not be fulfilling its mission if such contingencies were not planned for.

If 10 billion gallons of ethanol are required, it would be necessary to use the grain from about 38 million acres of land that has a reasonably high productivity (100 bushels per acre). Twenty-one million acres of sugar crop land would be necessary to achieve a 10 billion gallon goal, provided that the land were highly productive (3.5 tons per acre of fermentable sugars). To put these numbers into perspective, current U.S. cropland totals approximately 330 million acres. Thus, diversion of approximately 10 percent of U.S. cropland to carbohydrate crops for fuel production could lead to achievement of the goal. The need for an agricultural technology development program and a program to improve fuels production from carbohydrate crops arises from the fact that such a diversion would have an adverse impact on food prices. The goal of the program then is to reduce the adverse impact by increasing the quantity of fermentable sugars obtainable from each acre, reducing the quality of land required for carbohydrate crops, and improving the overall economics of fuels production from carbohydrate crops.

APPROACH

If we want to win ballgames, the first step is to build up a winning team. Ever since the original systems studies indicated a need for fuels from carbohydrate crops, a major activity of the Battelle program has been to build a knowledgeable team that is indigenous to the regions where the selected crop could be grown (Fig. 1). The team members not only understand the problems of increasing yields and expanding acreage in their own regions, but also have contacts with farmers and other elements of agribusiness to transfer new technology as it comes along.

The potential of the various carbohydrate crops to achieve the ambitious goals outlined above has been evaluated. Crucial problems and barriers have been and are being identified. Concepts are being generated to overcome these problems and barriers. The concepts that withstand initial critical and skeptical evaluation are subjected to field testing so that conclusions and recommendations can be made and implemented.

KEY RESULTS AND ACCOMPLISHMENTS

Whether grain or sugar crops are used as the source of fermentable sugars, biomass raw material cost is the preponderant cost component. The need to pay depreciation, interest, and other capital charges and the need to return a profit in the form of a return on investment constitute the second largest component. The cost of fermentation, distillation, [illegible] her processing is small, compared with raw [illegible] al and capital charges. Although either dry or wet milling of grain is relatively inexpensive, processing of stalk crops such as sugarcane or sweet sorghum requires a heavy capital investment. Steam required for distillation, stillage drying, and other process activities must be generated either from biomass or from coal; otherwise, an inordinate amount of fossil energy is consumed in the manufacture of ethanol. A steam generating facility based on agricultural residues has a capital cost roughly double that of a fuel oil facility, and a coal steam plant is approximately four times as costly as a fuel oil plant. Fortunately, this bad news on ethanol costs is at least partly offset by rapidly declining estimates of the quantity of steam required to distill ethanol. In the past 2 years, estimates have decreased from approximately 50 pounds of steam to approximately 22 pounds of steam per gallon of fuel ethanol. Pressure distillation accounts for most of this decrease.

For a given facility, the length of the ethanol production season has a highly significant impact on ethanol cost. The cost of sugarcane-derived ethanol can vary from $1.20 per gallon with a 330-day season to $1.94 per gallon with a 90-day season.

Our agricultural economics and processing studies indicate that attainment of 10 billion gallons of ethanol from carbohydrate crops can be achieved at reasonable food price levels only by utilizing several complementary biomass resources. Sugarcane has a superb yield in its tropical and subtropical habitats, but little of the United States has the appropriate climate. We need a sugarcane-like crop that will grow over a wide geographical range. Sweet sorghum, which is a crop now grown for the production of pancake syrups, has many (but not all) of the desired attributes.

High yields of sweet sorghum can be obtained over a wide range of geographical latitudes and climates. As would be expected for a member of the hardy sorghum family, this crop is not nearly as sensitive to drought or frost as is sugarcane or corn. However, sweet sorghum is no panacea. It has a short harvesting and processing season, which will impose working capital problems, even if the technical problems of the perishability of its stalks and raw juice are overcome. It is expected that, when sweet sorghum is fermented and distilled, its stillage will resemble that which is obtained from molasses, and disposing of it would be difficult, whereas the stillage from corn fermentation is an asset due to its $128 per ton price as an animal feed.

A major emphasis in this program has been on increasing the yields of stalk crops by narrow interrow spacing. Conventional sugarcane and sweet sorghum agronomic practices use rows that are approximately 5 to 6 feet apart, so that conventional harvesting and planting equipment can be employed. Studies conducted by Battelle and its subcontractors and cocontractors indicate that sweet sorghum yields can be increased 40 to 100 percent and sugarcane yields can be increased 40 to 50 percent by narrow interrow spacing. For example (Fig. 2), at Belle Glade, Florida, the yield of sweet sorghum was roughly doubled, as was the yield of total sugars, when the rows were about 1.5 feet apart, instead of approximately 3 feet apart. The importance of selecting the appropriate cultivar* is shown also in Fig. 2. Not only does Roma give a much lower yield than does Wray, but the favorable effect of narrow interrow spacing is not pronounced with Roma.

As excellent as these sweet sorghum biomass and sugar yields were at Belle Glade, they were only half as great as what was obtained with sugarcane at the same location. The advantages of growing sweet sorghum in areas that are also amenable to sugarcane growth are that cheaper land may be used and, because the sweet sorghum processing season precedes the sugarcane season, the processing and conversion facilities may be able to function over a more extended season.

Results for sweet sorghum at all of our test locations are presented in Table 1. These maximum biomass yields are for small plots, and lower yields would be expected for commercial plantings. On the other hand, improved cultivars and nutrient practices may tend to bring the yields back up toward these levels.

For comparison, sugarcane at three locations (Table 2) showed much higher yields of dry biomass and fermentable sugars. The difference between Baton Rouge and Houma, Louisiana yields arises from differences in research methodology. The Baton Rouge experiments were directed toward adaptation of conventional planting and harvesting equipment, while the Houma experiments were designed specifically to achieve genuine equidistant spacing. Thus, the goal of the Houma experiment was to explore the upper limit of sugarcane production in Louisiana, whereas the Baton Rouge experiment was designed to determine the effect of compromising the ideal to facilitate commercialization.

To obtain high yields of fermentable sugars, it is necessary to supply the plants with correspondingly large amounts of water and nutrients. Fortunately, the sugar stalk crops do not require as much nitrogen fertilization as do grain crops, because less protein is manufactured in the stalk crops. Before taking into account byproduct credits that are expected to be between \$0.20 and \$0.35 per gallon of ethanol, the fermentable sugars cost component of ethanol is likely to be between \$0.64 and \$1.20 per gallon. At best, these estimates lead to ethanol that sells for \$1 per gallon and, at worst, \$2 per gallon. Furthermore, the cheapest ethanol is likely to constitute only a fraction of the total ethanol that could be manufactured from these sources. A realistic viewpoint is that all of our replacements for imported petroleum are going to be expensive when produced in large quantities, and ethanol will not be an exception.

The Tilby process separates sweet sorghum, sugarcane, or other stalk crops into two major fractions: the pith which contains the fermentable sugars desired for ethanol production, and rind fibers in the form of segments or strands. The process principle is shown schematically in Fig. 3. It is well suited for small-scale use.
The FY 1978 program at Battelle Columbus Laboratories conducted for DOE involved operation of a 15-ton-per-hour Tilby C-10 separator pilot plant at Clewiston, Florida. Sugarcane was used as the stalk feedstock, but the process was also demonstrated on sweet sorghum stalks from which the leaves had been stripped by hand.

The organization of the project team was as follows: Battelle Columbus Laboratories was the overall program manager. Ander-Cane, Inc. assembled and operated the stalk-processing facility using some equipment purchased from Intercane, Inc. and some donated by the U.S. Department of Agriculture's Belle Glade Experiment Station. Intercane, Inc. developed the Tilby separators, key alignment equipment, and was the source of technical information on the process. Moeltner and Associates developed the strand board substitute for plywood and prepared samples derived from the Tilby fiber output. Joseph E. Atchison Consultants, Inc. investigated the markets for the fibrous coproducts and performed part of the evaluation of the pilot plant. F. C. Schaffer and Associates, a major engineering and construction organization that builds sugar mills and alcohol facilities, developed cost estimate comparisons of the Tilby process with conventional processes. This rather complex team was employed in this critical evaluation because of its capabilities and because of its ability to transfer this technology to the private sector with minimal delay.

The following tentative conclusions arose from the FY 1978 critical evaluation: (1) the pith and rind fibers obtained from sugarcane and sweet sorghum appear promising on technical and economic bases; (2) some method to remove sweet sorghum's leaf sheath must be developed before the Tilby C-10 separator can be used with sweet sorghum; (3) the slitters on the C-10 stalk separator are underpowered, and the mechanism for slitting of the rind segments into strands needs to be redesigned; (4) although the specific low-cost detrashing system employed in this program is both inefficient and ineffective, a more sophisticated system that Intercane has designed might well overcome these problems; (5) the capital investment for a facility that uses the Tilby C-10 separators in place of conventional sugar mills probably would not be less than conventional sugar mills, but a higher-valued product mix would be obtained; (6) the smaller C-5 Tilby separator could process sweet sorghum stalks effectively and would require less detrashing than the C-10 device; (7) if the producer is willing to forego making plywood substitutes, the Tilby C-5 system might be scaled up to produce fuels with a muc smaller capital investment; (8) the economies scale in the Tilby system are smaller than with conventional stalk processing equipment; therefore, smaller mills closer to the source of supply

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* Agronomists refer to specially bred varieties as "cultivars".

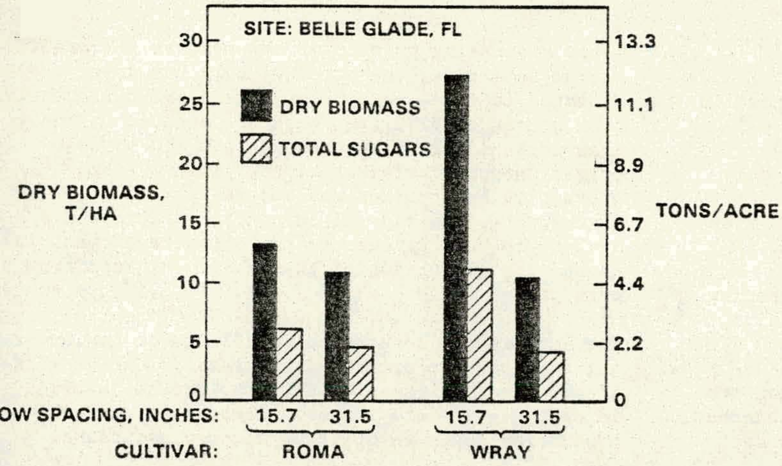

Fig. 2. Dry Biomass and Total Sugars Yields for Two Sweet Sorghum Cultivars at Belle Glade, Florida

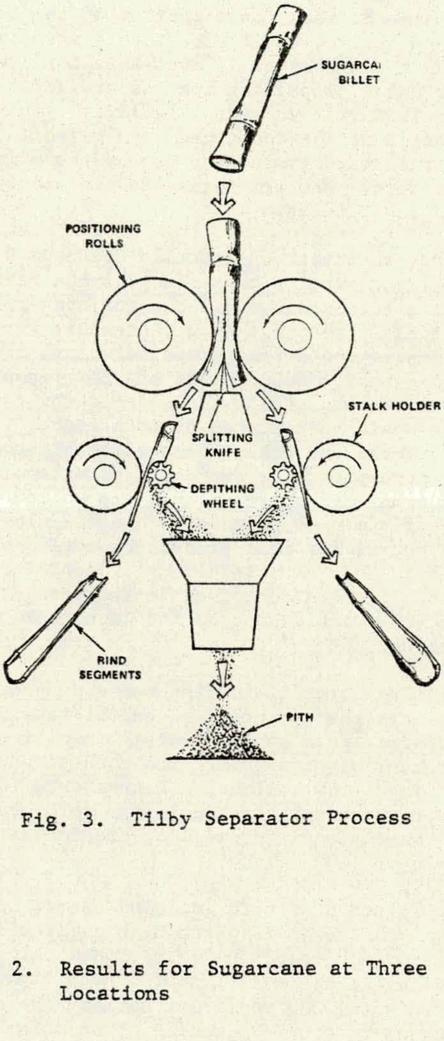

Fig. 3. Tilby Separator Process

Table 1. Results for Sweet Sorghum at All Test Locations

SWEET SORGHUM

LOCATION	CULTIVAR	MAXIMUM BIOMASS YIELD (T/HA)	MAXIMUM TOTAL SUGAR YIELD (T/HA)
BELLE GLADE, FL	WRAY	28.5	8.8
VESLACO, TX	SART	30.0	9.0
BATON ROUGE, LA	WRAY	28.8	8.5
N, MS	MN 1500	22.4	7.7 (MER 60-2)
COLUMBUS, OH	WRAY	22.2	6.5

Table 2. Results for Sugarcane at Three Locations

SUGARCANE

LOCATION	MAXIMUM BIOMASS YIELD (T/HA) PLANT CROP	MAXIMUM BIOMASS YIELD (T/HA) RATOON	MAXIMUM TOTAL SUGAR YIELD (T/HA) PLANT CROP	MAXIMUM TOTAL SUGAR YIELD (T/HA) RATOON
BELLE GLADE, FL	60.0	60.0	30.9	26.0
BATON ROUGE, LA	40.0	36.8	17.7	26.0
HOUMA, LA	61.2	—	21.1	—

appear feasible.

This critical evaluation led to the recommendation that DOE-funded Tilby development emphasize C-5-type Tilby separator system for use with sweet sorghum, sugarcane, and grain sorghum. Not only does the C-5-type system appear more suited for DOE's purposes, but it also avoids the more expensive capital investments of the C-10 system.

Intercane is redesigning the C-10 and is altering its detrashing system to include root removal. BCL recommends that investigation of the C-10 system be shelved until a revised system incorporating these changes is available for demonstration or until compelling reasons are advanced that DOE should invest in such a facility. This decision should not be construed as a rejection of the C-10 strandboard system for use by those in the private sector who are interested in manufacturing plywood substitutes.

Another task report by Battelle Columbus Division presents the results of a study of the feasibility of making corn grain available for conversion into motor fuel ingredients (e.g., ethanol) without undue disruption of the price structure for beef and other meat products that are now dependent on inexpensive corn grain. Corn grain for cattle feeding would be replaced by corn stover (the stalks, cobs, and other lignocellulosic portions of the corn plant), a byproduct of ethanol fermentation from corn grain. Thus, the unique capability of ruminant animals to use cellulose as the carbon source for meat production could be exploited, and, as a result of ethanol fermentation, the protein contained in the corn grain and in the yeast could be used for livestock production.

The primary objective of the research program was to evaluate the potential technical feasibility and economic promise for systems involving replacement of corn grain with corn stover and DDGS. The study includes a rather detailed evaluation of the cost of ethanol from corn grain, using conventional dry milling and fermentation technology.

The consumption of corn grain by cattle now occurs primarily in feedlots where high daily weight gains are required. The nutritional value (e.g., metabolizable energy) of stored corn stover in mixtures with DDGS would not satisfy the needs of feedlot operators. Upgrading of corn stover by the Iotech Process, Purdue Process, or alkali could increase the nutritional value but also would increase the cost of the cellulose material. Initial estimates of nutritional improvements afforded by these processes do not appear to correct the deficiencies of corn stover enough to offset the processing costs; however, these theoretical estimates need to be tested by feeding experiments.

The production of even one billion gallons of ethanol from corn grain in Midwestern states would lead to six times the present production of DDGS. This additional DDGS would compete with soybean meal, depressing prices for the soybean farmers. At the same time, the price of corn would be expected to rise, perhaps as high as $3 per bushel from a base of $2.27 per bushel. Because of these drastic anticipated changes in corn and soybean prices through implementation of this system, attractive beef prices may not result. Furthermore, the rise in corn grain prices and decline in DDGS prices are the opposite of what is needed for production of inexpensive ethanol from corn grain.

The anticipated adverse price effects of implementing this concept are based on classical price elasticity theory. These effects may not be observed in practice if land formerly devoted to soybean production could be utilized to grow additional corn.

Battelle has just completed a research program entitled "Identification of Key Policy Issues, Alternatives, and Implications Relating to Energy from Biomass". Some possible alternative energy policies were suggested. The implications of these alternatives relative to potential costs and benefits to the carbohydrate crop industry and U.S. taxpayers were examined.

One factor complicating the production of energy from biomass materials is that some forms of biomass, such as sugar crops and corn, contribute to the nation's food production. Use of these crops as an energy resource material meets with resistance from some observers who fear that U.S. food production capabilities could be threatened. However, during the periods of more than ample supplies and concurrent low prices for agriculture commodities, the U.S. taxpayer has subsidized crop producers in order that they might receive a "fair" income for their production. For example, in 1977 government payments to producers for the farm programs totaled $1.2 billion. Over the last 13 years government payments to producers have averaged $1.5 billion in real dollars. Several persons have suggested that there are alternative uses for the money going into the farm subsidy programs that might be more beneficial to the U.S. taxpayers and agricultural producers.

Various policy incentives programs have been suggested in Congress and elsewhere to stimulate biomass energy production and minimize distortions in the economy. Battelle's work suggests that the production of billions of gallons of ethanol would require a significant percentage of the current production of carbohydrate crops. The diversion of carbohydrate crops away from traditional food and feed uses would mean changes in the prices of agricultural products. Other obstacles also are present, including the development of efficient harvesting equipment, for development of high yielding varieties, and the development of multiple input crop processing and fermentation facilities. Policy incentives can help to minimize the various economic and institutional obstacles presently hindering the development of fuels from biomass systems; however, the policy alternatives should be related directly toward the most severe obstacles.

JT WORK

lthough there are anecdotal reports of qualita- ively high yields of sweet sorghum in various xtreme environments, an important first step is ɔ test its geographical range. For this reason, ɔrth Dakota State University, Kansas State Uni- ersity, and the University of Nebraska have been rought into the program in order to determine the ffects of short season, long-day length, and ryland conditions on several varieties of sweet ɔrghum. Commercial growers of corn and soybeans n Iowa, Illinois, and Ohio also are testing sweet ɔrghum varieties to determine whether standard gricultural equipment and practices will provide ɔod yields. These evaluations are being coordi- ated by the Farm Bureaus of the respective states, nich have many thousands of farmer members to nom the technology can be transferred rapidly.

ne need for a long fermentation season to achieve ɔw ethanol costs highlights the need for genetic mprovement. Dr. Fred Miller of the Texas A&M niversity has been experimenting with hybrids of rain sorghum and sweet sorghum, and these hybrids re being added to the program. Fermentable ugars in the stalk can provide in-season fermen- ation raw materials, while the starch stored in he grain head can stretch the season. The woody ibers can be used either for fuel or for plywood ubstitutes.

he introduction of a major new crop may require ew planting and harvesting equipment. In par- icular, the sweet sorghum leaf does not wither nd fall off, as does the sugarcane leaf. It is ecurely fastened to the stalk with a leaf sheath hat would jam a C-10 Tilby machine and decrease he effective throughput of conventional sugar rop stalk processing units. Modification of the -5 Tilby unit and the development of a new rocessing system that does not require leaf emoval will be undertaken to overcome these roblems.

reservation of sweet sorghum or sugarcane juices ithout energy-intensive evaporation is an impor- ant item requiring future work. Considerable rogress has been made at Battelle in a juice reservation process, for which patent rights are eing assigned to the Department of Energy.

ost evaluations of grain processing up to now ave pertained to dry milling of grain and sale of ll the byproducts as one animal feed called dis- illers' dry grains with solubles. As Carroll eim pointed out at a recent American Chemical ociety meeting [1], wet milling of starchy crops rovides the opportunity to obtain byproduct redits of corn oil, corn and gluten meal, and heir byproducts. The glucose output of this kind f facility is suitable for use in tower ermenters. Integration of wet milling of starch rops with ethanol production could have a highly avorable impact on ethanol costs, especially for he initial producers.

Motor fuels that contain at least 10 percent alcohol of biomass origin are not subject to the $0.04 per gallon federal excise tax. This provision of the Energy Act includes not only ethanol, but also methanol. Battelle's early studies included methanol from fibrous and leafy residues of sugar crop processing [2]. Methanol did not appear to be very attractive from this source, because large facilities were considered necessary to achieve economies of scale and because forest residues appeared to be cheaper. Furthermore, the opportunities to produce plywood substitutes from these residues appeared quite attractive. The development of the multisolid fluidized bed [3] as a means to make compact gasifiers that do not require oxygen facilities and the institution of incentives for methanol production bring methanol back into the picture for stalk crops. With stalks delivered to a central location, there are opportunities to integrate the production of ethanol with methanol, so that heat generated, but not fully consumed, in the gasification process can be used in the distillation of ethanol or in the evaporation of water from stillage or fermentable sugar solutions.

A nagging problem that could be converted into an opportunity is the need for disposing of stalk crop stillage. This product has only approximately 10 percent protein and is almost 40 percent soluble salts. Much of the valuable potash that is extracted from the soil by the plant is in the stillage. The high biochemical oxygen demand of stillage renders it a disposal problem, and at least 10 gallons of stillage is produced for every gallon of fuel ethanol. These factors have an adverse impact on commercialization decisions, because the technology does not exist to dispose of the stillage economically. The ideal recovery process would be one in which a protein-rich concentrate is obtained after separating the potassium salts for return to the soil. Alternatively, a facility that processed both stalk crops and grain could blend off at least some of the stalk crop stillage. The nutritional performance of such blends needs to be evaluated.

CONCLUSIONS

Carbohydrate crops have a significant role to play in an overall United States program to obtain liquid motor fuels from biomass. The total demand for motor fuels is so large that other resources, such as trees, also must make a large contribution. Sweet sorghum has been selected for development as an energy crop for the United States, but its shortcomings are also recognized. Yields and costs are being improved by narrow interrow spacing, and the geographical range is being tested by planting in various extreme environments. The genetic potential of the sorghum family is being tapped by making hybrids of sweet sorghum with grain sorghum, and improvements in stalk processing and ethanol production are leading to cost reductions and valuable byproducts. Americans should not expect that ethanol will be an inexpensive motor fuel; however, they can expect that this fuel will be available as the price of petroleum rises and the future availability of imported petroleum becomes ever more dubious. The U.S.' recent experiences with gasoline shortages indicate that availability of motor fuel is more important to the American motorist than its price.

ACKNOWLEDGMENTS

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ANALYSIS AND DESIGN OF A PILOT SILVICULTURAL BIOMASS FARM AT THE SAVANNAH RIVER PLANT

David J. Salo* The MITRE Corporation/Metrek Division
Jean-Francois Henry The MITRE Corporation/Metrek Division

ABSTRACT

In support to the Fuels from Biomass Systems Branch, MITRE/Metrek has developed an implementation plan for the installation and operation of a Pilot Silvicultural Biomass Farm of about 600 acres at the Savannah River Plant, South Carolina. The project included a comparative assessment of designs using either a trickle irrigation system or a traveling sprinkler system. Installation costs, operating costs, biomass production costs, net energy efficiency and sensitivity to design parameters were compared for the two designs. The trickle irrigation system is recommended for the Savannah River project: installation costs for this system are estimated to be \$1.1 million, operating costs are estimated at \$7.5 million for the 23 year lifetime of the project, biomass production cost is projected to be \$2.25/MM Btu for a large scale farm based on the Savannah River project design and the net energy efficiency is estimated to be 0.91. The selection of the trickle system results from site specific conditions at the Savannah River Plant which may not be found elsewhere.

INTRODUCTION

Silvicultural biomass energy farms are one of the options being considered as a long term source of fuel/feedstock. An extensive research program sponsored by the Department of Energy (DOE) is presently underway to develop the basic data required to implement energy farms -- selection of species, duration of rotations, biomass harvesting and handling, etc. There is a need, however, for pilot scale experiments to test new farm management techniques and new equipment and to generate the cost data needed to evaluate the economic potential of silvicultural energy farms. MITRE/Metrek has assisted the Fuels from Biomass Systems Branch of the DOE in planning and designing such a pilot silvicultural biomass farm (PSBF). This work is a continuation of the original MITRE study on silvicultural energy farms [1] and is discussed in two recent reports [2,3]. The program included four tasks: selection of a specific site for the pilot project, design and planning of the PSBF and estimates of installation and operation costs for the project. These tasks are discussed below.

SITE AREA SELECTION

The criteria adopted for selecting a site area for the pilot project included: a minimum of 20 to 25 inches of precipitation annually, the availability of arable land and the existence of terrain of no more than 30 percent grade. Because of the experimental nature of the PSBF and the desire to test a wide range of operating conditions, the availability of water for irrigation and the proximity of research facilities and personnel familiar with the site area were also considered desirable. DOE's Savannah River Plant (SRP) located near Aiken, South Carolina, meets these criteria and was selected as a site area for the PSBF. The SRP is attractive because of its location, size, multiple goal land use plan, existing research facilities and staff and because it is under DOE control. A word of caution regarding the selection of the SRP is however necessary: it may be difficult to extrapolate research findings from the SRP to other sites the Southeast because of its unique location in the sand hills of the Upper Coastal Plain.

FARM SITE SELECTION

Four alternative sites were proposed at the SRP. This choice was based on site quality and suitability for hardwood culture and on compatibility of biomass farming with other site uses. The Jackson site was selected for establishment of the PSBF. The general characteristics of this site are: fertile soil suitable for hardwoods, high productive potential, water availability, nearly level terrain and easy accessibility [2].

LAYOUT AND DESIGN

Because of the pilot nature of the project, it was decided that the PSBF design would include the capability for irrigation during a full six-year rotation. This is in contrast with the conceptual design proposed by MITRE for large scale farms in which irrigation is only provided during the first three years of a six-year growth cycle [1]. Early estimates indicated that irrigation would be a major component in the capital and operating costs of the PSBF. A comparative assessment of two designs, one including a trickle irrigation system, the other including a traveling sprinkler system, was performed to select the design most appropriate for the PSBF. The design characteristics for the PSBF are summarized in Table 1.

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Each irrigation system has specific features which will impact on the layout and capital and operating costs of the PSBF. The trickle system is easy to manage and operate and results in low operation and maintenance (O&M) costs. The trickle system does not require irrigation lanes and, therefore, for a given land area, about six percent more area can be planted than would be the case for a traveling sprinkler system. The trickle system applies water directly to the vicinity of the plants root system and, therefore, uses water more efficiently than the sprinkler system. For dense plantings, however, the capital cost of the trickle system will be high because of the cost of the drip tubing installed along each row of trees. For this reason, a 8' x 2' (8' between rows) planting pattern was proposed for the PSBF rather than the more conventional 4' x 4' pattern. Reverting to a 4' x 4' pattern would increase the capital cost of the trickle system by about $230 per planted acre. The impact of this change in planting pattern is discussed in the sensitivity analysis.

The traveling sprinkler system requires irrigation lanes which reduce the planting area per given land area. Operation costs are higher than for the trickle system because of the need for moving the sprinkler units from lane to lane during irrigation. Capital costs for self-propelled sprinkler systems are generally lower than for trickle systems ($/planted acre basis). However, the denivellation observed at the Jackson site, the friction losses within the water distribution network and the requirement for a pressure of 70 psi at the nozzle of the sprinkler result in a design pressure of about 200 psi for the water mains feeding the sprinklers. The need for high pressure mains will increase the capital cost of the sprinkler system. Increased pumping power will also increase the operating costs. Because of the conditions of the site, neither irrigation system has a clearcut advantage over the other one, and a detailed comparative assessment of the two systems for the Jackson site was performed.

Layouts for pilot farms including either irrigation system were drawn for the Jackson site. Both layouts have approximately the same overall area. The planted area for the sprinkler irrigation design, however, is smaller because of the presence of irrigation lanes (Table 2).

BIOMASS FARM INSTALLATION

The installation of the PSBF includes the following steps in approximate order: land clearing, land preparation, drainage ditching if needed, workroad construction, irrigation system installation and planting. For each operation, regional contractors were asked to estimate the cost and duration of the various operations. These data were used in costing and scheduling the installation of the farm.

Land clearing can be performed through timber sales at no cost to the project. Land preparation and improvement include root making, stump and slash burning, ash spreading and disking. The work roads are 25 feet wide, crowned and ditched. Drainage ditches are necessary in some areas of the farm. The characteristics of the irrigation systems are shown in Table 3. The higher capital cost of the sprinkler system results from the need for four wells rather than three wells for the trickle system and from the requirement for high working pressure. Year old bare root seedlings of the species proposed can be purchased from state nurseries and should be planted during the month of February. Two-row commercial planters can be used for this operation. Table 4 shows the estimated capital costs for the two designs. In both cases, the irrigation systems account for almost 60 percent of the total, and under the conditions prevailing at the site the trickle system is less costly to install than the sprinkler system. Planting of the first module is scheduled for February 1981 and will be followed by planting of the other modules at one year intervals. No scheduling problems are envisioned except for the first module. If the February 1981 deadline is to be met, some operations such as ordering of the seedlings must be initiated in the fall of 1979.

BIOMASS FARM OPERATION

Yearly farm operations include cultivation for weed control, fertilization, irrigation and harvesting. Table 5 summarizes the operational characteristics assumed for the PSBF. Periodic disking will be used to reduce competition from weeds. The higher application rates of fertilizer projected for the sprinkler system result from expected losses of fertilizer by run off and volatilization when it is applied through the sprinkler system. The higher cost of fertilization for the sprinkler system results from a combination of higher fertilizer costs and higher fuel costs for operating the sprinkler system. The difference in irrigation costs between the trickle and sprinkler systems result from the lower water utilization of the sprinkler system and from the high pumping costs associated with this system. No harvesting unit for silvicultural energy farming is available at present. Operational characteristics comparable to those of commercial wholetree chippers were used in the analysis [4].

Operating costs were estimated for the lifetime of the PSBF. It was assumed that the project would be pursued until all modules have been in operation for at least three (one first growth, two coppice) rotations, i.e., until the year 2003. Table 6 summarizes the projected operating costs for the lifetime of the project. The major component of the projected costs is personne
which includes a team at the site and a sup
ing team for analysis and reporting of the

cc[illegible]ted and for overall management of the project. Irrigation is the largest of the field related cost components followed by fertilization. Together these items amount to over 50 percent of the non-personnel related expenses. The total estimated operating cost of the project is 7.5 million dollars or about $300,000 per year and 8.4 million dollars or about $350,000 per year for the trickle and sprinkler designs respectively. These yearly budgets could be 10 to 15 percent higher than estimated.

BIOMASS PRODUCTION COSTS

Some of the capital and O&M cost components for the PSBF are not representative of the conceptual design proposed for large scale energy farms. For instance, personnel costs at the PSBF are higher ($/acre) than expected for a large scale farm, no land fees are included in the PSBF budget and irrigation is planned for the full duration of a rotation rather than for half of the rotation in the case of a large farm [1]. The cost data for the PSBF was adapted to reflect large scale farm operations. Cost items not included in the PSBF budget were taken from a previous analysis for a site located in Georgia [1]. These data were then used to estimate biomass production costs on a large energy farm. (Table 7). Biomass production costs are projected to be $2.25 and $2.94 per million Btu for the trickle and sprinkler designs respectively. With the exception of financial items, irrigation, fertilization and biomass handling and transport are the major components of the production costs.

NET ENERGY EFFICIENCY

The net energy efficiency (ratio of the net energy yield to the gross energy yield) was estimated to be 0.91 and 0.74 for the trickle and sprinkler designs respectively. The input energy includes the fuels for machinery operation and the energy for fertilizer synthesis.

SENSITIVITY ANALYSIS

The sensitivity of biomass production cost and efficiency to cost or management parameters relevant to the PSBF designs was analyzed. The impact of variations in installation costs, number of cultivations, amount of fertilizer and water, productivity, planting pattern and of the replacement of diesel fuel by electricity for water pumping was examined. Increasing productivity reduces biomass costs and improves efficiency for both the irrigation designs. Modifying the planting pattern from 8' x 2' to 4' x 4' has an unfavorable effect on the trickle design; the trickle design nevertheless remains more attractive (cost and efficiency) than the sprinkler design even for the 4' x 4' planting pattern. The impact of [illegible]s in other design parameters on the cost of [illegible]s and energy efficiency is relatively small and, in all cases, the cost of biomass and energy efficiency for the trickle design are always more favorable than for the sprinkler design. In principle, if electricity could be purchased at continuous service rates, electric pumping of irrigation water could be cheaper than diesel engine pumping. However, unless preferential rates for U.S. government activity is obtained, the interruptable rate that would be charged to the project would make the electric option more expensive.

CONCLUSION

Table 8 summarizes the comparison between the two irrigation systems designs. On the basis of this comparative analysis it is recommended that the trickle design be adopted at the SRP. It should be noted, however, that the preference for the trickle system results from site specific conditions (deep wells and terrain denivellation for instance) which affect the sprinkler system more than it affects the trickle system. Therefore, caution should be exercised when extrapolating the conclusions of the present analysis to other sites.

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TABLE 1

DESIGN CHARACTERISTICS OF THE PILOT SILVICULTURAL BIOMASS FARM

PLANTED AREA: ABOUT 600 ACRES

ROTATION PERIOD (INITIAL AND COPPICE): 6 YEARS

NUMBER OF ANNUAL MODULES: 6

PLANTING DENSITY: 2,725 TREES/ACRE (16 FEET SQUARE/TREE)

PLANTING PATTERN:

DISTANCE BETWEEN ROWS:	8 FEET	OR	4 FEET
DISTANCE BETWEEN TREES:	2 FEET	OR	4 FEET

IRRIGATION: TRICKLE/WELLS OR TRAVELING SPRINKLER/WELLS

FERTILIZATION: LIQUID FERTILIZER THROUGH IRRIGATION SYSTEM

SPECIES: AMERICAN SYCAMORE (PLATANUS OCCIDENTALIS)
EUROPEAN ALDER (ALNUS GLUTINOSA)
SLASH PINE (PINUS ELLIOTTII VAR. ELLIOTTII ENGELM)

PRODUCTIVITY:

BASE CASE: 8 DTE/A-Y
LOW CASE: 5 DTE/A-Y
HIGH CASE: 10 DTE/A-Y

TABLE 2

SITE CHARACTERISTICS OF THE PILOT SILVICULTURAL BIOMASS FARM

	TRICKLE IRRIGATION	SPRINKLER IRRIGATION
PLANTED AREA (ACRES)	620	570
WORK ROADS (ACRES)	39	47
IRRIGATION LANES AREA (ACRES)	—	28
RESEARCH AND STORAGE AREA (ACRES)	15	15
TOTAL SITE AREA (ACRES)	674	660

TABLE 3

CHARACTERISTICS OF THE IRRIGATION SYSTEMS

(BASE CASE)

	TRICKLE SYSTEM	SPRINKLER SYSTEM
PEAK DAILY WATER DEMAND (ACRE FOOT/DAY)	9.7	13.9
AREA IRRIGATED AT ONE TIME BY ONE UNIT (ACRES)	50	11
NUMBER OF UNITS OPERATING SIMULTANEOUSLY	3	5
SPACING BETWEEN DRIP TUBING OR IRRIGATION LANES (FT)	8	330
SPACING BETWEEN DRIP HOLES IN DRIP TUBING (FT)	2	--
LENGTH OF DRIP TUBING PER ACRE (FT)	5,460	--
WORKING PRESSURE OF IRRIGATION SYSTEM (PSI)	10	70 AT NOZZLE 170 IN PIPING
NUMBER OF 1,000 GPM WELLS PER SYSTEM	3	4
DEPTH OF WELLS (FT)	450	450
RATING OF DIESEL ENGINE DRIVE (HP)	117	250
INSTALLATION COST EXCLUDING WELLS ($/PLANTED ACRE)	530*	637
REPLACEMENT OF PLASTIC TUBING ($/PLANTED ACRE)	230	—
COST OF WELL, PUMP AND DRIVE ($/PLANTED ACRE)	435	674

*COULD REACH $760 PER ACRE FOR A 4' X 4' PLANTING PATTERN.

TABLE 4

INSTALLED CAPITAL COSTS FOR THE TRICKLE IRRIGATION AND TRAVELING SPRINKLER IRRIGATION SYSTEMS

(1978 DOLLARS PER PLANTED ACRE)

		TRICKLE IRRIGATION		TRAVELING SPRINKLER IRRIGATION	
	ITEMS	COST	PERCENT	COST	PERCENT
1.	SITE IMPROVEMENT (ROADS, DITCHES)	161	9.2	189	8.8
2.	LAND PREPARATION	218	12.5	232	10.8
3.	SUBTOTAL	379	21.7	421	19.6
4.	WATER APPLICATION	230	13.1	243	11.3
5.	WATER DISTRIBUTION	301	17.2	393	18.3
6.	WELLS	435	24.8	674	31.3
7.	SUBTOTAL	966	55.1	1,310	60.9
8.	SEEDLINGS	85	4.9	77	3.6
9.	PLANTING	55	3.1	55	2.6
10.	SUBTOTAL	140	8.0	132	6.2
11.	BUILDINGS, SERVICES, EQUIPMENT	266	15.2	289	13.3
12.	TOTAL	1,751	100.0	2,152	100.0

TABLE 5

OPERATIONAL CHARACTERISTICS OF THE PILOT SILVICULTURAL BIOMASS FARM

	TRICKLE DESIGN	SPRINKLER DESIGN
NUMBER OF CULTIVATIONS PER GROWING SEASON		
1st YEAR	7	7
2nd YEAR	5	5
COST PER CULTIVATION ($/ACRE)	10	10
FERTILIZER AVERAGE APPLICATION RATES		
NITROGEN AS N (LBS/ACRE-YEAR)	100	140
PHOSPHORUS AS P_2O_5 (LBS/ACRE-YEAR)	43	60
POTASSIUM AS K_2O (LBS/ACRE-YEAR)	143	200
AVERAGE COST PER ACRE-YEAR ($/ACRE YEAR)	49	98
IRRIGATION NEEDS (ACRE-FOOT/ACRE-YEAR)	1	1.4
COST PER ACRE-YEAR ($/ACRE-YEAR)	97*	137
HARVESTING (EXPERIMENTAL UNIT)		
CAPACITY (DRYTON/HOUR)	10	10
COST ($/ACRE)	290	290

*INCLUDES REPLACEMENT OF DRIP TUBING EVERY SIX YEARS

TABLE 6

SUMMARY OF THE PROJECTED OPERATING COSTS FOR THE PILOT SILVICULTURAL BIOMASS FARM (1978 Dollars)

ITEMS	TRICKLE SYSTEM BASE CASE 10^3\$	TRICKLE SYSTEM BASE CASE %	SPRINKLER SYSTEM BASE CASE 10^3\$	SPRINKLER SYSTEM BASE CASE %
CULTIVATION	278	3.7	265	3.1
FERTILIZATION	604	8.0	1,125	13.3
IRRIGATION	1,180	15.7	1,630	19.3
HARVESTING	507	6.8	483	5.7
REPLANTING	42	0.6	42	0.5
MAINTENANCE AND REPAIR	381	5.1	402	4.8
EQUIPMENT REPLACEMENT	247	3.3	247	2.9
UTILITIES AND SERVICES	282	3.8	282	3.3
PERSONNEL	3,984	53.0	3,984	47.1
TOTAL	7,505	100.0	8,460	100.0
AVERAGE COST PER PLANTED ACRE PER YEAR ($/Planted Acre-Year)	504		618	

TABLE 7

ESTIMATED BIOMASS PRODUCTION COSTS AND NET ENERGY EFFICIENCY FOR A LARGE SCALE ENERGY FARM

	SITE BASED ON SRP DATA			
	TRICKLE DESIGN		SPRINKLER DESIGN	
ITEMS	COST (1978$/DT)	PERCENT OF TOTAL	COST (1978$/DT)	PERCENT OF TOTAL
PLANNING/SUPERVISION	1.99	5.2	1.99	4.0
LAND BASE	1.90	5.0	2.02	4.0
LAND PREPARATION	0.91	2.4	0.97	1.9
ROADS	0.66	1.7	0.80	1.6
PLANTING	0.46	1.2	0.46	0.9
IRRIGATION	9.62	25.2	10.06	20.1
FERTILIZATION	6.06	15.9	14.44	28.9
WEED CONTROL	2.50	6.5	2.50	5.0
HARVESTING	1.11	2.9	1.11	2.2
BIOMASS HANDLING AND TRANSPORT	4.02	10.5	4.02	8.0
INTEREST	1.26	3.3	1.30	2.6
TAXES	3.90	10.2	5.32	10.6
RETURN TO INVESTOR	3.82	10.0	4.99	10.0
TOTAL	38.21	100.0	49.99	100.0
COST PER MILLION BTU (1978 $/MM BTU)	2.25	--	2.94	--
NET ENERGY EFFICIENCY	0.91	--	0.74	--

TABLE 8

SUMMARY COMPARISON OF THE TWO IRRIGATION DESIGNS

ITEMS	TRICKLE DESIGN	SPRINKLER DESIGN
TOTAL CAPITAL COST (THOUSAND 1978 DOLLARS)	1,086	1,280
TOTAL O&M COSTS (THOUSAND 1978 DOLLARS)	7,505	8,460
ESTIMATED BIOMASS PRODUCTION COST (1978 $/MM BTU)	2.25	2.94
NET ENERGY EFFICIENCY	0.91	0.74
SENSITIVITY TO DESIGN PARAMETERS	Variable*	Variable*

* BIOMASS COST AND ENERGY EFFICIENCY ALWAYS REMAIN MORE FAVORABLE FOR THE TRICKLE SYSTEM THAN FOR THE SPRINKLER SYSTEM.

NOTES

HERBACEOUS SPECIES SCREENING PROGRAM - PHASE I

DOE Contract ET-78-C-02-5035-A000

Starting Date: September 1978

Kathryn A. Saterson and Marianne K. Luppold

Arthur D. Little, Inc.
Cambridge, MA 02140

ABSTRACT

The overall objective of the DOE Fuels From Biomass Herbaceous Species Screening Program is to identify the herbaceous plant species that might be the most promising candidates for biomass production in approximately six different regions of the United States (Phase I) and to then conduct field screenings on each species in order to recommend the most promising species for future research and development (Phase II). The United States was divided into ten regions, based on the SCS Land Resource Areas, and the six that appear to offer the most potential for commercial biomass production were identified. A literature review and field interviews generated a list of 280 promising species, on which botanical and economic screenings were conducted. The result of the screening is a list, for each region, of up to 20 species that appear to have potential for high biomass productivity and which should be considered for inclusion in Phase II field screening. A total of 70 different species were included in the regional lists.

DESCRIPTION OF TASK

The Department of Energy's Fuels from Biomass Systems Branch is coordinating a two phase Herbaceous Species Screening Program in order to identify those species with the greatest potential for use on terrestrial energy farms. Arthur D. Little, Inc., was responsible for completion of Phase I of this program; Phase II has not yet been initiated. The overall objective of Phase I was to identify up to 20 promising species for whole-plant biomass production in approximately six different regions of the United States, while Phase II will involve field screening of those species.

OBJECTIVES/COST AND PERFORMANCE TARGETS

Objectives

The Department of Energy's Fuels from Biomass Program has sponsored considerable past and current research on the potential for energy production from silvicultural and aquatic botanical systems. The objective of the Herbaceous Species Screening m is to identify herbaceous species that have ial for high biomass productivity on terrestrial energy farms. The emphasis is primarily on previously unexploited, non-woody, annual and perennial species, and includes consideration of developed agricultural crops with potential for high whole-plant biomass productivity.

Phase I of the program involved identifying the most promising species for inclusion in a Phase II field screening in six regions of the U.S. and designing a three-year field screening program for those species.

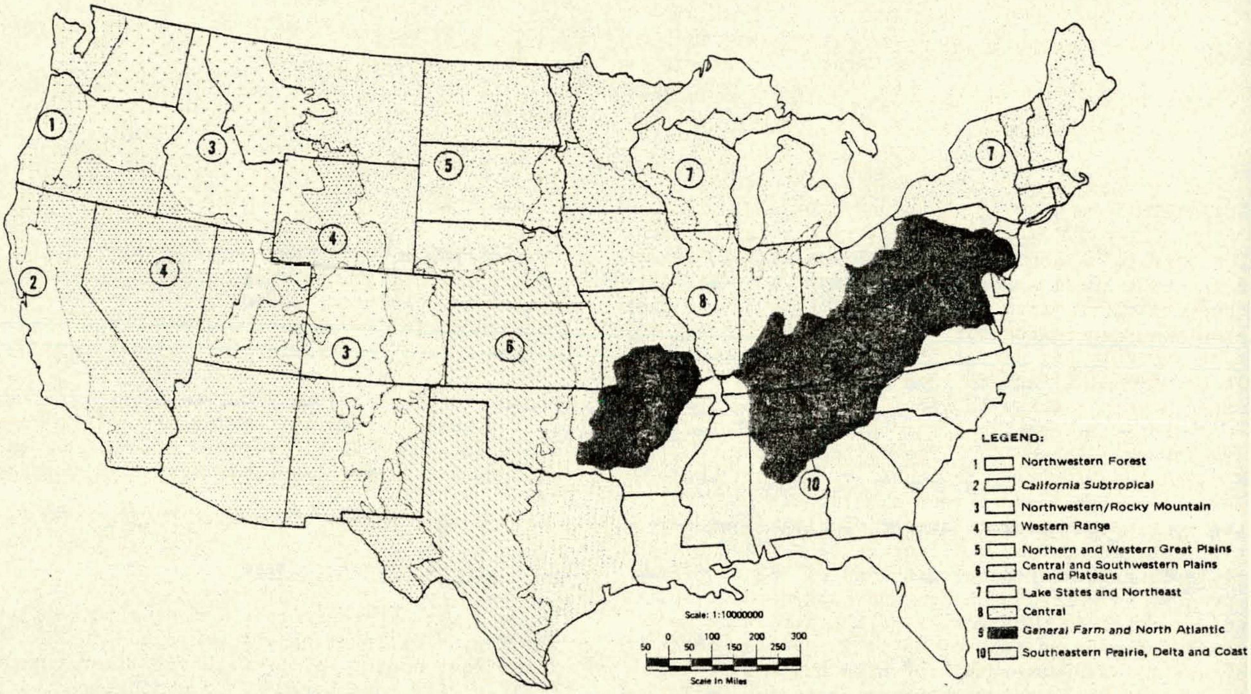
Cost and Performance Targets

Phase I took approximately 7 months to complete. The region delineation was performed during the first four months, along with the literature review, field visits, and personal communications with noted specialists. By the end of the fifth month, the literature review was complete as well as the botanical and economic screenings of the candidate species. Final selection of species recommended for field testing, which involved integration of the botanical and economic screenings, required an additional month. The results were presented at a public meeting during the 7th month. The cost for Phase I work was $123,000.

APPROACH

Literature Review and Data Collection

In order to determine which species should be included in this Phase I screening, we contacted over 150 agronomists and botanists through phone and field interviews. After describing the programs' objectives, we asked each person to recommend those herbaceous species in his/her region that would produce the greatest whole plant biomass. A list of 280 suggested species was compiled for the United States. The literature review provided limited information, especially in regard to the productivity, agronomic needs and economic parameters of unexploited plant species. Therefore, it was necessary to rely on extensive interviews with noted specialists in botanical, agronomic and other biomass areas for much of the cultural information and some yield estimates.



Source: Agriculture Handbook 296. March 1972.
Soil Conservation Service. U.S. Department of Agriculture.

FIGURE 1 REGION DELINEATION BASED ON MAJOR LAND RESOURCE AREAS OF THE UNITED STATES
(exclusive of Alaska and Hawaii)

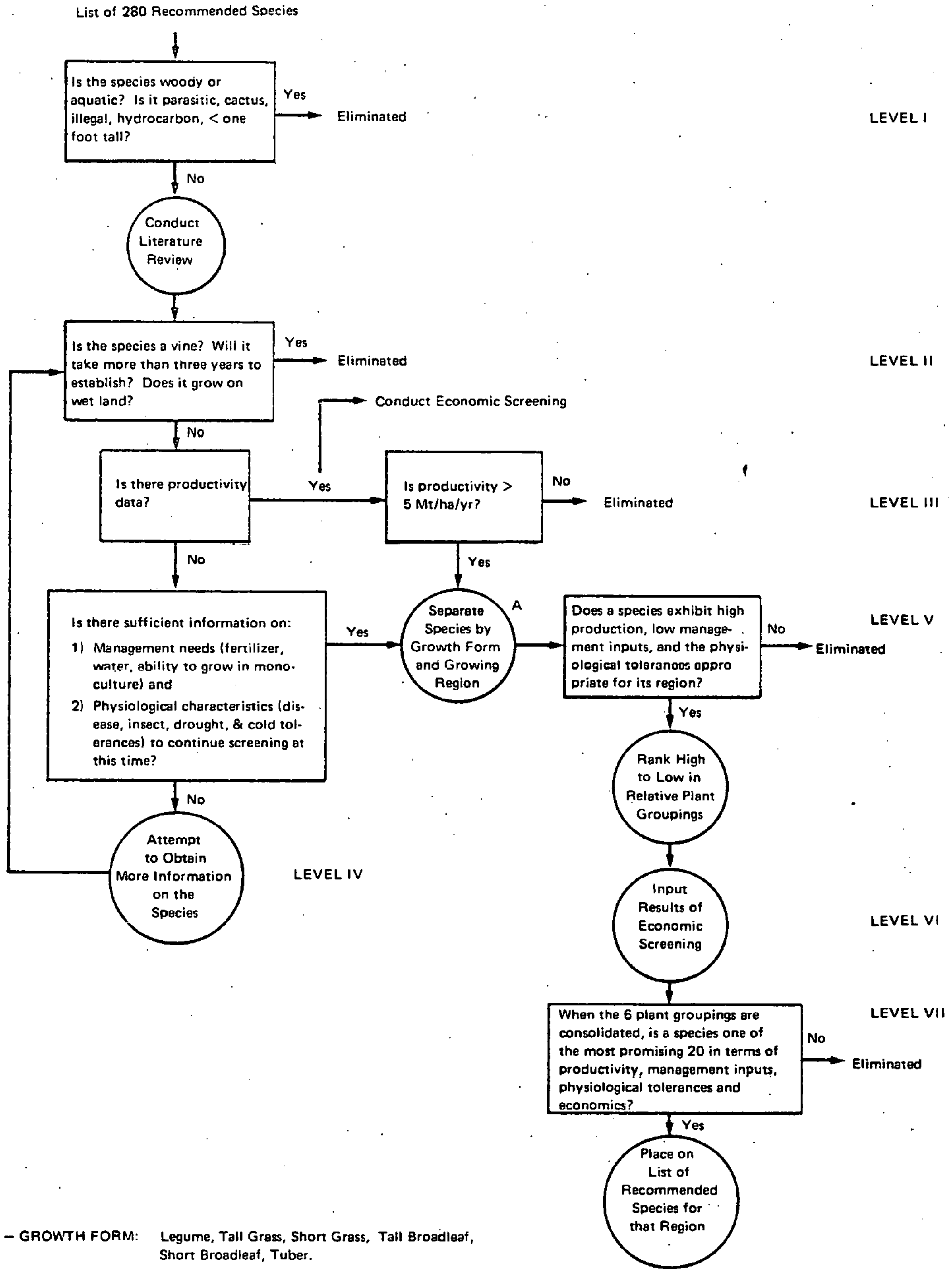
Region Delineation

In order to conduct the species screening and to evaluate regions with potential for commercial biomass production, the U.S. was divided into 10 ecological regions, based on the 20 Soil Conservation Service Land Resource areas [1] (see Figure 1). Although region delineation systems based on both agricultural characteristics and vegetation types were reviewed, the SCS classification was selected because it incorporated characteristics such as soil type, rainfall, temperature, and topography, as well as land use.

The regions that appear to offer the greatest potential for herbaceous biomass production on a commercial basis are Regions 5, 6, 7, 8, 9, and 10, primarily because:

- A majority of promising biomass candidate species are adapted to or grow productively under the conditions found in these regions,
- Climatological characteristics of these regions are conducive to crop production practices that are moderately energy intensive (i.e., less need for irrigation),
- Cropland is more abundant, therefore, the potential quad impact would be higher. For example, even in established agricultural regions such as 8 and 10 nearly 20% of the total land suitable for crop production was available for production in 1974 [2,3].

Regions 3 and 4 were ranked seventh and eighth and slightly lower than the top ranked regions. While these regions have many positive features in the southern portions (a good growing season with high insolation, some unutilized cropland and an abundant supply of rangelands) they ranked lower due to a lack of natural rainfall, necessitating irrigation for crop production in most unuti areas. This results in more energy intensiv agricultural practices and higher costs.



A – GROWTH FORM: Legume, Tall Grass, Short Grass, Tall Broadleaf, Short Broadleaf, Tuber.

FIGURE 2 SPECIES SCREENING METHODOLOGY

Regions 1 and 2 (Northwestern Forest and California Subtropical) were considered to have the least feasibility due to their relatively limited land availability and highly urbanized areas.

Species Screening

Botanical. The screening methodology used for each of the 280 recommended species is outlined in Figure 2. Since this program is focused on terrestrial, herbaceous plants, we began at

Screening Level I by eliminating aquatic and woody species. In addition, we eliminated species that are either 1) parasitic, 2) cacti (succulent), 3) illegal to grow, 4) being investigated under other D.O.E. contracts (i.e., hydrocarbon and sugar crop projects), or 5) less than one foot tall (because of low yield potential and minimum harvest height). Seventy-three species were included in these categories.

Information obtained from the literature review and interviews was used to describe a minimum of the following characteristics for each of the remaining species: productivity, monoculture ability, height, water needs, fertilizer needs and response, habitat, disease and insect resistance, and cold, salt, and drought tolerance. Since it was not possible to obtain information on many of the more important characteristics for all species (i.e., a productivity figure for each plant), the screening methodology was designed to give weight to other characteristics that influence the potential for high productivity, such as cultural requirements or environmental tolerances.

Once the information was assembled, several steps were taken to narrow the list of species at Screening Level II. Plants growing only in wetland areas[A] were eliminated due to environmental/regulatory considerations, lack of land availability, and potential harvesting difficulty. Vines were eliminated because of potential harvesting difficulty. Since the Phase II field screening program was defined for three growing seasons, we eliminated perennial species which are known to require more than three years to establish a productive stand. Twenty-four species were eliminated at this level and included cattails (<u>Typha</u> Sp.), bamboo (<u>Arundinaria</u> Sp.) and Kudzu (<u>Pueraria lobata</u>).

The next step (Screening Level III) was to examine available productivity information for each species. Plants with an annual productivity of less than 2.2 tons/acre (5 MT/ha) were eliminated. This was felt to represent a conservative minimum productivity requirement, given the fact that we recorded productivities from 0.4 to 22.3 tons/acre (1-50 MT/ha) and that the majority of plants produced between 4.5 and 6.7 tons/acre (10-15 MT/ha). Only eighteen species were eliminated on the basis of productivity. The economic evaluation was conducted on all species with productivity data.

Species with either a productivity of more than 5 MT/ha, or with sufficient data to continue screening, were then grouped by the regions in which they grow, with many species occurring in more than one region. There was insufficient data to adequately screen seventy species and they are awaiting further information.

The next step Screening Level V was to separate the regionalized species into six growth form groups: legumes, tall[B] and short[C] grasses, t and short broadleaves, and tubers. It was as that rough productivity and characteristic comparisons could be made among the plants within each group in order to facilitate further screening of species without specific yield information. In addition, the growth form groups enabled the assignment of proxies to previously unexploited species, which was necessary for the economic screening.

The species that appeared to have the least potential in terms of productivity, management, and tolerances were eliminated at Level V (12 species). The remaining species were then ranked in each growth form type and region in order of decreasing potential for biomass productivity. The results of the economic screening were incorporated before the final selection of species recommended for inclusion in Phase II field screening in each region.

<u>Economic</u>. The economic screening was conducted as a supplement to the botanical screening. Since a majority of the species screened have never been commercially produced, primary technical and economic data on inputs such as seeding rates, fertilization and pesticide schedules, labor, irrigation and mechanization needs, as well as harvestin costs, have yet to be developed for most of the candidate species. Therefore, the well-documented commercial crops shown in Table 1 were utilized as proxies for the six growth form groups of candidate species. This provided the basis for our development of indicators of the potential economic feasibility of each herbaceous species.

After species with reliable productivity estimates[D] were identified, analytical criteria were applied to the following three distinct indices:

- Estimated costs of production, based on prototypical crop budgets for each proxy;
- Estimated value of production, as a measure of a potential candidate species' relative competitiveness; and
- Estimated cropland availability in each species known regions of habitat.

A typical 1976 crop budget was developed for each proxy and is summarized in Table 1. A maximum cos of production per ton was established as the first screening criterion, in order to eliminate those

---

[A] defined as marshes, swamps, river banks, lake shores

[B] tall >4 feet (average height)

[C] short <4 feet

[D] Although reported yields from research plots are often higher than those obtained on a commercial scale, the maximum reported yields were utilized in this analysis. Since many of the candidate species have not had the benefit of genetic cultural research and development, it was f maximum figure would be a better indication future potential.

TABLE 1

AVERAGE COST OF PRODUCTION FOR PROXY CROPS, 1976[a]

rowth orm Groups	Proxy	Annual or Perennial	Average Yield[b] Tons/Acre (MT/ha)	Cost per Acre (per Hectare) ($)	Cost per Ton (per Metric Ton) ($)
all grasses	Corn	A	7.7 (17.3) [6]	159 (393) [7]	21 (23)
hort grasses	Wheat	A	4.4 (9.8) [6]	83 (205) [8]	19 (21)
all broadleaves	Sunflowers	A	6.7 (15.0) [9]	92 (228) [10]	14 (15)
hort broadleaves	Sugar beets	A	6.2 (14.0) [6]	525 (1296) [11]	85 (93)
egumes	Alfalfa	P	6.1 (13.7) [6]	143 (354) [8]	23 (26)
ubers	Potatoes	A	4.1 (9.2) [6]	613 (1514) [12]	150 (165)
verage Cost of Production for all Proxy Crops					52 (57)

Numbers in brackets refer to references at end of paper.

Whole plant on a dry matter basis.

pecies that would not offer enough incentive for ventual commercial production due to lack of ttractive margins for the producer. A value of 52/ton was the maximum for this analysis, based n the representative proxies' average costs/ton.

he relative competitiveness of biomass crops with ommercialized agronomic crops has a direct influ-nce on their potential for successful commercial-zation. It was felt that a species which could be efined as potentially attractive should have a inimum value/ha equal to or greater than 25% of he current value of all crops in a given region. n order to calculate the average value/ha, the alue of all crops per region and the amount of all ropland per region was initially determined. econdly, an average potential value of $41.25/dry on ($45.20/dry MT) for candidate biomass species as estimated [4,5], based on current values of ompetitive fuels. It was felt that this price was good indicator of the minimum price level at hich biomass would be available to the market, ssuming efficient conversion technology is readily vailable.

t was also felt that a species that grew only in limited area would not be as suitable as one dapted to a wider range. Given the present state f non-monoculture and non-commercial production of ost of the candidate species, we felt it would be rejudicial to eliminate a species on the basis of ts current habitat alone. There is much potential or introduction of some of these species into new reas. Therefore, cropland availability was used s a supplementary measure, but was not a species liminating parameter.

ach species was assigned a relative value for the egree to which it met each criterion. These values ere then multiplied in order to determine each urviving species' final economic ranking. Species ere segregated by ecological region, and arrayed ost likely feasible" to "least likely feasi- thin each region.

Integration. At Screening Level VI (see Figure 2) the economic feasibility assessments resulting from the economic screen were integrated with the botanical ranking. The six plant groupings were then consolidated and the most promising twenty species within each region, in terms of productivity, management needs, physiological tolerances and economic ranking, were selected.

The seventy recommended species included previously unexploited species with undetermined agronomic needs (such as redroot pigweed, lambsquarters, Colorado river hemp, and ragweed) as well as species which have received years of agronomic research and development (such as Bermuda grass, kenaf, reed canary grass and sudan grass).

KEY RESULTS/ACCOMPLISHMENTS

The Phase I species screening produced a list of seventy species that appear to offer potential for high biomass productivity and should be considered for inclusion in field screening trials in one or more regions. Although Regions 5 to 10 appear to offer the greatest potential for commercial biomass production, species suggested in Regions 4, 3 and 2 were also screened. The species that appear to offer potential for biomass production are listed alphabetically, by region, in Table 2.

The four species with the greatest reported productivities for Region 10 (Southeastern Prairie Delta and Coast) are Bermuda grass (Cynodon dactylon), kenaf (Hibiscus cannabinus), napiergrass (Pennisetum purpureum), and forage sorghum (Sorghum bicolor). Okra (Hibiscus esculentus), roselle (Hibiscus sabdariffa), and perennial ryegrass (Lolium perenne) were recommended only in this region. Napiergrass is a tropical species that is adapted to the southern portions of Region 10, while either the Coastal or Tifton 44 varieties of Bermuda grass, a warm season grass, could be grown throughout the region.

TABLE 2

SPECIES RECOMMENDED FOR INCLUSION IN FIELD SCREENING

Scientific Name	Common Name	Maximum Reported Productivity Dry tons/acre(MT/hectare)	Referenc
REGION 10 - Southeastern Prairie, Delta and Coast			
Amaranthus retroflexus	Redroot pigweed	—	
Ambrosia trifida	Giant ragweed	—	
Arundo donax	Giant reed	8.3 (18.7, wildstand)	[13]
Chenopodium album	Lambsquarters	—	
Crotalaria juncea	Sunn hemp	6.9 (15.4 Georgia)	[14]
Cynodon dactylon	Bermuda grass	12 (27 Tifton, GA Coastal variety)	
Festuca arundinacea	Tall fescue	3.7 (8.3 Alabama)	[15]
Hibiscus cannabinus	Kenaf	13 (29.2 Southeast)	[15]
Hisbiscus esculentus	Okra	3.3 (7.4 Georgia)	[16]
Hisbiscus sabdariffa	Roselle	7.6 (17 Georgia)	[9]
Lespedeza cuneata	Sericea lespedeza	5.0 (13 Georgia)	[17]
Lolium perenne	Perennial rye grass	4.9 (11)	[9]
Paspalum dilatatum	Dallis grass	5.4 (12 South)	[18]
Pennisetum americanum	Pearl millet	9.5 (21.3 Georgia)	
Pennisetum purpureum	Napiergrass	12.7 (28.5 Georgia)	[19]
Phalaris tuberosa	Harding grass	4.5 (10.1 Alabama)	[20]
Sesbania exaltata	Colorado river hemp	7.1 (16 Arkansas)	
Sorghum bicolor	Forage sorghum	12 (27 Louisiana)	[9]
Sorghum halepense	Johnsongrass	6.2 (14 South)	[18]
Sorghum sudanense x S. bicolor	Sorghum hybrid	7.9 (17.6 Georgia)	[21
REGION 9 - General Farm and North Atlantic			
Amaranthus retroflexus	Redroot pigweed	—	
Asclepias syriaca	Common milkweed	5.5 (12.3 Illinois)	[22
Bromus inermis	Smooth bromegrass	6.2 (14 W. Virginia)	[9
Coronilla varia	Crown vetch	4.5 (10 W. Virginia)	[9
Cynodon dactylon	Bermuda grass	7.1 (16 South)	[18
Dactylis glomerata	Orchardgrass	5.8 (13 W. Virginia)	[9
Festuca arundinacea	Tall fescue	5.4 (12 W. Virginia)	[9
Hibiscus cannabinus	Kenaf	8.3 (18.5 Appalachian)	[15
Kochia scoparia	Kochia	3.1 (7.0 Pennsylvania)	[23
Lathyrus sylvestris	Flat pea	5 (11.2 Northeast)	
Lespedeza cuneata	Sericea lespedeza	4.1 (9.1 Tennessee)	[24
Miscanthus sinensis	Eulalia	6 (13.5 Tennessee)	
Panicum virgatum	Switchgrass	6 (13.5 Kentucky)	
Paspalum dilatatum	Dallis grass	54 (12 South)	[18
Pennisetum americanum	Pearl millet	3.8 (8.6 Tennessee)	[25
Phalaris arundinacea	Reed canary grass	5.8 (13 Northeast)	[18
Phleum pratense	Common timothy	4.5 (10 Pennsylvania)	[26
Phytolacca americana	Pokeweed	--	
Sorghum sudanense	Sudan grass	5.0 (11.1 W. Virginia)	[27
Sorghum sudanese x S. bicolor	Sorghum hybrid	8.2 (18.4 Tennessee)	[15
REGION 8 - Central			
Amaranthus retroflexus	Redroot pigweed	--	
Ambrosia trifida	Giant ragweed	--	
Apocynum cannabinum	Hemp dogbane	--	
Asclepias syriaca	Common milkweed	5.5 (12.3 Illinois)	[22
Bromus inermis	Smooth bromegrass	5.6 (12.5)	[18
Chenopodium album	Lambsquarters	--	
Crambe cordifolia	Colewort	--	
Dactylis glomerata	Orchardgrass	6.9 (15.5 Indiana)	[15
Datura stramonium	Jimsonweed	--	
Festuca arundinacea	Tall fescue	7.0 (15.6 Indiana)	
Kochia scoparia	Kochia	6.7 (15 South Dakota)	

TABLE 2 (contd)

cientific Name	Common Name	Maximum Reported Productivity Dry tons/acre (MT/hectare)	References*
edicago sativa	Alfalfa	6.1 (13.7 Corn Belt)	[15]
anicum virgatum	Switchgrass	4.4 (9.9 Iowa)	[28]
halaris arundinacea	Reed canarygrass	7.6 (17 Indiana)	[18]
hleum pratense	Common timothy	4.6 (10.4 Ohio)	[15]
hytolacca americana	Pokeweed	--	
orghum bicolor	Forage sorghum	11.4 (25.6 Ohio)	[29]
orghum sudanense	Sudan grass	4.8 (10.8 Iowa)	[28]
orghum sudanese x S. bicolor	Hybrid sorghum	8.5 (19.1 Iowa)	[28]
ripsacum dactyloides	Gama grass	3.6 (8)	
EGION 7 - Lake States and Northeast			
mbrosia artemisiifolia	Common ragweed	--	
mbrosia trifida	Giant ragweed	--	
ndropogon gerardii	Big bluestem	3.8 (8.5 Northeast)	
pocynum cannabinum	Hemp dogbane	--	
sclepias syriaca	Common milkweed	5.5 (12.3 Illinois)	[22]
triplex patula	Spreading saltbush	3.5 (7.8)	[31]
vena fatua	Wild oats	--	
romum inermis	Smooth bromegrass	5.0 (11.1 Wisconsin)	[18]
henopodium album	Lambsquarters	--	
oronilla varia	Crown vetch	4.0 (9 Northeast)	
actylis glomerata	Orchardgrass	4.9 (11 Wisconsin)	[18]
chinochloa crusgalli	Barnyard grass	--	
estuca arundinacea	Tall fescue	--	
elianthus annuus	Sunflower	8.9 (20 Minnesota)	[9]
elianthus tuberosus	Jerusalem artichoke	14.3 (32)	
athyrus sylvestris	Flat pea	5.0 (11.2 Northeast)	
edicago sativa	Alfalfa	4.7 (10.6 Michigan)	[15]
halaris arundinacea	Reed canarygrass	6.1 (13.7 New York)	[28]
hleum pratense	Common timothy	4.8 (10.8 Wisconsin)	[18]
hytolacca americana	Pokeweed	--	
EGION 6 - Central and Southwestern Plains and Plateaus			
maranthus retroflexos	Redroot pigweed	--	
mbrosia psilostachya	Perennial ragweed	--	
ndropogon virginicus	Broomsedge	--	
rundo donax	Giant reed	8.3 (18.6 Texas)	[9]
othriochloa barbinodis	Cane bluestem	4.8 (10.8 Texas)	[30]
rotalaria juncea	Sunn hemp	9.5 (21.3 Kansas)	[14]
ynodon dactylon	Bermuda grass	5.4 (12.1 Oklahoma)	[32]
ragrostis curvula	Weeping love grass	--	
ibiscus cannabinus	Kenaf	14.7 (33 Texas)	
edicago sativa	Alfalfa	8.0 (17.9 New Mexico)	[33]
anicum dichomotiflorum	Fall panicum	--	
anicum coloratum	Kleingrass	3.8 (8.6 Texas)	[15]
anicum maximum	Guineagrass	--	
anicum miliaceum	Proso millet	--	
anicum virgatum	Switchgrass	10.0 (22.4 Texas)	
ennisetum americanum	Pearl millet	4.0 (9 South)	[18]
ennisetum purpureum	Napiergrass	--	
esbania exaltata	Colorado river hemp	11.2 (25 Texas)	
orghum halepense	Johnsongrass	6.2 (14 South)	[18]
orghum Sudanense	Sudan grass	3.7 (8.4 Texas)	[34]
GION 5 - Northern and Western Great Plains			
gropyron repens	Quackgrass	"high"	[35]
dropogon gerardii	Big bluestem	60 (13.4 Nebraska)	[36]
as syriaca	Common milkweed	5.5 (12.3 Illinois)	[22]
lus cicer	Milkvetch	7.2 (16.2 Montana)	[37]

TABLE 2 (contd)

Scientific Name	Common Name	Maximum Reported Productivity Dry tons/acre (MT hectare)	Refere
Bromus inermis	Smooth bromegrass	"high"	[35]
Coronilla varia	Crownvetch	5.8 (12.9 Colorado)	[38]
Dactylis glomerata	Orchardgrass	"high"	[35]
Helianthus annuus	Sunflower	8.8 (19.7 Minnesota)	[39]
Helianthus annuus x H.tuberosus	Sunchoke	12.7 (28.5 Texas)	[40]
Helianthus maximiliani	Perennial sunflower	--	
Helianthus tuberosus	Jerusalem artichoke	14.3 (32)	
Kochia scoparia	Kochia	6.7 (15 South Dakota)	[41]
Medicago sativa	Alfalfa	5.6 (12.5 North Dakota)	
Melilotus alba	White sweet clover	3.3 (7.5 Texas)	[42]
Panicum miliaceum	Proso millet	--	
Panicum virgatum	Switchgrass	6.4 (14.3 Nebraska)	[36]
Salsola kali	Russian thistle	7.1 (16 New Mexico)	[43]
Tripsacum dactyloides	Gama grass	3.6 (8)	
REGION 4 (Most Require Irrigation) - Western Range			
Amaranthus hybridus	Grain amaranth	2.3 (5.2 South)	[44]
Amaranthus retroflexus	Rodroot pigweed	--	
Ambrosia psilostachya	Perennial ragweed	--	
Ambrosia trifida	Giant ragweed	--	
Artemisia filifolia	Sandsage	--	
Bothriochloa barbinodis	Cane bluestem	4.8 (10.8 Texas)	[30]
Cucurbita foetidissima	Buffalo gourd	4.5 (10 Arizona)	
Cynodon dactylon	Bermuda grass	--	
Datura stramonium	Jimsonweed	--	
Eragrostis curvula	Weeping love grass	--	
Erianthus ravennae	Plume grass	--	
Medicago sativa	Alfalfa	8.0 (17.9 New Mexico)	[33]
Melilotus alba	White sweet clover	3.3 (7.5 Texas)	[42]
Panicum antidotale	Blue panic grass	8.0 (18 Arizona - without irrigation)	
Phalaris tuberosa	Harding grass	--	
Proboscidea parviflora	Devil's claw	--	
REGION 3 - Northwestern/Rocky Mountain			
Agropyron repens	Quackgrass	"high"	[35]
Amaranthus retroflexus	Redroot pigweed	--	
Astragalus cicer	Milkvetch	5.4 (12.0 Colorado)	[37]
Bromus marginatus	Mountain bromegrass	"high"	[35]
Kochia scoparia	Kochia	4.9 (11)	[45]
Medicago sativa	Alfalfa	3.6 (8 Washington)	[18]
Phalaris canariensis	Canary grass	--	
Phleum pratense	Common timothy	--	
Salsola kali	Russian thistle	4.5 (10.1 West)	
Sisymbrium altissimum	Tumble mustard	--	
REGION 2 (Most Require Irrigation) - California Subtropical			
Amaranthus retroflexus	Redroot pigweed	--	
Avena fatua	Wild oats	--	
Cynodon dactylon	Bermuda grass	--	
Datura stramonium	Jimsonweed	--	
Festuca arundinacea	Tall fescue	--	
Medicago sativa	Alfalfa	8.5 (19.1 California)	[46
Melilotus alba	White sweet clover	--	
Oryzopsis miliacea	Smilo grass	--	
Panicum maximum	Guineagrass	--	
Pennisetum clandestinum	Kikuyugrass	--	
Pluchea sericea	Arrowweed	--	
Sesbania exaltata	Colorado river hemp	--	
Sisymbrium altissiumum	Tumble mustard	--	
Sorghum bicolor	Forage sorghum	12.9 (29 California)	[40
Sorghum halepense	Johnsongrass	--	
Sorghum sudanense	Sudan grass	16.0 (35.9 California)	[33
Sorghum sudanese x S.bicolor	Sudan-Sorghum hybrid	14.1 (31.7 California)	[15

*Unreferenced yields are estimates obtained in field interviews.

he species with the highest reported yields in egion 9 (General Farm and North Atlantic) are mooth bromegrass (Bromus inermis), Bermuda grass, enaf, eulalia (Miscanthus sinensis), switchgrass Panicum virgatum), reed canary grass (Phalaris rundinacea) and a sorghum hybrid (Sorghum sudanense S. bicolor). Eulalia is the only species that as unique to this region. Region 8 (Central) ffers potential for growing many productive cooleason grasses such as orchardgrass (Dactylis lomerata), tall fescue (Festuca arundinacea) and eed canary grass. Colewort (Crambe cordifolia) s the only species that was recommended solely in egion 8.

he maximum reported productivities for the species ecommended for Region 7 (Lake States and Northeast) re somewhat lower than those reported for Region 0. The promising species for this region include mooth bromegrass, orchardgrass, sunflower and reed anarygrass. The two species that are recommended nly in this region are common ragweed (Ambrosia rtemisiifolia) and spreading saltbush (Artriplex atula). Common ragweed is a self-seeding annual hat tolerates dry soils and grows in dense natural onocultures. Sunflower, which has the greatest eported productivity in this region, grows on dry, andy loams and is drought tolerant.

our of the species recommended for Region 6 (Cenral and Southwestern Plains and Plateaus) are nique to that region: broomsedge (Andropogon irginicus), fall panicum (Panicum dichomotiflorum), leingrass (Panicum coloratum) and guineagrass Panicum maximum). Guineagrass (for which yields f 30 MT/ha have been recorded in Nigeria) is a erennial that could grow in southern Florida, exas and California. The species with the greatst reported productivities in this region are iant reed (Arundo donax), sunn hemp (Crotelaria uncea), kenaf, switchgrass, and Colorado river emp.

he species that offer high productivity potential n Region 5 (Northern and Western Great Plains) inlude milkvetch (Astragalus cicer), sunflower, sunhoke (Helianthus annus x H. tuberosus), kochia Kochia scoparia) and Russian thistle (Salsola ali). Sunchoke and perennial sunflower (Helianthus aximiliani) are the two species recommended only n this region.

ince less than 20 candidate species remained for egions 4, 3, and 2 (Western Range, Northwestern ocky Mountain, and California Subtropical) at creening Level VI (see Figure 2), all of those reaining species were included on Table 2. Mountain romegrass (Bromus marginatus) and canary grass Phalaris canariensis) are recommended only in egion 3, while smilo grass (Oryzopsis miliacea), ikuyugrass (Pennisetum clandestinum) and arrowweed Pluchea sericea are the three recommended species nique to Region 2.

ost of the species offer potential in more than one egion, while twelve species are recommended for our or more regions. They are: redroot pigweed, ragweed, common milkweed, smooth bromegrass, Bermuda grass, orchardgrass, tall fescue, kochia, alfalfa, switchgrass, common timothy, and hybrid sorghum.

Included in the Phase I work was a preliminary design for the Phase II field screening. The purpose of the field testing is to determine the biomass productivity of the recommended species and to generate information on the environmental and cultural parameters influencing that productivity. The results of Phase II will determine which productive species can be established at a reasonble cost. Our field screening design considers two categories of herbaceous species: agronomically certain and previously unexploited. The primary variables introduced in the initial field testing program will be harvest time and frequency. Data will also be collected on green weight, dry weight, BTU content, disease and insect problems and environmental conditions.

FUTURE PLANS

Although Phase II was originally planned for three years, it is now anticipated that field screening of promising herbaceous plants will be conducted for a greater number of years, by subcontractors in different areas of the U.S. It is anticipated that an RFP will be issued and subcontractors selected prior to the 1980 growing season.

ACKNOWLEDGEMENTS - Significant contributions were made by case team members: Robert E. Lee, Fred W. Besley, Kate Scow and Edmund C. Payne.

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Session IV B

LIQUID FUEL AND CHEMICAL PRODUCTION FROM CELLULOSIC BIOMASS - HEMICELLULOSE RECOVERY AND PENTOSE UTILIZATION IN A BIOMASS PROCESSING COMPLEX,

DE-AS05-79ET23051

R. P. Chambers, Y. Y. Lee and T. A. McCaskey
Auburn University, Auburn, Alabama 36830

ESCRIPTION OF RESEARCH PROGRAM OBJECTIVE

he objectives and research program tasks include he following:

. Develop a selective acid hydrolysis process uitable for large scale applications which can electively extract and recover the hemicellulose ortion of biomass in the form of its constituent ugars (primarily xylose).

. To utilize this hemicellulose recovery process s a pretreatment process for cellulose hydrolyis--the solids from which the hemicellulose has een removed have a much higher susceptibility to ellulose hydrolysis.

. To experimentally screen and isolate microrganisms that can ferment the xylose-rich acid ydrolyzate with good yields to liquid fuel onstituents, chemical by-products and fuel interediates.

. To develop a process for the conversion of cid hydrolyzate sugars to fuels, by-product hemicals and fuel intermediates.

. To coordinate the processes described above ith the remaining processes under study elsewhere n order to develop an integrated biomass refining omplex.

ARGETS

. Process and Equipment Specifications for the elective acid-catalyzed hydrolysis of hemicelluose including separation and recovery of constiuent sugars in a large scale biomass refining omplex. This process shall produce constituent ugars at investment and operating costs such that he fuels and chemicals can be produced at econoically satisfactory prices.

. Process, plus Microorganism Isolation and haracterization, for the fermentation of hemiellulose constituent sugars (selective acid ydrolyzate) at attractive yields to liquid fuel omponents, by-product chemicals and fuel interediates. The process shall produce fuels and hemicals at investment and operating costs that nomically viable.

3. Process for the conversion of fuel intermediates (such as butanediol) to gasoline extenders at economically viable costs.

APPROACH

The approach of this research program is to perform experimental research process and equipment designs and process economics in an interactive and coordinated manner.

Hemicellulose utilization research has been performed with large numbers of batch autoclave experiments. The batch reaction studies have determined conditions at which maximum xylose yield from hemicellulose can be obtained. Extensive mapping of xylose and by-product formation (furfural glucose, polysaccharides and acetic acid) has been accomplished throughout the projected operating range of the hemicellulose hydrolysis reaction. The kinetics of the sequence of reactions occurring in the hydrolysis are now under investigation to properly design a large scale unit. The batch studies, of necessity, were performed at lower solid/liquid ratios than those of a large scale process. Consequently, an experimental reactor is being designed and constructed to obtain data at industrially feasible solid/liquid ratios.

Experiments have been conducted to assist in the selection of suitable equipment for the subsequent separation of the solid and liquid products.

Contacts were made with equipment manufacturers to determine the suitability of pulp digesters and washers for use in hemicellulose hydrolysis.

A screening program was developed to select for microorganisms capable of fermenting hemicellulose hydrolyzate. Five microorganisms were isolated from cellulosic waste which grew in the hydrolyzate. One microorganism which we have designated AU-1-d3 ferments D-xylose to ethanol, acetic acid and 2,3-butanediol. The yield of fermentation products is about 60% based on xylose fermented.

Growth conditions of culture AU-1-d3 were studied to optimize growth and fermentation potential of the organism. Growth conditions were characterized with respect to pH, xylose concentration, nitrogen for growth, temperature, and growth rate.

Fermentation studies were conducted using pure xylose, as a preliminary step of fermenting hemicellulose hydrolyzate. With AU-1-d3 a number of bench scale batch fermentations were carried out with the control of pH, temperature and D.O. in order to optimize these parameters. Column fermentation technology was then employed to improve the fermentation process providing a number of benefits including reducing fermentation time and raising product concentration.

KEY RESULTS

The following is a summary of the results for each of the four major tasks which are described in the previous sections.

Hemicellulose Conversion to Xylose by Selective Acid Hydrolysis

This section deals with the hydrolysis of hemicellulose from oak hardwood using dilute aqueous H_2SO_4 as the catalyst. The objective here is to recover the carbohydrates in hemicellulose by maximization of the xylose yield in the hydrolyzate. Since hemicellulose is an amorphous solid which consists of a xylose backbone (xylan) with side chains of various sugars [1,2], it hydrolyzes much faster than crystalline cellulose. This makes it possible for the reaction to be selective with respect to the hemicellulose fraction. Conditions must be chosen so that the conversion of hemicellulose into xylose is favored over the conversion of cellulose into glucose, the decomposition of xylose into furfural, and the production of by-products [3]. Since hemicellulose accounts for 10-40% of the total weight of various biomass materials [4,5], it is important that the carbohydrates in hemicellulose be recovered.

Maximization of Xylose Yield: Results of the investigation show that a maximum xylose yield of 83% can be obtained by the hydrolysis of hemicellulose in oak hardwood at 150°C and 0.20% H_2SO_4 by weight. The hydrolysis reactions were carried out in a batch autoclave containing a 10 to 1 liquid to solid ratio (volume of liquid to mass of oven dry solids). Wood chips were used which were retained between A.S.T.M. (Tyler) 4 and 8 mesh screens. Figures 1 and 2 show the effect of temperature and acid catalyst concentration on the yield of xylose. As the temperature is increased the maximum xylose yields for each acid catalyst concentration occur at earlier times and the decomposition of xylose is more rapid. It is shown that, at 150°C, an H_2SO_4 concentration of 0.20% is required before a relative maximum will occur within three hours of hydrolysis time.

The major products of oak hydrolysis at the conditions for the maximum yield of xylose are given in Table I. Results from this table show that hydrolysis at 150°C and 0.20% H_2SO_4 is selective with respect to the hemicellulose fraction. At these conditions the conversion of cellulose to glucose is only 17% and the conv sion of hemicellulose to furfural by xylose

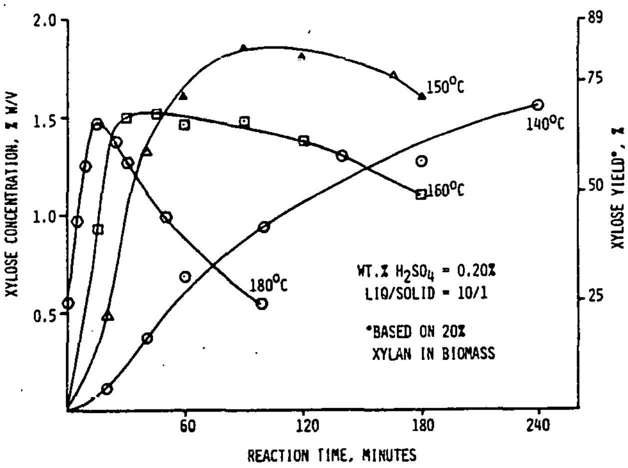

Fig. 1. Effect of temperature on xylose yields during hydrolysis of A.S.T.M. 4-8 mesh oak at 0.2 wt. % H_2SO_4.

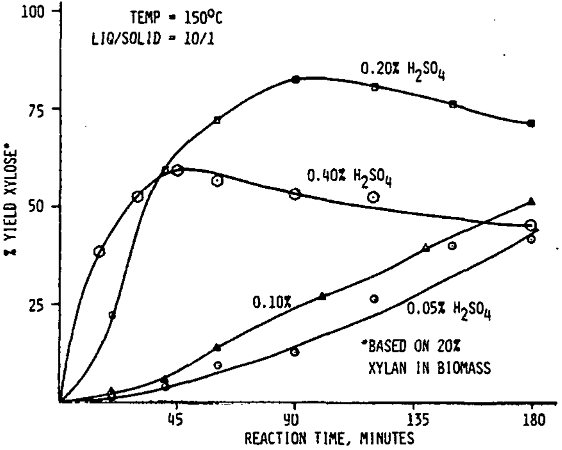

Fig. 2. Effect of H_2SO_4 catalyst concentration on xylose yields during hydrolysis of A.S.T.M. 4-8 mesh oak at 150°C.

TABLE I
HYDROLYSIS PRODUCTS AT 150°C, 0.2% H_2SO_4, AND 90 MIN. (MAXIMUM XYLOSE YIELD)

component	wt.% of soluble solids
xylose	50
glucose	26
polysaccharide	10
acetic acid	10
furfural	3
methanol	1
	100

ecomnosition is only 8%. The ratio of solid oak o te acid solution for the optimization of he rating conditions, temperature and % H_2SO_4, as maintained at 1 to 10. The yields were calulated based on Southern Red Oak hardwood conaining 20% xylan and 50% cellulose on a dry eight basis.

ne investigation of xylose yields and by-product ormation was carried out over a temperature range f 140°C - 180°C and an acid concentration range f 0.0% - 0.4% H_2SO_4. Table II gives the main nformation on the production of xylose for this nvestigation.

TABLE II
XYLOSE MAXIMUM YIELDS AND REACTION TIMES VS. TEMPERATURE AND % H_2SO_4

	% H_2SO_4			
emp.	0.05	0.10	0.20	0.40
40°C	*	*	*	*
50°C	*	*	83% 90 min	59% 90 min
50°C	*	55% 150 min	74% 45 min	71% 30 min
70°C	48% 90 min	64% 60 min	65% 30 min	69% 15 min
30°C	41% 70 min	55% 50 min	66% 15 min	74% 10 min

* Maximum yield not within 180 min

t 150°C and 0.20% H_2SO_4 the yield of xylose hanges very gradually with time in the region of aximum yield. This means that the exact resience time is not critical in a reactor design. everal reactor configurations appear practical rom the standpoint of chemical kinetics due to he relatively flat maximum in xylose yield.

y-products: The production of by-products is lso affected by the operating conditions. The ield of furfural is shown in Fig. 3 to be a strong unction of temperature and acid concentration. At 80°C and 0.40% H_2SO_4, furfural actually shows a aximum which gives evidence of furfural breakdown.

lucose occurs as a constituent of various crystalline celluloses, amorphous cellulose and side hains on hemicellulose. When exposed to acid, he amorphous cellulose and the hemicellulose side hains are hydrolyzed simultaneously with the ylan backbone of hemicellulose. However, at hese conditions most of the crystalline cellulose s unreacted [1].

aximum yield of glucose was realized at the most evere conditions investigated, 0.4% H_2SO_4 and 80°C. Figure 4 shows that glucose peaks after ylose and then begins to decline at a modest ate. Xylose degrades at a significantly faster ate than glucose [6]. The cellulose conversion o glucose is still quite low at 21%.

n sequence: The reaction sequence for the sis of hemicellulose has been postulated as a series of reactions. First, solid hemicellulose is converted into soluble xylose oligosaccharides. Next, the soluble xylose oligomers are hydrolyzed into monomeric xylose. Finally, xylose decomposes into furfural [1]. The xylose decomposition step has been reported to be first order in xylose and acid catalyst concentration [7]. Solubility limits of large polysaccharides and internal mass transfer resistance for the oak chip may affect the observed reaction rate.

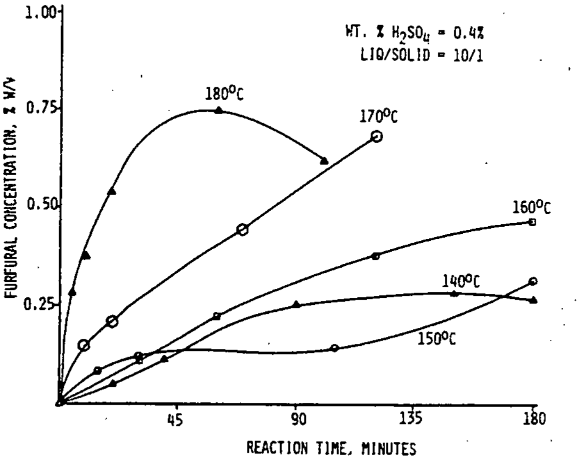

Fig. 3. Effect of temperature on the production of furfural at 0.4% H_2SO_4.

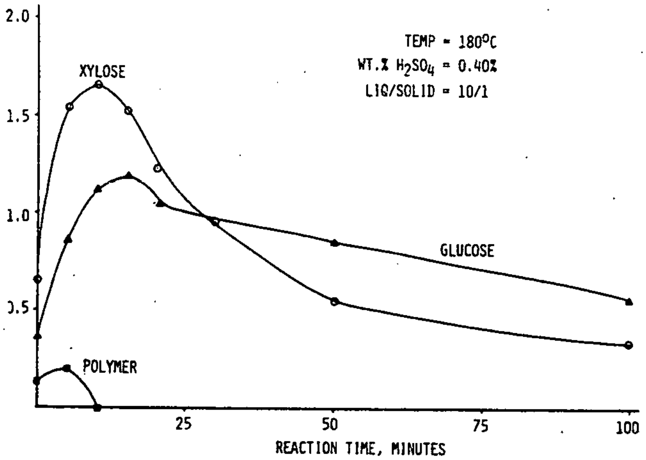

Fig. 4. Hydrolysis of A.S.T.M. 4-8 mesh oak at 180°C and 0.4% H_2SO_4. Sugar concentration (% W/V) vs. time (min.).

The carbohydrate analysis in this study was carried out using liquid chromatography at moderate pressures [8]. The product designated as polymer consists of oligosaccharides which have chain lengths of seven monomer units and above. It is probable that this polymer contains polysaccharides of glucose as well as oligomers of xylose due to the lack of separation between high molecular weight components. Analysis of individual components with chain lengths less than seven is possible with increasing separation as the chain length decreases.

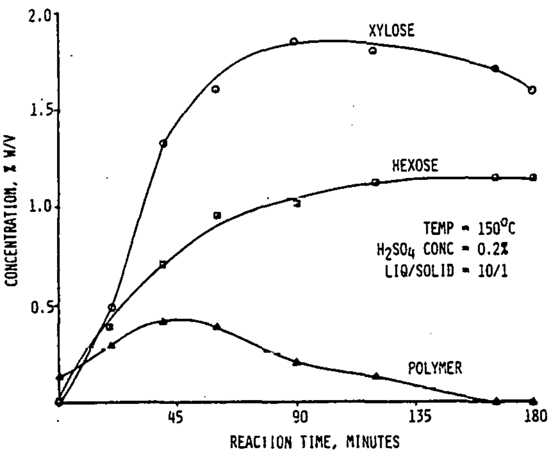

Fig. 5. Hydrolysis of A.S.T.M. 4-8 mesh oak at 150°C and 0.2% H_2SO_4.

Figure 6 illustrates the effect of temperature on the production of high molecular weight polymer, the first carbohydrate product to appear during hemicellulose hydrolysis. It is interesting to note that the temperature appears to affect only the reaction time of the maximum polymer concentration.

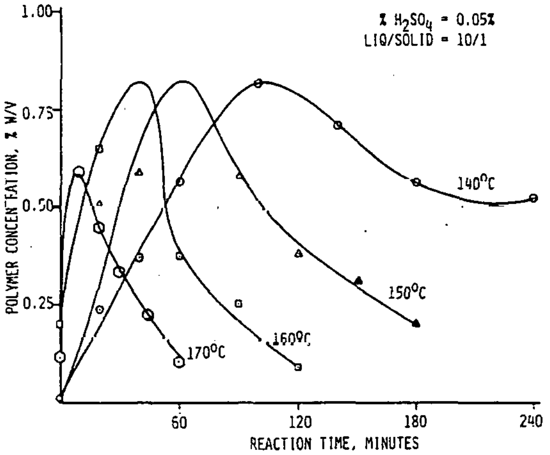

Fig. 6. Effect of temperature on polymer production at 0.05% H_2SO_4.

The amounts of individual xylose oligosaccharides with chain lengths less than seven vary considerably with the operating conditions. A maximum yield of 35% for xylobiose was observed at 150°C and 0.05% H_2SO_4. Figure 7 shows the effect of temperature on xylobiose production.

The concurrence of the decline of polymer and the maximum yield of xylose is illustrated in Fig. 5. This is in agreement with the reaction sequence where polymer preceeds monomeric xylose. The following list gives the components in order of their appearance in the series of reactions:

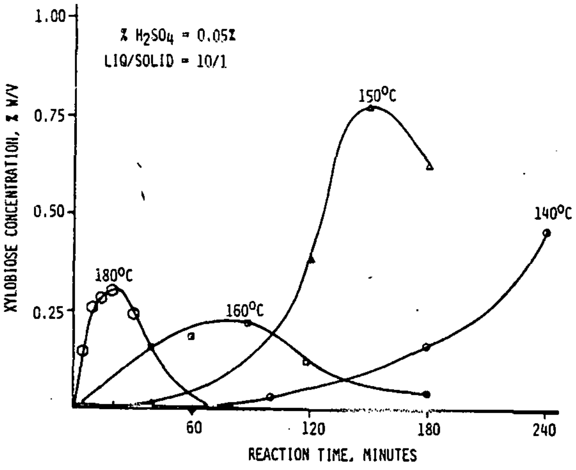

Fig. 7. Effect of temperature on xylobiose production at 0.05% H_2SO_4.

1. solid hemicellulose
2. soluble xylose oligosaccharides
3. xylobiose
4. xylose
5. furfural

In conclusion, the major results of this section are as follows. A maximum xylose yield of 83% is obtained at 150°C and 0.20% H_2SO_4. The hydrolysis of oak hardwood exhibits a high degree of selectivity with respect to hemicellulose. Concentrations of by-products in oak hydrolyzate at conditions for maximum xylose yield were relatively low. Also, substantial amounts of xylose oligosaccharides were observed at 0.05% H_2SO_4 and temperatures in the experimental range (140°C - 180°C).

Equipment Selection for Hemicellulose Hydrolysis and Solid-Liquid Separation

Based on the results of preliminary experiments and conversations with equipment suppliers, the following process has tentatively been selected for the production and separation of xylose from hemicellulose. As shown in Fig. 8, debarked wood is directed into chip preparation equipment which is responsible for chipping the wood into 4-8 cm mesh size. The chips are then metered into steam pretreating equipment in which the chips are both brought to process temperature and made more receptive to acid impregnation. During presteaming, the conditions are suitable to initiate hydrolysis of some of the hemicellulose present.

The chips are then conveyed to the hydrolysis reactor where the hemicellulose undergoes further reaction. Operating conditions are 140-160°C 0.02-0.06 percent sulfuric acid. It appears both continuous vertical and inclined hydrolysis

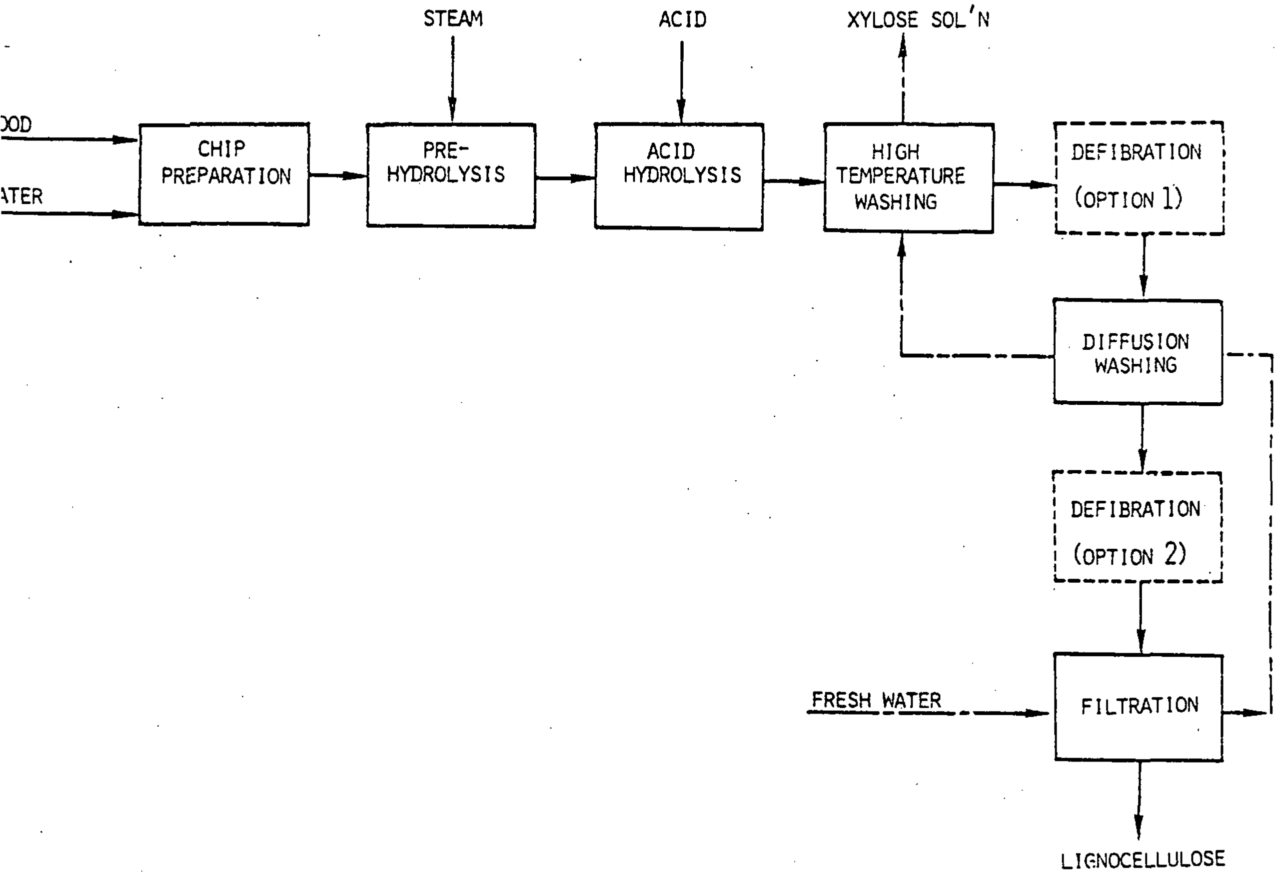

Fig. 8. Schematic for hemicellulose hydrolysis and xylose-lignocellulose separation.

actors will be suitable as well as the continuous draulic reactors in which the high temperature sh stages are an integral part of the reactor wer. Following a 20 minute hydrolysis period, e chips will consist of the original cellulose d lignin matrix (lignocellulose), the converted micellulose (now present in the chip interior as luble xylose) plus a small amount of unreacted micellulose.

e chip slurry is next passed to a high tempera- re wash which favors the diffusion of the luble xylose from the chip interior. Since the ip interior will contain acid, performing the rst wash at high temperature allows for further drolysis, thus increasing the effective residence ne of the reaction.

significant amount of recoverable material will ill be present within the chip after the first sh, therefore multistage countercurrent diffusion shing will provide for the transfer of this terial into the advancing wash. In order to omote high mass transfer rates, a size reduction the post-hydrolyzed chip is undertaken using :ion equipment. This size reduction will advantageous in the subsequent processing of cellulose to produce glucose. Whether this step is completed before or after diffusion washing will depend on the operating characteristics of the diffusion washing equipment. After washing, the solid lignocellulose is dewatered using conventional filtration and continuous pressing equipment.

Fermentation of Hemicellulose Hydrolyzates

The objective in this section of the program is to develop a fermentation process whereby the hemicellulose hydrolyzates are converted into fuel substances. The emphasis during this time span was placed on fermentation of xylose, the predominant component in the hydrolyzate. Specifically, the factors effecting the fermentation were identified and optimized. Improvements were then sought in fermentation from the standpoint of a chemical process.

Batch fermentation: As described in the previous section, an extensive microorganism screening study was made on a large number of bacteria and yeasts. On the basis of growth, product formation and xylose consumption data, a bacterium locally isolated from decaying wood (tentatively named AU-1-d3) was selected for fermentation studies.

Using this organism a series of batch fermentations were carried out. The medium contained xylose as a sole source, mineral salts (0.5% $(NH_4)_2HPO_4$ and others in minor quantities) and yeast extract (0.15%).

The initial xylose concentration was 10%. After 75 hours 1.9% xylose remained unfermented (Fig. 9). The product yield based on consumed sugar was 61%. Product distribution at this stage was as follows:

		% wt. of xylose consumed
Butanediol	73%	44.7
Ethanol	9.4%	5.7
Acetic Acid	8%	4.9
Acetoin	9%	5.5
		60.8

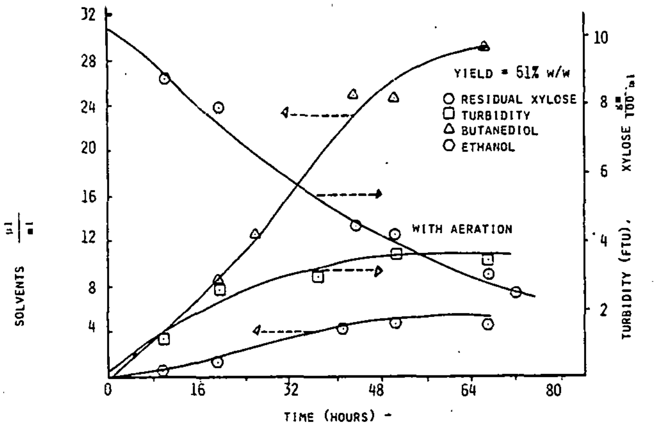

Fig. 9. Cell growth, solvent formation, and xylose consumption profiles with 10% initial xylose in stirred tank.

In addition to xylose, the hemicellulose hydrolyzate contains other hexoses and pentoses, namely glucose, mannose and arabinose. The ability of AU-1-d3 to metabolize these sugars was tested. A medium containing equal amounts of these sugars (1.25% each) was fermented. As shown in Fig. 10, all the sugars were consumed except arabinose. The products were the same as those from xylose. It is therefore expected that all the sugars from hemicellulose except arabinose will be converted in the fermentation process using AU-1-d3.

Modeling in product formation: A preliminary kinetic study was made with regard to product formation using Leudeking-Piret empirical model; $\frac{dP}{dt} = 2\frac{dX}{dt} +, \beta X$. The present fermentation data were in good agreement with the model (Fig. 11). Moreover the constants in the model were determined. From these constants, the relative magnitudes of growth-associated and non-growth associated terms were determined at a fixed time (t = 23 hr was chosen) as shown below.

Ethanol: $\alpha_E \frac{dx}{dt} = 0.018$

$\beta_E X = 0.208$

Butanediol: $\alpha_B \frac{dx}{dt} = 0.05$

$\beta_B X = 0.29$

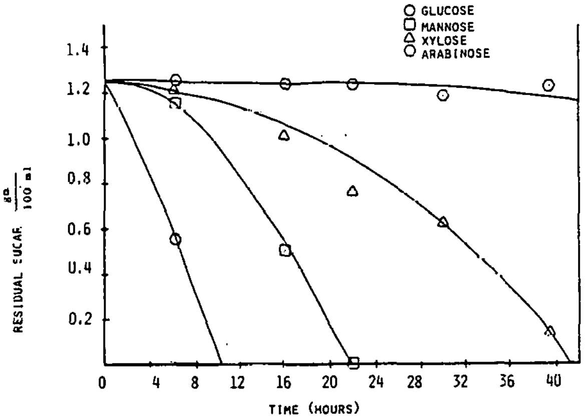

Fig. 10. Microbial sugar preference.

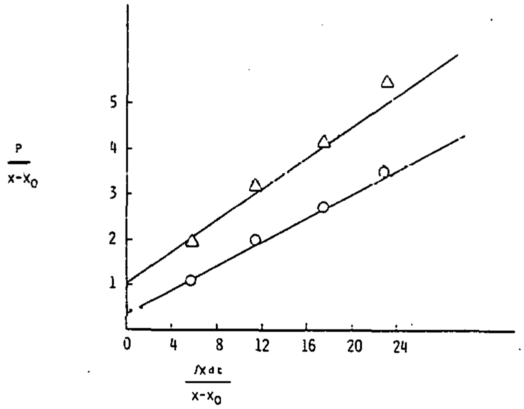

Fig. 11. Mathematical analysis of product formation in stirred tank fermentor.

Judging from the above data, it is apparent that product formation in this fermentation is essentially proportional to cell concentration. This in turn led us to investigate non-conventional fermentation schemes in which high bacterial concentrations are employed.

Column fermentation: In order to achieve high cel concentration during fermentation a method of affixing microorganisms on a solid support wa xp l. For this purpose, in cooperation with om and Griffith of Oak Ridge National Laboratory, a column fermenter was constructed using schig rings as support material. This column rmenter was attached to the main fermenter and e system was operated in a recirculation mode ig. 12).

ter initially carrying out runs with 5% W/V lose, the xylose content was raised to 10% W/V. e rate of fermentation was significantly imoved compared to that of the batch fermenter ig. 13). Of our particular interest was the ct that the maximum xylose consumption rate in e column was three times that of the batch rmenter. This also implies that the fermentaon time can be reduced by the same factor. The plus % W/V of xylose was totally consumed during e run. The concentrations of the solvents at eir maximum level were as follows:

	W/V %
Butanediol	3.08
Ethanol	1.15
Acetic Acid	0.46
Acetoin	0.2

An overall yield of 55% was obtained based on the amount of xylose consumed. The level of xylose fell to less than 0.6 gm%/100 ml within 51 hours. After 51 hours, the fermentation almost ceased. It would also be important to note that the final turbidity values obtained from process stream were only half of that obtained in batch fermentor. This, coupled with initial high fermentation rates shows that fermentation took place mainly by microorganisms attached to the column packing.

<u>Screening and Characterization of Wood Hydrolyzate Fermenting Bacteria</u>

Screening: Microorganisms were obtained from culture collections based on their reported potential of utilizing wood hydrolyzates or ability to produce fermentation products. Bacteria representing the following genera: Clostridium (10 cultures), Bacillus (11), Pseudomonas (1), Aeromonas (1), and 16 yeasts, representing 7 genera, were evaluated for their ability to initiate growth in an extract derived from acid hydrolyzed wood. Due to the acidity (pH 2.0) and low level of nitrogen (.005%) the hydrolyzate was

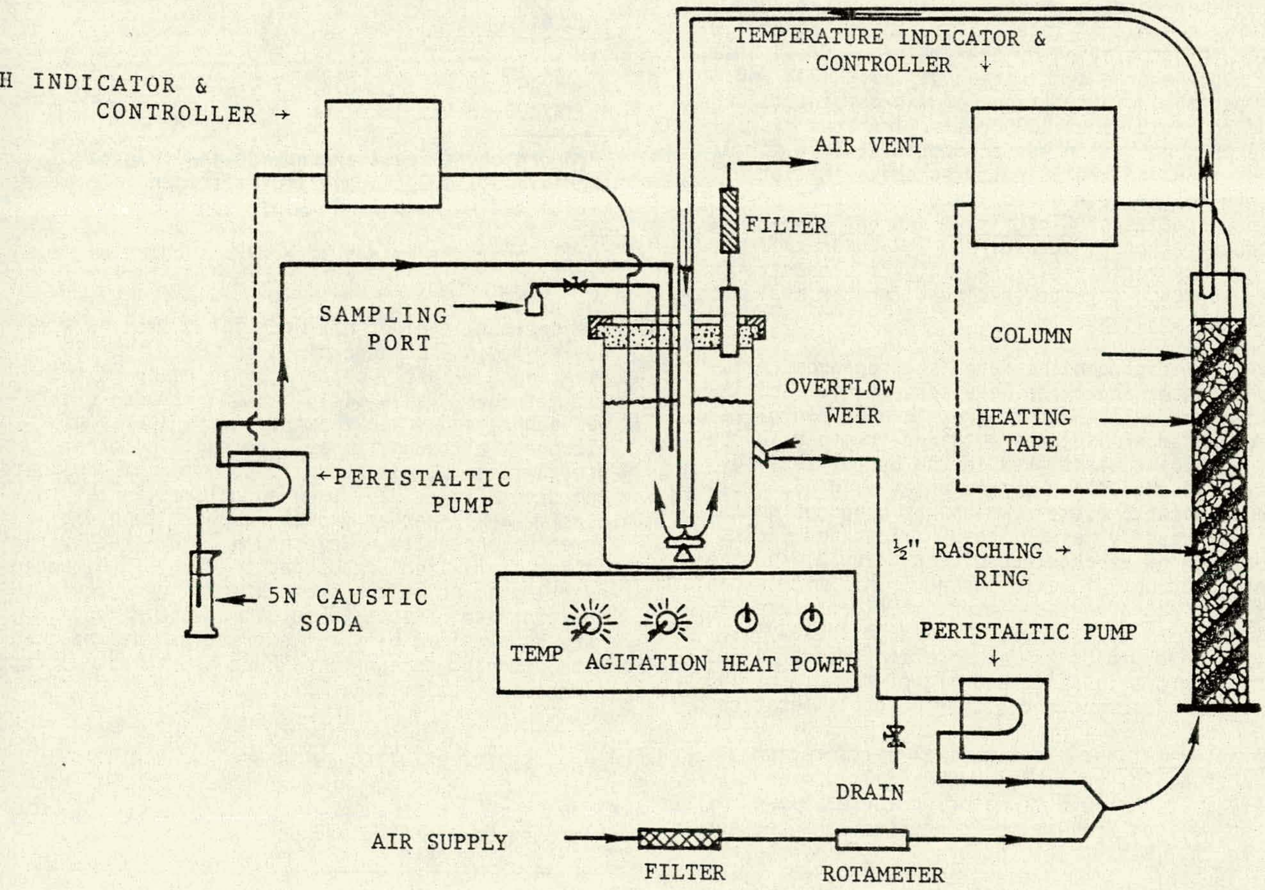

Fig. 12. Column fermentor set-up.

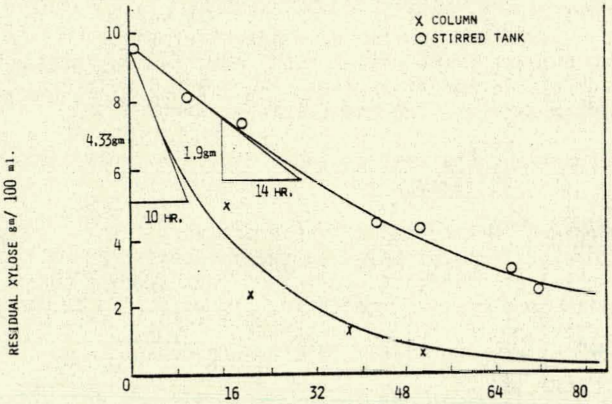

Fig. 13. Residual xylose vs. time in hours in column and stirred tank under similar starting conditions.

neutralized with NaOH or $CaCO_3$ and also supplemented with nitrogen. Adjustments in pH were calculated to give pH 4, 5, 6 and 7 after autoclaving the hydrolyzate. Nitrogen was added prior to autoclaving by the addition of .1% urea, .1% peptone or a combination of .1% KH_2PO_4 and .1% peptone. Inoculations of microorganisms were made to yield about 1000 cells/ml of the hydrolyzate and incubation was conducted at 32°C for 7 days. None of the 38 cultures above initiated growth in the acid hydrolyzate. Although some of the species of Clostridium and Bacillus produced ethanol, butanol and 2,3-butanediol in laboratory media their inability to propagate in acid wood hydrolyzate precluded further evaluation of these species.

Several microorganisms were isolated from cellulosic wastes and these were screened for ability to ferment xylose and to grow in wood hydrolyzate. Five bacterial isolates obtained from one source of cellulosic waste grew in the hydrolyzate adjusted to pH 6.0 and supplemented with .5% peptone. One of these isolates, which has been designated AU-1-d3 readily metabolizes D-xylose and yields about 60% of ethanol, acetic acid and 2,3-butanediol based on xylose fermented. This microorganism has distinct advantages over other organisms we evaluated in that it readily metabolizes xylose giving good yields of fermentation products, and also it grows in acid wood hydrolyzate requiring only pH adjustment and nitrogen supplementation.

Characterization: In a minimal salts medium ($NH_4H_2PO_4$, 1g; NaCl, 5g; $MgSO_4$-$7H_2O$, .2g and KH_2PO_4, 1g; in 1000 ml water) culture AU-1-d3 grows well at D(+) xylose concentrations between .1 to 2%. Best growth occurs with 2% xylose and levels above 3% suppress growth. However, after the maximum stationary growth phase is reached the organism readily metabolizes 10% xylose to ethanol and 2,3-butanediol. The organism is capable of utilizing a variety of carbohydrates. Mannose, sucrose, lactose, xylose, maltose, glucose, arabinose and fructose all support good growth. Culture AU-1-d3 also tolerates a wide range of pH for growth. Good growth occurs between pH 5.0 to 7.0 with the optimum range between pH 5.5 to 6.5. The organism has a rapid growth rate. With an inoculation rate yielding about 100 cells/ml of minimal salts medium initially, the maximum stationary growth phase is reached in 10.4 hr at 37°C. The optimum growth temperature for culture AU-1-d3 is 37°C; however, there is little difference in growth rate between 32 and 37°C. No growth occurs at 45°C.

TABLE III
EFFECT OF NITROGEN LEVEL AND SOURCE
ON GROWTH OF AU-1-d3

0.1% Nitrogen	Cells/ml
100/0[A]	8.1×10^8 [B]
60/40	7.8×10^8
50/50	6.9×10^8
20/80	6.8×10^8
10/90	6.3×10^8
0.5% Nitrogen	
100/0	1.9×10^9
1.0% Nitrogen	
10/90	$< 10^5$
1/99	$< 10^6$
0/100	$< 10^5$

[A] Proportion of yeast extract to $NH_4H_2PO_4$ to achieve 0.1%, 0.5%, and 1.0% nitrogen in minimal salts medium with 1% D-xylose (pH 6.0).

[B] Cell count determined after 72H incubation at 32°C.

The optimum level of nitrogen for growth of AU-1-d3 is within the range of .1 to .5%, with .5% yielding a slightly better growth response than .1% nitrogen (Table III). At 1% nitrogen growth was suppressed. Yeast extract as a source of nitrogen, although it may have supplied other growth factors, was superior to ammonium phosphate and urea (Table IV). Urea autoclaved in the substrate gave a better growth response than did ammonium phosphate. However, a later study using filter sterilized urea added to give .1% nitrogen after autoclaving the substrate was very toxic to the organism. About 40 to 50% of the nitrogen need as supplied by yeast extract can be replaced with ammonium phosphate without appreciable affect on growth (Table III).

TABLE IV
EFFECT OF NITROGEN SOURCE ON GROWTH
OF AU-1-d3*

Nitrogen Source (Equivalent at 0.1% N	OD (610 nm)
Yeast extract	.509
Phenol Red Broth Base	.161
$NH_4H_2PO_4$	.076
Urea	.268

*Growth in minimal salts medium at pH 6.0 with 1%

-xylose. Optical density determined after 72H ncubation at 32C.

u' AU-1-d3 grows in acid and alkaline wood y... zates adjusted to pH 6.0 to 6.5 and supplemented with .1% nitrogen, as ammonium phosphate, nd 1% D-xylose (Table V). Growth was better in he undiluted alkaline hydrolyzate than in the acid ydrolyzate with increased growth occurring in both ubstrates upon dilution with minimal salts medium. t about 40 to 50% of acid hydrolyzate or alkaline ydrolyzate in minimal salts medium, growth of ulture AU-1-d3 was similar in both substrates. lkaline hydrolyzate without nitrogen and xylose upplementation also supported good growth of the ulture.

urther studies are underway to optimize growth and ermentation yields of culture AU-1-d3.

TABLE V
GROWTH OF ISOLATE AU-1-d3 IN
ACID AND ALKALINE HYDROLYZATES

ydrolyzate %	Type Wood Hydrolyzate Acid^A	Alkaline^A	Alkaline^B
100	3.0×10^4	6.8×10^8	6.9×10^8
80	1.2×10^6	7.2×10^8	8.5×10^8
60	6.6×10^7	7.3×10^8	9.6×10^8
50	8.2×10^8	1.1×10^9	1.3×10^9
40	1.1×10^9	1.5×10^9	
20	1.3×10^9	3.3×10^9	
10	5.4×10^9	6.2×10^9	
0	8.5×10^9	8.9×10^9	

).1% nitrogen ($NH_4H_2PO_4$) and 1% D-xylose added. ydrolyzates diluted in minimal salts medium .1% N and 1% xylose) pH 6.0.

o xylose or nitrogen added. Buffered at pH 6.5 ith Na_2HPO_4 (1.79%) and K_2HPO_4 (1.17%), hydrolyate diluted with sterile water. All samples incubated 72H at 32C.

JTURE PLANS

e selection of hemicellulose hydrolysis conditions ill be investigated to optimize xylose yield in rms of process variables (biomass type, chip size, etreatment conditions, acid concentration, rention time, temperature, and liquid to solid tio). This work thrust will directly affect the lection of the large-scale hydrolysis reactor d be performed in the experimental high solids actor.

eas for future study in the area of hemicellulose drolysis include investigation of the decomposion kinetics of xylose at the operating condition nges of this investigation, the kinetics of drolysis of various oligosaccharides, and the fect of chip size on observed reaction rates, lectivity and yields to determine the significnce of internal mass transfer resistance. The fect of the solid to liquid ratio in the reactor is also extremely important since increasing this ratio is a preference to increasing the number of recycles of the hydrolyzate in a continuous operation for production of a fermentation substrate medium with an adequately high concentration of xylose.

Further diffusion experiments will be performed to determine the best washing conditions (including chip size, temperature, length and nature of wash cycle). Filtration studies will be conducted to characterize the solid material and investigate the economics of the proposed solid-liquid separation schemes.

The future work on batch fermentation will focus closely on the parameters of fermentation. Attempts will be made to direct the pathways toward a single major product (either ethanol or butanediol) using a specific inhibitor. Other fermentation schemes will be investigated which produce high yield of potential liquid fuels (butanol, acetone, etc) for comparison with AU-1-d3 xylose conversion to butanediol and ethanol.

The results obtained with column fermentation have been very encouraging. Attempts will be made to further increase the rates of fermentation and substrate concentration by continuously acclimating microorganisms to high product concentrations. Adapting the microorganism to hemicellulose hydrolyzate by gradually increasing the proportion of hydrolyzate in the column process (feed) stream needs to be carried out. The effect of reducing the nitrogen medium especially yeast extract will be studied in column fermentation. Under the present set up, mass transfer effects for oxygen, substrate, and product may be playing significant role in fermentation. These effects need to be investigated.

Studies are planned to further optimize the fermentation potential of the organism. The screening program will continue to search for other organisms or strains of culture AU-1-d3 which are better suited for growth and fermentation of acid wood hydrolyzate and which also tolerate high levels of products in the substrate. Organisms which are tolerant of products resulting from acid hydrolysis of wood and also tolerate high levels of fermentation products will be selected for further study. Modification of the acid hydrolyzate process or supplementation of the hydrolyzate will be evaluated to further optimize the yield of fermentation products.

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IMPROVEMENT OF YIELDS AND RATES USING IMMOBILIZED ENZYMES DURING THE HYDROLYSIS OF CELLULOSE TO GLUCOSE

H. E. Klei, D. W. Sundstrom, R. W. Coughlin,
K. Ziolkowski, and G. Biederman

Department of Chemical Engineering
University of Connecticut
Storrs, Connecticut

STRACT

e objective of this program is to show that the nversion of cellulose to glucose can be increased gnificantly by enzymatically removing the inhibi- ry cellobiose from the reaction system using im- ilized β-glucosidase (β-G). The enzyme β-G was oduced and isolated from *Aspergillis phoenicis* d was immobilized on Corning controlled-pore alu- na by adsorption. Activity retentions upon im- ilization have been above 40%. When this en- natic catalyst was used in a fluidized bed with llobiose as the only substrate, a 10% loss of tivity was observed during a 500 hour period. other experiments cellulose was hydrolyzed in batch reactors operated side-by-side, with one actor containing cellulase, immobilized β-G and llulose and the other reactor containing no im- ilized β-G but otherwise identical. After 30 rs the reactor containing the immobilized β-G d 100% more glucose, indicating that the cataly- removal of the cellobiose had a significant fect upon the production of glucose.

SCRIPTION OF TASK

mass is produced continuously by photosynthetic rage of solar energy in green plants and repre- ts a significant potential source of renewable rgy for the production of liquid and gaseous ls. About 50% of this solar produced biomass is the form of cellulose, which must be broken down o smaller molecules before it is conveniently ble as a liquid fuel. Most routes for the con- sion of the cellulose involve the hydrolytic de- ymerization of the cellulose to glucose and lobiose with minor amounts of other sugars being duced. Enzymatic hydrolysis using a mixture of ymes has received much attention over the older d hydrolysis because of the fewer side products erated. The enzyme mixture is collectively led cellulase and contains three major enzymes: loglucanase and cellobiohydrolase, which cleave β-1,4 linkages and produce cellobiose, and β- cosidase which hydrolyzes the cellobiose mole- e to form two glucose molecules. However, the ymatic reaction slows down appreciably when the centrations of cellobiose and glucose builds up. enzymatic conversion of the cellulose is a ti-step process, in which a most important reac- talyzed by β-glucosidase (β-G), is the con- f the inhibitory cellobiose to glucose. hough many microorganisms produce β-G as a component of the cellulase, generally it is in insufficient amounts to prevent the accumulation of inhibitory levels of cellobiose in the reactor. The cellobiose is considered to be an uneconomical and unwanted byproduct which, if it were hydrolyzed, could significantly increase the yield of glucose.

OBJECTIVES

Since the levels of β-G in the cellulase enzyme mixtures are usually insufficient to prevent the accumulation of cellobiose, our research program is investigating the use of additional β-G immobilized on a suitable support to convert the unwanted cellobiose to desirable glucose. By immobilizing the β-G and thereby allowing its reuse, the cost of glucose produced by the enzymatic hydrolysis process should be reduced.

The immobilized β-G can be used in a variety of reactor configurations. It may be placed into the same reactor as that containing the crude cellulase mixture of enzymes, or it can be operated separately in its own reactor and connected to the cellulase system by a recycle loop. This research program is directed towards immobilization of β-G and its evaluation in cellulose reactor systems. Its goal are to:

1. Immobilize the β-G on supports suitable for commercial reactors.
2. Study the dependence of immobilization yields on the purity of the β-G and the method of immobilization
3. Operate a cellulase reactor containing immobilized β-G to show that yields of glucose are increased by the β-G catalyzed removal of the inhibitory cellobiose.
4. Develop at UConn the capability of producing the β-G enzyme required for the research from the organism *A. phoenicis*.

EXPERIMENTAL APPROACH

1. Preparation of the microbial β-G from A. phoenicis. Samples of *A. phoenicis* from the QM collection at the University of Massachusetts were obtained and grown in a 14 liter fermentor in a starch medium recommended by Dr. Mandels and Dr. Sternberg of the U.S. Army Natick Laboratories.

The culture was grown at 28-30°C for 14 days during which time the pH was adjusted to prevent it from dropping below 4.0. If the pH dropped below 4.0 the yield of β-G decreased.

After 10-14 days the broth was concentrated by ultrafiltration to 1/10 of the original volume using a 30,000 MW membrane. At this point the solution was either freeze dried or further purified by acetone precipitation to give a dry powder.

2. Effect of purification on immobilization yields. The acetone-precipitated material was compared with the crude, freeze-dried material to see whether the enzyme loading (EU/gm of support) and the immobilization yield (EU actively adsorbed/EU removed from contacting solution) are significantly increased by the purification. The goal was to use as impure material as practical to save on purification costs.

3. Immobilize the β-G on suitable supports. The initial screening included selecting a suitable support and investigating both the extent of loading possible on the supports and the effect on the activity of crosslinking with a 0.25% glutaraldehyde solution.

One important performance criterion is the life of the immobilized enzyme under reaction conditions. After a suitable support had been chosen based upon loading and activity retention tests, the corresponding immobilized β-G was tested in a fluidized-bed reactor for several days at 50°C using the apparatus shown in Figure 1. A substrate solution of cellobiose was pumped through the fluidized-bed reactor at a flow rate sufficiently high for fluidization, but sufficiently low for a residence time to permit a high conversion. Fresh cellobiose solution was continuously fed to each recycle reactor system at the same rate as the product solution was withdrawn, therby allowing the system to attain continuous steady-state operation. The glucose concentration in each product stream was periodically measured to determine the loss of enzyme activity with time.

4. Operation of a cellulase reactor with immobilized β-G. The main objective of this program is to test the immobilized β-G in a reaction system containing cellulase and using cellulose as a feed material. Two batch reactors of 1.0 liters were run in parallel, one without the immobilized β-G and the other containing it. The initial cellulose concentration was 55 gm/ℓ. The cellulase enzyme concentration was at 225 EU/ℓ and the temperature was held at 50°C during the reaction. One reactor also contained 5.0 gms of immobilized enzyme containing 40 EU of β-G per gm of alumina support. Samples were periodically withdrawn from the reactor and assayed for cellobiose and glucose concentrations using a liquid chromatograph and a glucose analyzer.

Solka-floc was a convenient pure cellulosic material for use in these first experiments. More realistic materials will be employed in future experiments.

RESULTS

β-G production and purification. Broth concentrations of β-G after about 10 days of fermentation were found to be > 3.0 EU/mℓ, provided the pH during the fermentation did not drop below 4.0 After filtration to remove the cells and ultrafiltration to concentrate the enzyme, the solution was freeze-dried to yield a crude powder with an activity of 0.9 EU/mg. This material was further purified using acetone precipitation to yield a powder having an activity of 16 EU/mg. Both enzyme preparations were immobilized and evaluated.

Immobilization. The supports which were tested together with the respective fractions of activity retention are given in Table I. Activity retention is defined as the measured activity of the solid/(original activity of the contacting solution less the residual activity of contacting solution). Thus, for example, 42% of the activity lost during adsorption from the contacting solution was found on the support No. 1. The TiO$_2$ support (#4 and #7) was found to break up easily under agitation and therefore was not used further. The SiO$_2$ and the DP-1 ion exchange resin showed no immobilized activity as a result of our simple preliminary procedure. One possible reason for this result may reside in the relationship of immobilization pH to the isoelectric point of the enzyme. The best support tested was the alumina (Type 7881-A) from Corning. This material did not break up under fluidization and still gave a high activity recovery on immobilization.

Table I

Support Screening

Supports Tested	Activity Retention
1. Alumina Type 7881-A 715A°	42%
2. Alumina Type 7851-A 245A°	21%
3. SiO$_2$ Type 7391-SX 660A° + Glutaraldehyde	0
4. TiO$_2$ - from Corning	43%
5. Alumina Type 7881-A coated w/TiO$_2$	37%
6. Alumina Type 7881-A + Glutaraldehyde (.25% soln.)	30%
7. TiO$_2$ - from Corning + Glutaraldehyde (.25% soln.)	44%
8. Ion exchange resin DP-1	0
9. Alumina Type 7881-A coated w/TiO$_2$ + Glutaraldehyde	28%

n(e alumina (support 1) showed good initial sults, it was tested further to explore increased zyme loading and associated retention. In the mobilization procedure, 5 gms of alumina were alys contacted with 50 mℓ of enzyme solution conining different amounts of powdered enzyme. The sults for the crude freeze-dried powder are given Figure 2, and for the purified powder are given Figure 3. The effect of glutaraldehyde cross ıking was also examined in both cases. For the eeze-dried powder, the EU adsorbed upon the alumi- was maximum at a solution concentration of about 5 EU/50 mℓ. Cross linking with glutaraldehyde peared to have little effect on either the amount EU adsorbed for the extent of activity retention ring immobilization. The activity retention reined relatively constant at about 50%.

en purified enzyme powder was used, the EU adrbed upon the support continuously increased as e solution concentration increased. Final immolized enzyme levels were about 375 EU/5 gm of umina, twice the maximum level reached using the ıde freeze-dried powder. As the amount of enme adsorbed upon the alumina increased, the actiy retention decreased, although the range still nained between 40 and 50%. Glutaraldehyde cross ıking reduced both the mass loading and the actiy retention of the purified enzyme. The purified zyme offers sufficiently higher support loadings make it worth considering for the final reactor sign.

e decrease with time of the enzyme activity dur- hydrolysis of cellobiose to glucose is shown in gures 4 for non-crosslinked enzyme and in Figure for immobilized enzyme that had been crosslinked ing glutaraldehyde. Glutaraldehyde does not apar to change the active life of the supported enme. In both cases about 10% of the enzyme actiy had disappeared after 500 hours of continuous eration; in both these cases the crude freezeied enzyme was the source material. This immolized enzyme is quite stable and offers a practil catalyst for further studies. The active life the purified material will also be investigated future work.

ıobilized β-G and cellulase reactor. The results om operating the batch cellulase reactor both h and without the addition of immobilized β-G lition are shown in Figure 6. For the case of no addition the cellobiose and glucose concentraons continuously increased, such that at 30 hours, ir concentrations were approximately equal. en immobilized β-G was present, however, the gluse concentration observed was approximately 100% eater than in the former case; moreover the resial cellobiose concentration was significantly er and remained constant at about 0.3 gm/ℓ. It apparent from these experiments that the immoized β-G caused the hydrolysis of considerable lobiose to glucose. Owing to crystallinity not the cellulose content of Solka-floc is accessi- he cellulase enzymes. Nevertheless, the of immobilized β-G to the cellulase mix- dramatically increases the yield of glucose at the expense of the undesirable cellobiose. Other practical cellulosic materials with higher accessibility, such as pretreated corn stover and poplar wood, will be tried in the future.

FUTURE PLANS

The following is a description of the experimental work we plan to undertake during the second year of this research project.

Feedstocks (Substrates). Initial experiments have been aimed at learning technical feasibility, and at developing new methods and techniques to immobilize the enzyme for several reactor configurations. Most of this initial work was done with rather pure and refined forms of cellulose for the sake of reproducibility and in order to have a basis for comparing the results with related work of others. Plans are underway to obtain samples of other less refined cellulosic feedstocks and it is expected that it will be possible to use some of these materials in some initial trial experiments towards the end of the first year of this project. During the proposed second year we plan to concentrate on the use of such feedstocks as:

A. Poplar wood, milled and acid-treated by the Dartmouth group.
B. Pine prepared and pretreated as in A. when it becomes available.
C. Corn stover milled, with and without acid pretreatment, depending on availability from the Dartmouth group.
D. Wood pretreated by exploding it through a die by a new Canadian process.
E. Various cellulose samples pretreated by the Purdue process insofar as they become available from Purdue.

Immobilization of β-glucosidase (β-G). Our work during the first months of this project has resulted in a highly active immobilized form of the β-G enzyme obtained from Aspergillis phoenicis; moreover this immobilized preparation has shown a remarkable half-life during on-stream tests while processing a continuous flow of cellobiose solution. This immobilized enzyme was prepared by adsorption on porous alumina particles. Although this accomplishment represents a significant step forward, we do not have reason to believe we have achieved the optimum activity or half life. We plan to continue to study variations in support and method of immobilization in an effort to improve further the activity and half life of immobilized β-G as well as possibly lower the cost of its preparation. We propose to pursue this goal during the next year as time and resources permit.

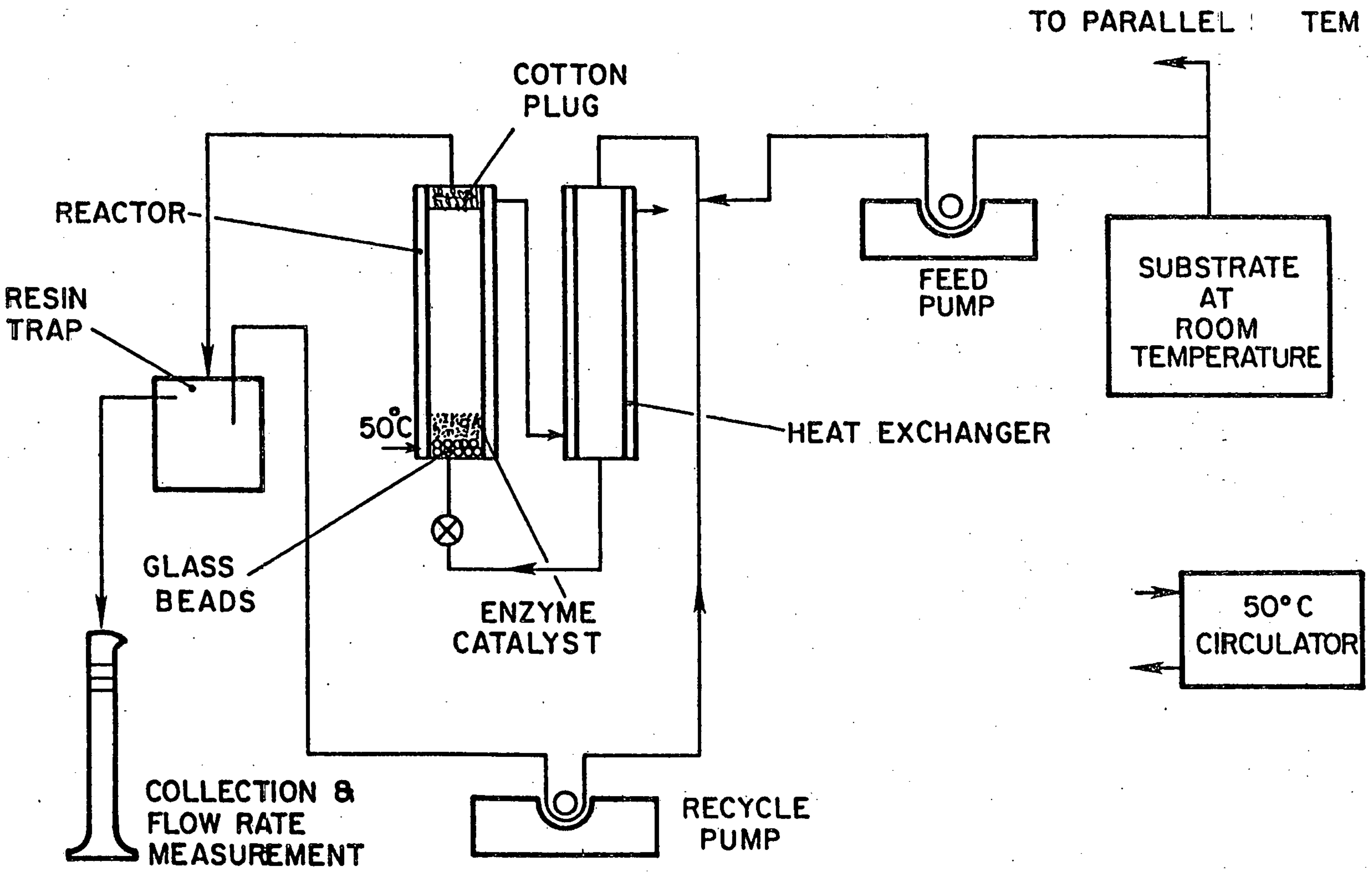
We also intend to study various reactor configurations during the next year that will require immobilized β-G enzyme in the form of membranes rather than in the form of particles. This will entail developing methods for entrapment and immobilization of β-G within various membranes. Such membranes must be carefully chosen and designed to provide a favorable environment for the enzyme (thereby providing stability and long lifetime) as well as to possess suitable porosity to insure rapid transport of the cellobiose substrate and the glucose product.

ACKNOWLEDGMENTS

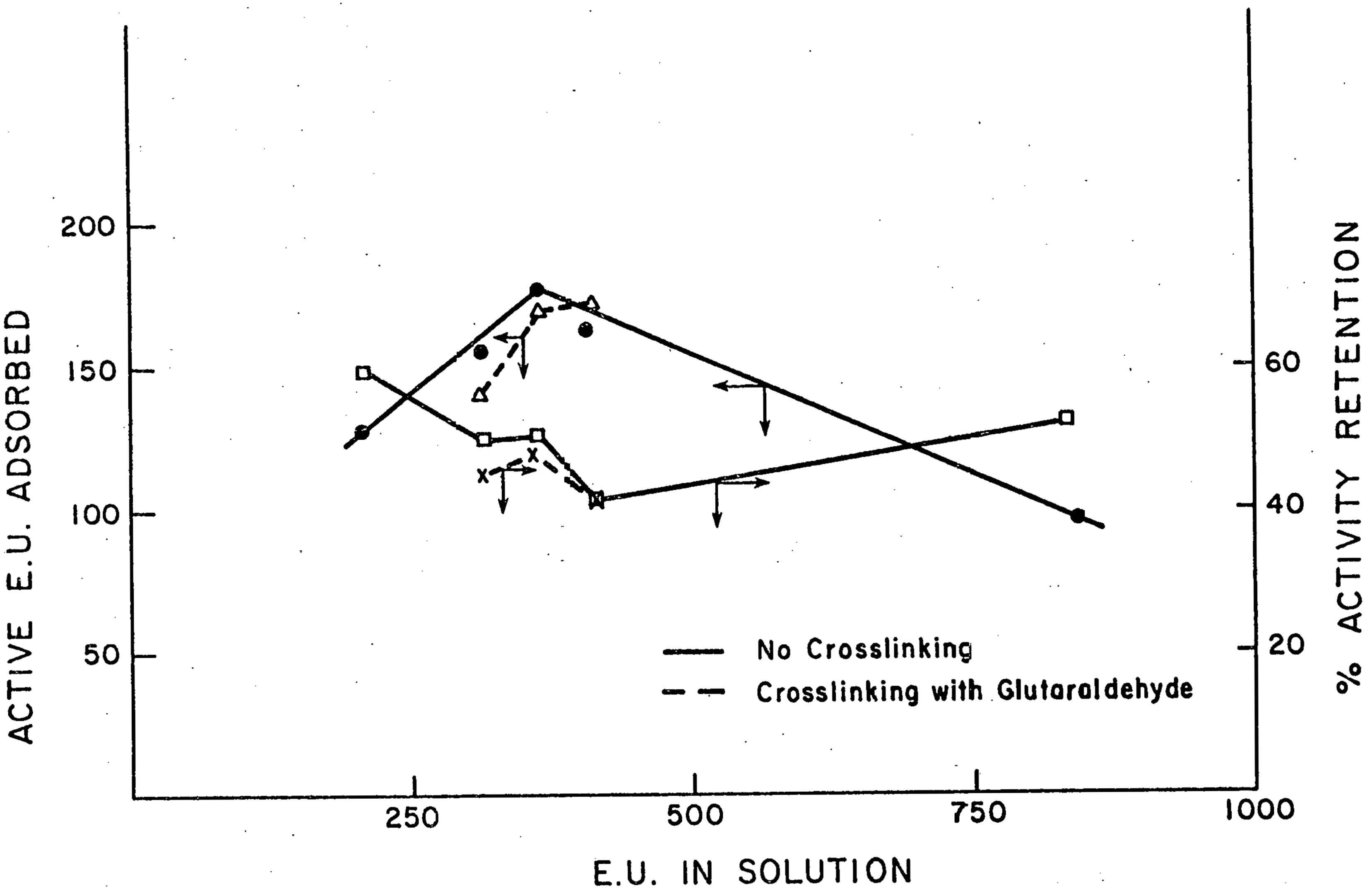
We wish to thank Dr. Mandels and Dr. Sternberg at the U.S. Army Natick Laboratories for supplying us with initial amounts of enzyme and technical assitance. Dr. Bungay provided helpful suggestions and encouragement. We are indepted to the U.S. Department of Energy, Fuels From Biomass Program, for financial support.

FIGURE TITLES

Figure 1.	Fluidized Bed Apparatus for Evaluating Immobilized Enzyme Half-Life.
Figure 2.	Solid Loading and Activity Retention of 5 gms of Alumina when Exposed to 50mℓ of Buffer Containing Various Amounts of Crude Enzyme.
Figure 3.	Solid Loading and Activity Retention of 5 gms of Alumina when Exposed to 50 mℓ of Buffer Containing Various Amounts of Purified Enzyme.
Figure 4.	Immobilized Activity versus Reaction Time with Cellobiose for No Crosslinking.
Figure 5.	Immobilized Activity versus Reaction Time with Cellobiose for Crosslinking with 0.25% Glutaraldehyde.
Figure 6.	Glucose and Cellobiose Concentration in Batch Reactors with/without Addition of Immobilized β-Glucosidase.



FLUIDIZED BED WITH RECYCLE


ACTIVE E.U. ADSORBED
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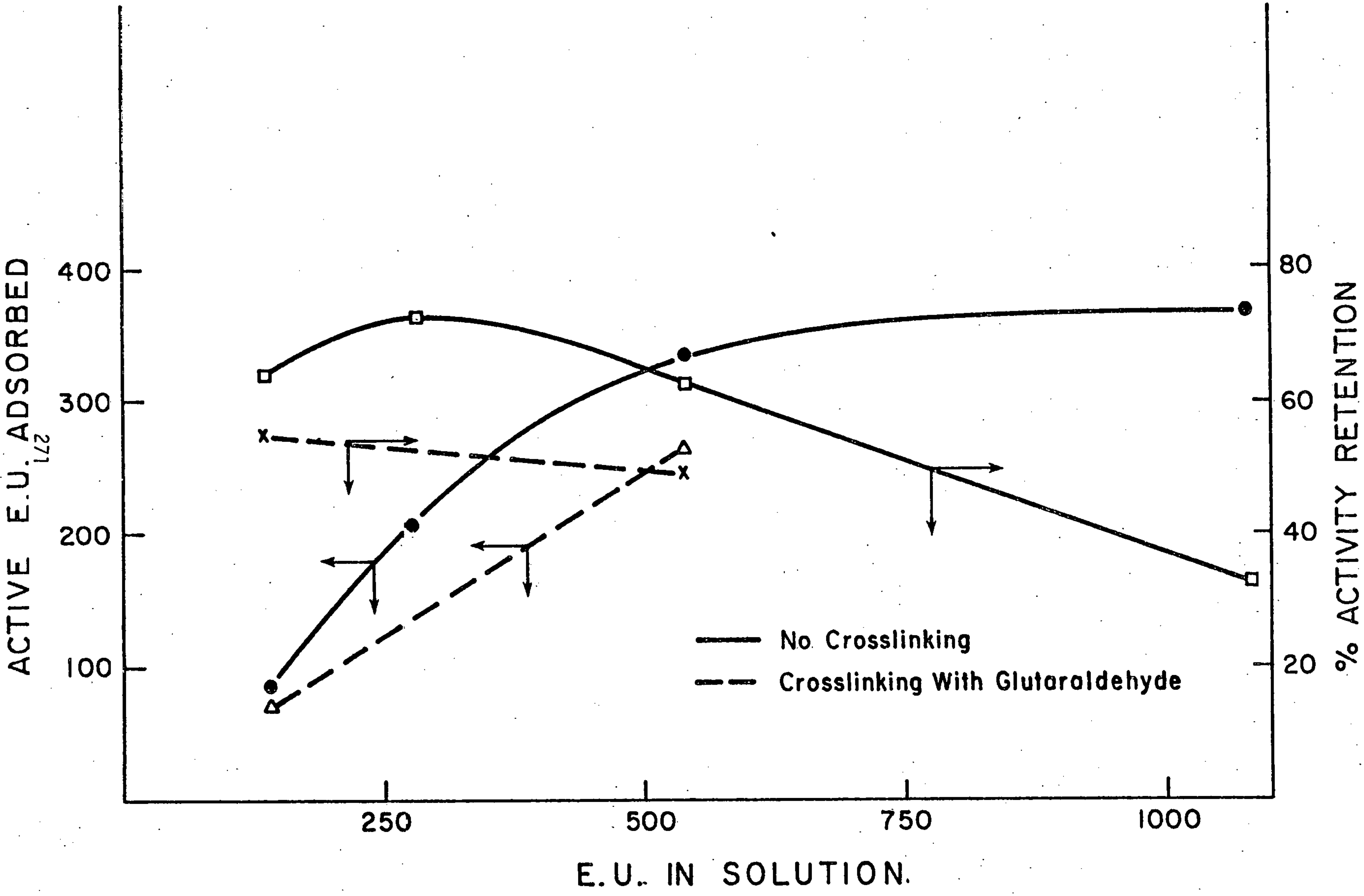
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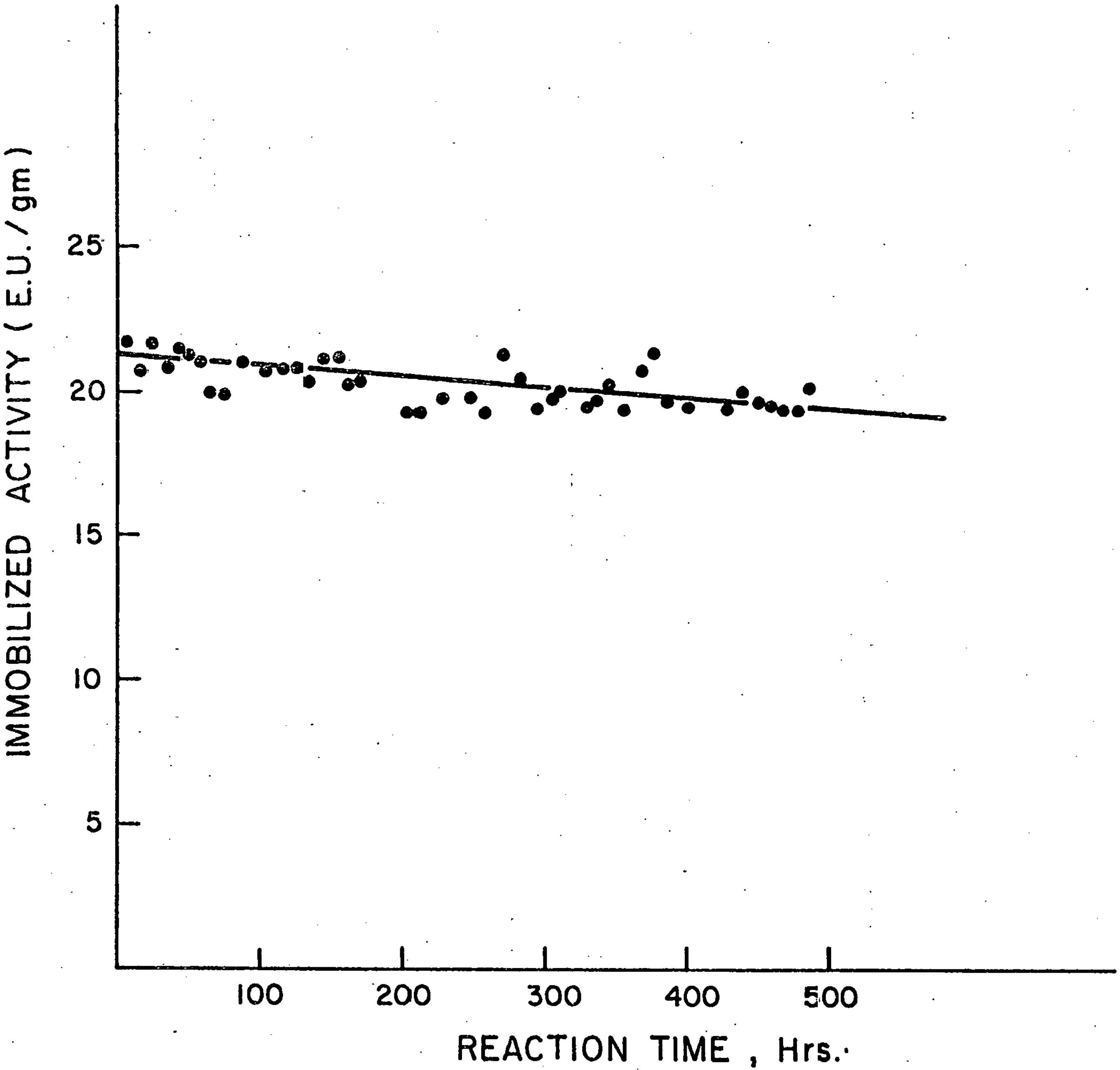
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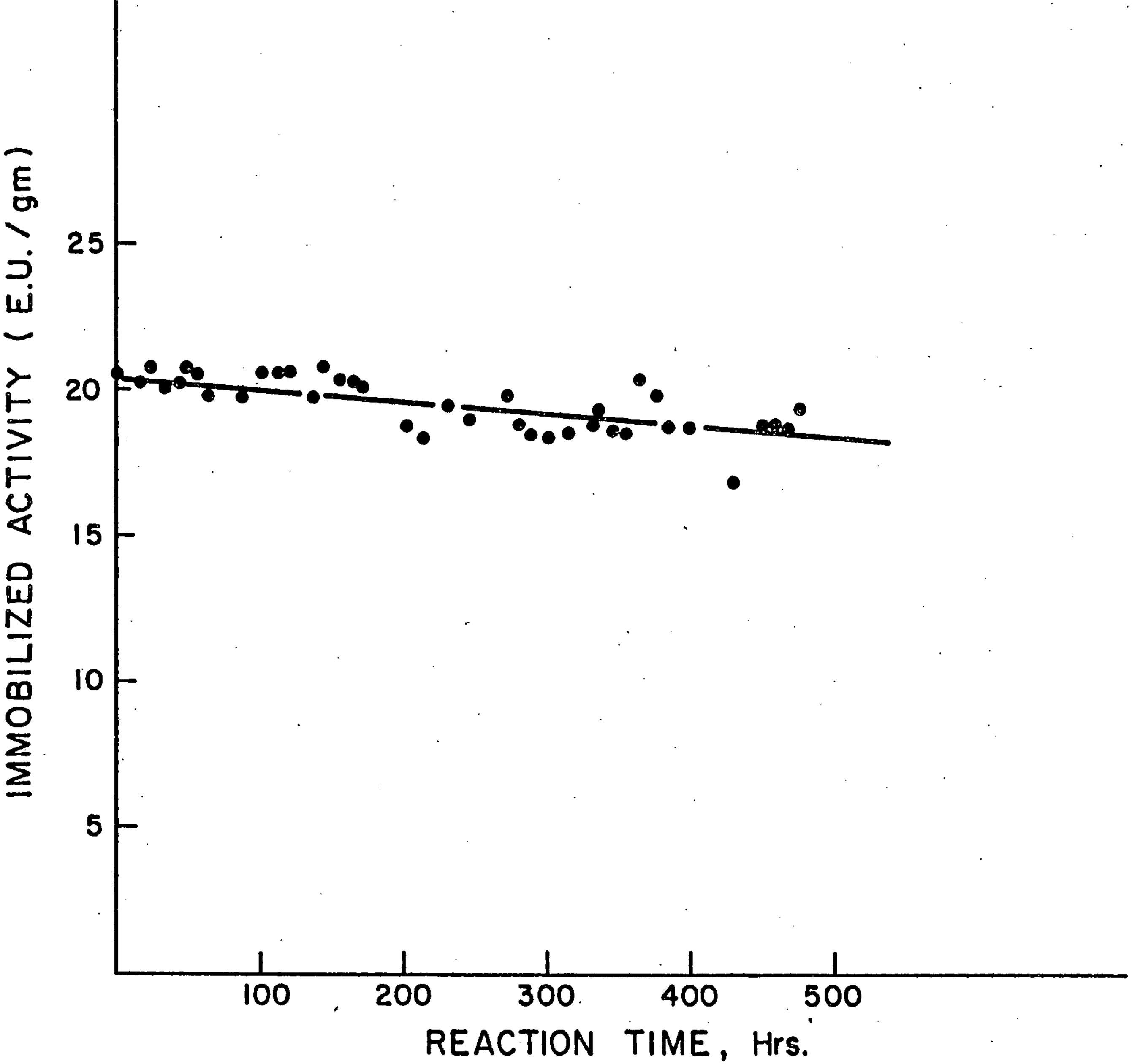
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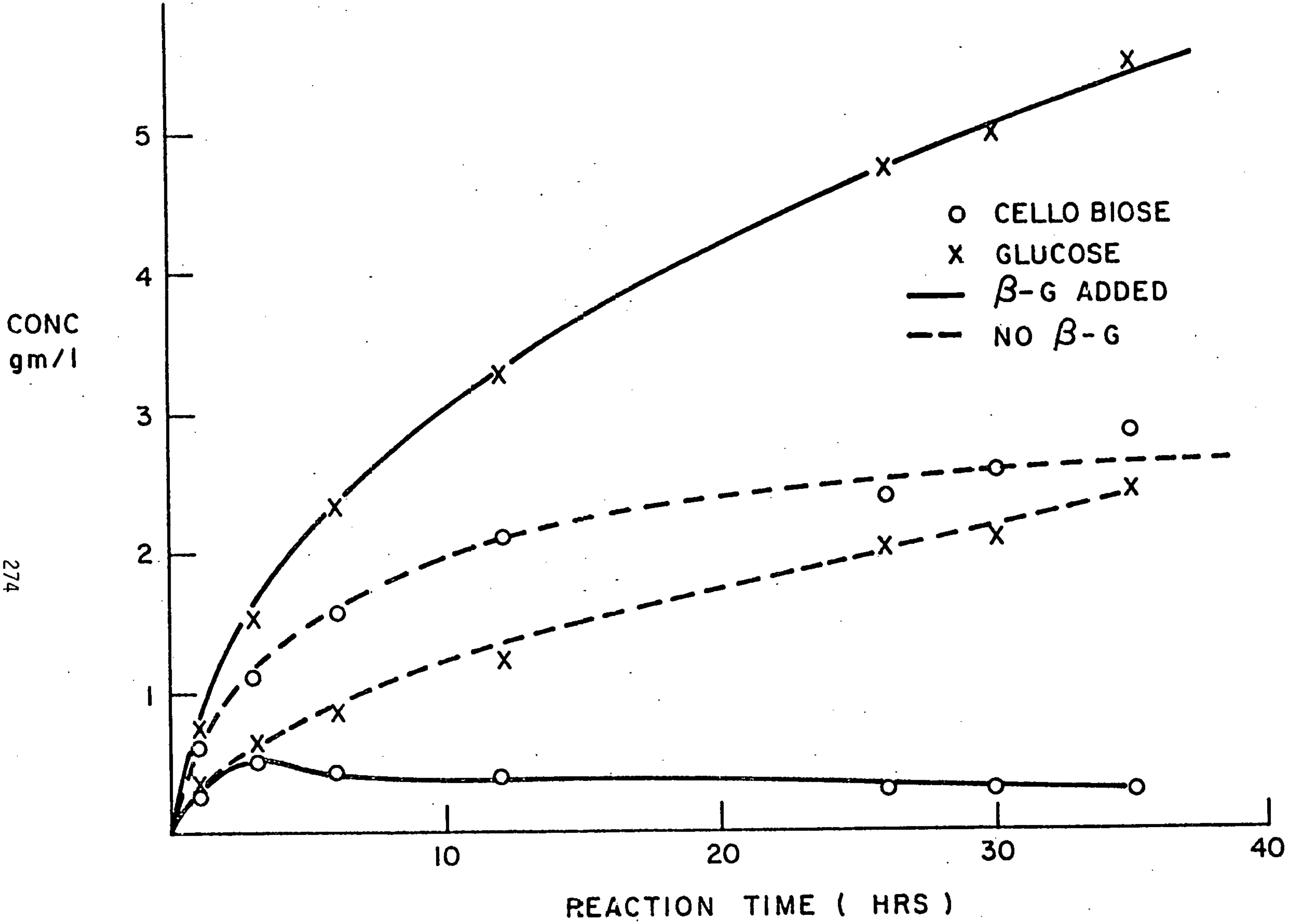
CONC
gm/l
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30
40
REACTION TIME (HRS)
o CELLO BIOSE
X GLUCOSE
β-G ADDED
NO β-G

BIOCONVERSION OF PLANT BIOMASS TO ETHANOL

Ronald Brooks, Tah-Mun Su, Michael Brennan, and Joanne Frick

General Electric Corporate Research and Development, Schenectady, NY 12301

ABSTRACT

ne General Electric Corporate Research and Deelopment (GE/CRD) process employs low pressure nemically augmented hardwood steaming and rapid ecompression to produce a readily digestible subtrate. A mixed culture of Clostridum thermocellum nd Clostridum thermosaccharolyticum ferments the retreated wood at 60°C directly to ethanol. C. nermocellum is used primarily to solubilize cellupse and to convert cellobiose to ethanol. Coulture with C. thermosaccharolyticum enhances the ate of cellulose degradation and permits the xylose roduced during pretreatment to also be fermented to thanol. Product recovery is accomplished via coninuous withdrawal of the fermentation broth to a eparate vacuum distillation chamber (modified vacuerm) and subsequent distillation to produce 95% thanol. After ethanol separation, the cell mass is eturned to the fermentor to maintain cell density nile lignin is discharged, partially dried, and sed to fuel the pretreatment boiler. If practical, pent silage is recovered and used as fertilizer.

ne preliminary ethanol cost analysis is encouragng; however, pretreatment optimization and improveents in the overall mixed culture ethanol tolerance nd yield is required before larger scale process valuation is justified.

INTRODUCTION

ost competitive ethanol from biomass requires 1. n efficient means of pretreating plant biomass to nhance its susceptibility to microbial attack, 2. apid and efficient fermentation of cellulose and emicellulose at high substrate concentrations and . ethanol concentration high enough to minimize the ecovery cost (in Btus and dollars).

ne research effort described in this report is irected toward solving the technical barriers to ne production of ethanol from wood. The objective f the research is to demonstrate the technical and conomic feasibility of fermenting SO_2/steamed hardood directly to ethanol using a mixed thermophilic ulture of C. thermocellum and C. thermoaccharolyticum. Experimental results are summaized and discussed in terms of two major research culture physiology and development, and wood tment and evaluation. The former studies were undertaken to gain additional insight into nutritional requirements, pathway and limitations on the thermophilic fermentation of cellulose to ethanol. The studies of wood pretreatment were undertaken to determine the primary constraint(s) to microbial utilization of wood carbohydrates and to develop a process which renders the latter susceptible to enzymic or microbial attack.

RESULTS

A. Culture Physiology and Development.

1. Organisms. C. thermocellum LQ8 was provided by J.G. Zeikus, Department of Bacteriology, University of Wisconsin, Madison, Wisconsin. C. thermocellum Q was provided by L.Y. Quinn, Department of Bacteriology, Iowa State University, Ames, Iowa. C. thermosaccharolyticum ZC was isolated as a contaminant of LQ8.

2. Growth substrates and products. C. thermocellum Q grows on cellulose, cellobiose and glucose. Growth on cellobiose results in the production of glucose, lactic acid, ethanol, acetic acid, hydrogen and carbon dioxide. Glucose represents about half of the soluble sugar procustion arising from growth on cellulose and the remainder was assumed to be cellobiose. C. thermosaccharolyticum is noncellulolytic but forms a very stable mixed culture with C. thermocellum and grows on cellobiose, glucose and xylose. The fermentation products of C. thermosaccharolyticum from growth on cellobiose are similar to that of C. thermocellum except for variable amounts of butryic acid. Both cultures require relatively high levels (0.2-0.5%) of yeast extract for growth. A quantitative fermentation balance was deferred until a defined medium was developed.

The comparative growth of Q on cellulose, cellobiose, and glucose is shown in Fig. 1. Growth on cellobiose was faster than that observed on cellulose. This suggests that saccharification is limiting during growth on cellulose. Growth on glucose was preceded by a long lag period which was shortened but not eliminated by sequential transfer on glucose.

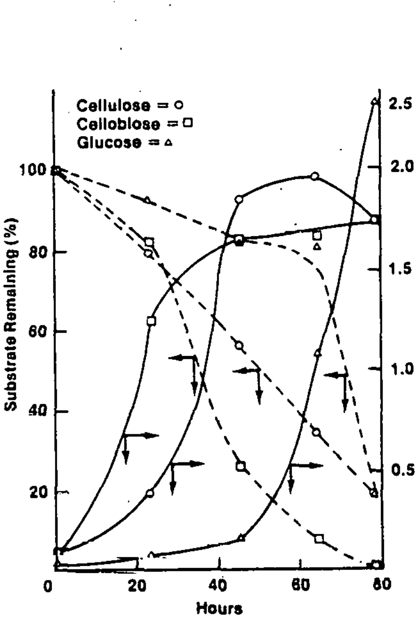


Fig. 1. Growth of C. Thermocellum Q on Cellulose, Cellobiose and Glucose

Product formation by C. thermocellum Q was found to depend on substrate (Fig. 2). The increased ethanol concentration from growth on cellobiose was due to greater substrate utilization. During growth on cellulose and cellobiose, glucose initially accumulated in the medium and was subsequently utilized.

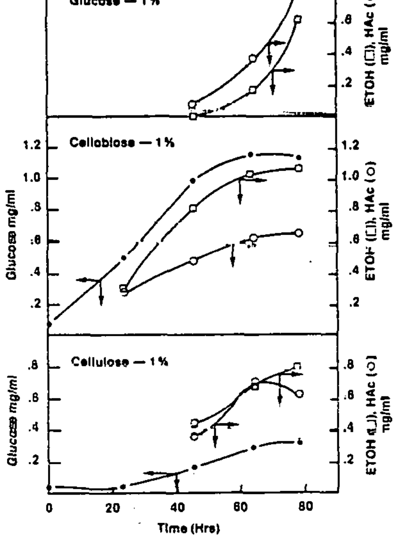


Fig. 2. Product Formation by C. thermocellum Q Grown on Cellulose, Cellobiose and Glucose

Cellulase production on cellulose, cellobiose, and glucose varied somewhat but was generally comparable and paralleled growth. These findings suggest that C. thermocellum cellulase is constitutive.

3. Cellulose from C. Thermocellum Q. During C. thermocellum Q growth on cellulose, the cellulase activity was primarily located in the supernatant and on cellulose. When cells were separated from the cellulose particles by settling, very little cell-associated Avicelase activity was detected even after an eight fold concentration of the cells. Glucose production during incubation of cell-free filtrate or the cell mass pellet with Avicel was not detected. Neither the whole culture nor the cell-free filtrate showed detectable -glucosidase activity as measured by the liberation of p-nitrophenol from p-nitrophenylglucoside.

The thermal stability of C. thermocellum Q cellulase (pH optimum 6.0) was examined between 40°C and 60°C. At 60°C, some loss of activity was observed, but the decline was gradual after 24 hours.

Glucose or cellobiose, in concentrations from 0.1 to 0.5% was not found to affect the cellulase activity of C. thermocellum.

4. Growth of C. Thermosaccharolyticum ZC. C. thermosaccharolyticum ZC grew equally well on cellobiose, glucose, and xylose. A lag period of about 20 hours was observed during growth on cellobiose (Fig. 3). Ethanol and acetic acid were produced in comparable amounts, and the ratio was generally between 1 and 1.5. The yield of ethanol was higher on glucose than on cellobiose. Product formation was growth associated but continued to increase after cell growth leveled off. As with C. thermocellum Q, C. thermosaccharolyticum ZC also accumulated glucose during growth on cellobiose.

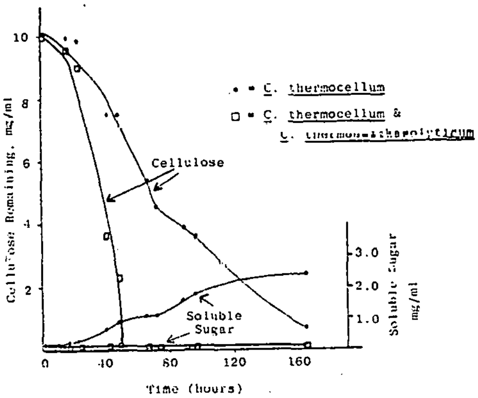


Fig. 3. Cellulose Degradation by Mixed Culture and Mono-Culture

5. Mixed Culture Cellulose Degradation. Co-culture of C. thermosaccharolyticum ZC with C. thermocellum Q on cellulose resulted in an enhanced rate of cellulose degradation and negligible soluble sugar accumulation (Fig. 3). The yield of ethanol and acetic acid, generally about 25% of the cellulose consumed, was comparable with that observed in the pure culture fermentation. Soluble sugars di accumulate because of utilization by ZC. How since exogenous sugar was not observed to stimula

ız production or to inhibit enzyme activity, the ıhanced cellulose degradation rate could not be :tributed to the reduced soluble sugar ɔcumulation.

. *Culture Development*. The necessity of improving ıe ethanol yield and tolerance of both *C. thermo-* ɪllum and *C. thermosacchoarlyticum* was recognized : the beginning of this research program. To ıhance the ethanol tolerance, cultures were either ɪquentially transferred into media containing in- ʼementally (0.5%) higher amounts of ethanol or ɪparated (via centrifugation) and incubated in ʼesh medium containing exogenous ethanol. A strain ʼ *C. thermocellum* Q-5 which was able to grow in ɪdia containing 5% ethanol was isolated by the ıtter procedure. In media containing no added :hanol, the ratio of ethanol to acetic acid produced by Q-5 increased to more than twice that ɔserved with the parent strain Q.

.milar efforts with *C. thermosaccharolyticum* were ısuccessful. More extensive subculturing and ıaller incremental increases in exogenous ethanol ɔncentrations are apparently required to overcome ıe greater sensitivity of *C. thermosaccharolyticum* ɔ ethanol.

. Wood. Pretreatment and Evaluation

1. *Pretreatment Reactor.* The principle comɔnents of the experimental apparatus for wood preʼeatment are schematically presented in Fig. 4. ıe steam generator is a high pressure vessel that ın withstand 1800 psi. The capacity of the vessel ɜ one gallon, which is large enough to supply steam ɔ the pretreatment chamber. Both sides of the :eam pretreatment chamber are connected to ʼdraulically operated ball valves capable of with- :anding 500 psi at 250°C. The ball valves have a ıpid response time to ensure uniform fiber dis- ıarge. The use of ball valves allows for rapid ʼont-end loading and discharge of wood chips and .ber, respectively. Under these experimental con- .tions, the minimum residence time is limited ʼimarily by the time required to load the ʼetreatment chamber. The addition of chemicals ːfore steaming is accomplished via a high pressure .s port on the input side of the pretreatment .amber. The pretreatment chamber is a one inch ID ainless steel threaded pipe that is insulated and ated to preset temperatures. The discharge port connected to a sample collector.

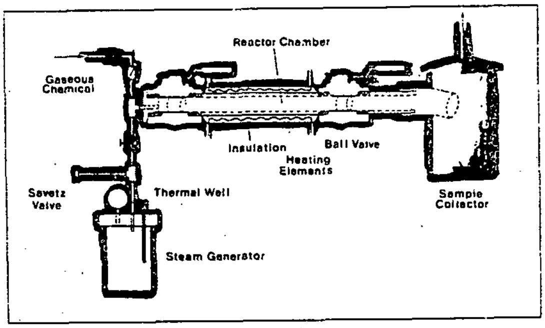


Fig. 4 Pretreatment Reactor Schematic

2. *Pretreatment.* The effect of steam essure and reaction time on fiber composition, rmentable sugar recovery, and susceptibility of eated poplar to enzymic hydrolysis were initially amined to determine a suitable pretreatment erating range. The yield of soluble sugars from zymatic hydrolysis of poplar fibers, steamed at D psi (205°C) was about four times greater than ıld be obtained from fibers treated at 200 psi 96°C). At these pressures, reaction time beyond minutes, generally resulted in decreased yields ɔle sugars; however, reaction time was less .ial than temperature. Subsequently, the ɔf separate additions of sulfur dioxide and nonia were also examined and compared with the effects of steam alone. Chemically augmented steaming was performed by initially contacting the poplar in the preheated (200°C) reactor with gaseous sulfur dioxide (30 spi) for about two minutes. The wood was decompressively discharged from the reactor and analyzed.

Steam and SO_2/steam treatment caused a decrease in the xylan and ash content of the fibers. However, operating on the basic side with ammonia significantly retarded xylose degradation. About 30% of the original Klason lignin was altered during each pretreatment to the point where it became soluble in 72% H_2SO_4. The yield of fermentable sugars (xylose and glucose) after steam or sulfur dioxide pretreatment at 240 psi was reduced unless the pH in the reactor was controlled.

The susceptibilities of treated and untreated poplar fibers to acid hydrolysis and enzymatic saccharification are summarized in Table I. Hydrolysis in 6N HCl provides a rough measure of the amorphous carbohydrate content of the fiber. The results indicated that little or no decrystallization of cellulose occurred as a result of the pretreatment. The relatively higher soluble sugar production from untreated fibers was because of their higher xylan content.

Enzymic hydrolysis of cellulose with *T. reesei* cellulase leads to the formation of both glucose and cellobiose. Since the soluble sugar yields reported in Table I were determined by glucostat assay, the yields of soluble sugars should be higher once cellobiose is accounted for. Extending the incubation time from 24 to 48 hours results in additional soluble sugar production, it is therefore likely that for SO_2-treated fibers, all of the carbohydrate can be saccharified in a reasonable period of time. Steam pretreatment alone resulted in a less dramatic improvement in fiber accessibility and ammonia-treated fibers were as resistant to saccharification as untreated fibers.

The ability of *C. thermocellum* Q to utilize sulfur dioxide pretreated poplar fibers for ethanol

TABLE I

SACCHARIFICATION OF PRETREATED POPLAR

	Soluble Sugar Production (mg)		
		Enzyme[b]	
	Acid[a]	24 hr	48 hr
Untreated	6.3	1.4 (6.1%)	2.1 (9.2%)
Steam	3.0	7.1 (30.6%)	10.3 (44.5%)
Steam + SO_2	2.0	12.9 (50.0%)	18.3 (70.0%)
Steam + NH_3	1.5	... (3.8%)	...

[a] 50 mg of fiber in 5 ml 6N HCl, incubated at 96 °C for 1 hr; soluble sugars were determined by dinitrosalicylic acid method.

[b] 50 mg of fiber in 2 ml enzyme solution which had Avicelase activity of 15 mg/ml (10% Avicel incubated for 24 hrs); sugars were determined by glucostat method.

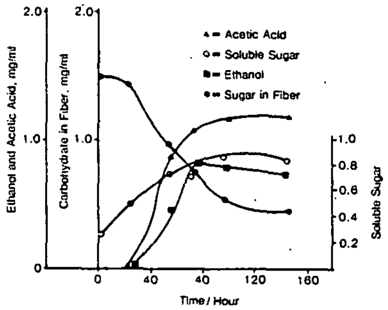


Fig. 5 Growth of C. Thermocellum Q on SO_2 - Treated Fiber

production was evaluated at 60°C in a pH (7.2) controlled fermentor (Fig. 5). Approximately 70% of the fiber was utilized in 48 hrs., indicating susceptibility comparable to that observed after incubation with R. reesei cellulase enzymes. The low ethanol concentration in fermentation was attributed to its removal by the nitrogen purge gas.

Treated and untreated fibers were examined by transmission electron microscopy to ascertain the nature of the structural alterations in fiber morphology which occurred as a result of pretreatment. Steam treatment resulted in extensive swelling in the S_3 layer and partial separation of the S_2/S_3 layers. Lignin diffusion (altered or native) and coagulation at the S_2/S_3 and, in some cases, at the P/S_1 interfaces were also observed. The loss of some intercellular substance (lignin) was apparent in some electron micrographs, but more extensive analysis is required to quantify the amount of cell wall material removed. Swelling of the primary wall and the S_1 and S_2 layers were limited relative to the extensive swelling which occurred in the S_3 layer.

The SO_2/steam treatment resulted in more extensive alterations in fiber morphology, and frequently, complete loss of cell wall structure was observed. The S_3 layer was generally absent, and a considerable portion of the intercellular substance, the primary wall, and the S_1 layer was removed. The S_2 layer was still discernible, but the loss of structure was apparent.

Ammonia treated poplar fibers retained considerable structural integrity and displayed extensive swelling of the S_3 layer. However, separation at the S_2/S_3 interface was infrequent, and completely detached lumen walls were not observed. Loss of middle lamella lignin appeared to be comparable or somewhat less than with samples treated with steam alone. Lignin coagulation and/or diffusion into the S_2/S_3 interface was not readily observed. The pH of ammonia treated fibers was about eight, which indicated that pH or a specific chemical interaction with sulfur dioxide or both were required to make most of the wood carbohydrate susceptible to microbial utilization.

DISCUSSION

1. Microbiology and Biochemistry of C. thermocellum Cellulose Fermentation. The experimental results permit the advancement of a tentative picture of the microbiology of pure anc mixed culture utilization of cellulose by C. thermocellum Q and C. thermosaccharolyticum ZC. During growth of C. thermocellum Q on cellulose, the culture elaborates extracellular cellulase whict hydrolizes the latter primarily cellobiose, that i: metabolized by C. thermocellum Q to produce glucose lactic acid, ethanol, acetic acid, hydrogen, an carbon dioxide. Similar end-products are produced b C. thermosaccharolyticum ZC grown on cellobiose. The failure to detect β glucosidose activity or glucos as an end-product of C. thermocellum cell-fre cellulase attack on Avicel suggests that the glucos in the broth arises from the metabolism o cellobiose. However, additional studies are require to substantiate this conclusion. Since glucose i expected as one end-product of endo-glucanase attac on cellulose, the luck of glucose in the hydrolizat suggests that endo-gluconase activity may be low The latter may explain the relatively low observe cellulose depletion rates.

Glucose accumulation in the broth also depend on the growth substrate. During the growth of Q o cellulose saccharification of the latter is rat limiting, the concentration of the actual growt substrate (cellobiose) remains low, which limits cel growth, and influences the amount of glucose that i utilized. However, during growth on cellobiose, th carbon substrate is plentiful and glucose accumu lation is controlled primarily by substrat preference.

The enhanced rate of cellulose depletion durin growth of C. thermocellum Q in the presence of C thermosaccharolyticum ZC appears not to be the resul of soluble growth factor provided by the latter. N demonstrable cellulose depletion rate enhancement were observed when C. thermocellum Q was grown c cellulose in the presence of added C. saccharolyticum ZC cell free supernatant. potential for achieving further increases in su strate depletion rates by coculturing is appare m comparison of the mixed culture cellulose lc... rate of C. thermocellum Q and C. rmosaccharolyticum ZC with that of the taminated C. thermocellum LQ8. The latter exded the former by almost a factor of three, even n basal medium yeast extract level was only 0.2%.

2. Wood Pretreatment. The experimental ults on wood pretreatment clearly confirm our ginal hypothesis that low pressure sulfur dioxide mented hardwood steaming should produce a readily essible substrate. Since the treated wood ains 75% or more of the original Klason lignin, latter does not appear to be the primary barrier enzymatic hydrolysis.

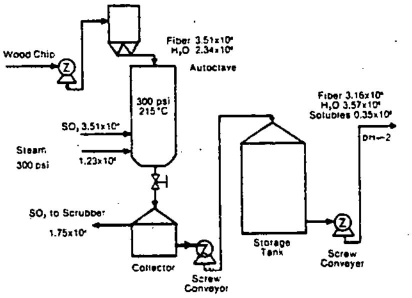
While the prelimimary experimental findings do permit identification of the primary siochemical barrier to enzyme accessibility, they provide considerable new insight and a basis for mulating a working hypothesis to guide future dies and the design of an optimum pretreatment cess. The electron micrographs are particularly pful in this regard. However, the results of the ctron microscopy analysis are very preliminary, several representative samples must be examined ore a general picture of the effects of steam and /steam treatment can be established. The erved changes in fiber morphology arise primarily m the combined effects of steam temperature, pH, rapid decompression; and these factors can be ependently evaluated in the reactor.

PRELIMINARY ECONOMIC CONSIDERATIONS

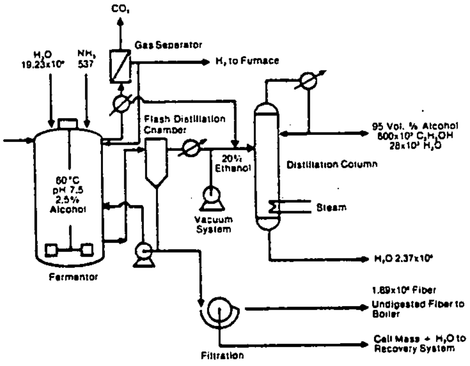
In the GE/CRD process (Fig. 6), relatively rse hardwood (and/or softwood, if amenable) chips steamed at low pressure in the presence of plemental amounts of sulfur dioxide (or ivalent reagent) for a brief time and are rapidly ompressed. The partially defibrated wood is tralized with ammonia gas and fed directly into a mentor which operates at a temperature of roximately 60°C. A mixed culture of Clostridium rmocellum and thermosaccharolyticum is employed ferment the readily digestible substrate to anol. C. thermocellum is used to solubilize lulose and to convert cellobiose to ethanol. C. rmosaccharolyticum is employed to ferment the toses produced during pretreatment but not lized by C. thermocellum, to ethanol. Product overy and cell recycle are accomplished by tinuous withdrawal of the broth to a vacuum tillation chamber (modified vacuferm) and sequent distillation to produce 95% ethanol. er ethanol separation, cell mass is returned to fermentor to maintain cell density while lignin discharged, partially dried, and used to fuel the treatment steam bioler. If practical, spent age is recovered and used as fertilizer.

A preliminary costs analysis of the GE/CRD cess was performed to estimate the costs of anol production and to determine areas where hnical improvement would lead to significant cost ...tion.

cost of producing ethanol via the GE/CRD cess was evaluated for a 30 x 10^6 gallon/year (300 ton/day) plant producing 95 volume percent ethanol from poplar. The size of the plant was dictated by available data and does not indicate an optimum or even desirable size. The material balance used to develop the cost estimate are shown in Fig. 6.



SUBSTRATE PRETREATMENT SECTION



FERMENTATION SECTION

Fig. 6 GE/CRD Direct Ethanol Fermentation Process

The following assumptions were made to arrive at a calculated manufacturing cost of ethanol.

1. Poplar wood available at $30/ton on a dry weight basis is fed into the reactor as chips containing 40 weight percent water as received.
2. Moist wood chips are contacted with gaseous sulfur dioxide (wood/SO_2 weight ratio = 100/1) under steam pressure of about 300 psi for 10 to 15 minutes.
3. Fiber recovery after pretreatment is 90% of the charge.

TABLE II

CAPITAL INVESTMENT

Pretreatment	Fermentation
Two 120 ton/hr 10" x 120" screw conveyors	Twenty-one 500 x 10^3 gallon fermentors
	Twenty-one agitators (100 hp)
Two 72 ton/hr 10" x 120" screw conveyors	Distillation System
Two 3 x 12' cylindrical autoclaves	Vacuum System and Compressor
Two 500 x 10^3 gallon storage tanks	Fermentation Condenser
Two 4000 gallon collector tanks	Process Piping
	Transfer Pump
	Package Bioler Plant
	Cooling Tower (6000 gal/min)
Fixed Capital. 1.72 x 10^6	Fixed Capital: 24.38 x 10^6
Total Capital: 2.24 x 10^6	Total Capital: 35.06 x 10^6

4. One-half of the pretreated fiber is fermentable carbohydrate and 90% of this carbonydrate is utilized during fermentation. The ethanol yield is 40%, based on the amount of sugar utilized.

5. The ethanol concentration in the fermentor broth is 2.5%.

6. Material costs for SO_2, NH_3, and steam are \$143/ton, \$120/ton, and \$2.50/1000 pounds, respectively.

7. Lang multiplication factor of 4.6 may be used to estimate the practical investment on delivered equipment costs (1). When 1978 equipment was not available, costs were updated using the December 1978 Chemical Engineering Plant Construction Cost Index of 245.2 (2).

8. Total capital investment includes contingency (15% of fixed investment) and working capital (20% of total capital investment).

9. Fixed charges for manufacturing costs include 8% depreciation, 8% interest, 1% property tax, 1% insurance, and 3% maintenance costs.

Nolan (3) has estimated the capital investment cost for a 30 x 10^6 gallon/year ethanol plant using combined saccharification and vacuum fermentation. Since similar equipment was used, the fermentation and vacuum distilation equipment costs were based on the estimates of Nolan and adjusted to 1978 costs using Chemical Engineering Cost Indexes of 245.5 and 226.2 for 1978 and 1979, respectively.

Based on these assumptions and the capital investment cost presented in Table II for a 30 x 10^6 gallon/year 95% ethanol plant, the manufacturing cost for a gallon of ethanol is calculated to be about \$1.06 (Table III). Ethanol costs of about \$1.25 gallon (pure ethanol) are considered achievable even with vacuum fermentation and ethanol yields which are only 50% of theoretical (3). The proposed modified vacuum fermentation process would use a much smaller system coupled to a conventional fermentor which should further reduce the capital investment.

TABLE III

PRELIMINARY COST ESTIMATES*

	Raw Material	Pretreatment	Fermentation
Total Capital Investment		\$2.24 x 10^6	\$35.08 x 10^6
Fixed Charge (¢/gal)		1.73	24.5
Direct Cost (¢/gal)			
Fiber	57.9		
Chemical		2.79	7.58
Steam + Lab + Utility		3.37	7.20
Manufacturing Cost (¢/gal)	57.9	7.70	39.30

*Cost does not include general expense, plant overhead cost, laboratory charge, land purchasing, building, service facility, yard improvement and startup.

The favorable ethanol cost estimate is a reflection of engineering simplicity, low cost pretreatment, high carbohydrate recovery, and the integration of the fermentation and ethanol recovery operations. Biomass costs represent about 58% of the ethanol manufacturing costs. With small-scale decentralized plants, the biomass costs may be lower than the \$30/ton assumed in the foregoing calculations. If the grower is also the ethanol manufacturer, the need and pressure to sell biomass for profit would be reduced because of the opportunity to recover a satisfactory profit on the sale of ethanol.

Based on these studies, the direct mixed culture thermophilic fermentation of SO_2/steam pretreated hardwood to ethanol is technically feasible and offers the potential of becoming economically attractive. The ethanol yield and tolerance of the microorganisms must be increased to justify the assumptions used in the cost estimate. Additionally, several avenues for improving the pretreatment operation and the overall productivity of the process have been suggested for future study and the prospects for progress are promising.

ACKNOWLEDGMENTS

This research was supported by the Fuels From Biomass Program, Department of Energy under Contract No. EG-77-C-02-4147.

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ENZYMATIC SACCHARIFICATION OF WASTE CELLULOSE

Mary Mandels
Enzyme and Biochemical Engineering Group
Environmental Sciences and Engineering Division
Food Sciences Laboratory
U.S. Army Natick Research and Development Command
Natick, Massachusetts 01760

ABSTRACT

The Natick contract with the Department of Energy (Interagency Agreement No. E49-28-1007, dated 23 February 1976) has the objective to develop a practical process for the conversion of cellulose in biomass to ethanol, a liquid fuel, via enzymatic conversion of the cellulose and yeast fermentation of the glucose. The process can be divided into five steps: 1. Selection of a suitable substrate; 2. Pretreatment of the substrate to enhance its enzyme susceptibility; 3. Production of active cellulase; 4. Utilization of the cellulase to saccharify cellulose; and, 5. Fermentation of the saccharification syrups to ethanol. The principal contributions of Natick have been in defining the basic microbiology and biochemistry of the process, particularly the identification of Trichoderma reesei as the best source of active cellulase, the realization that "cellulase" is a complex of enzymes that act synergistically with each other, and the demonstration of rapid extensive saccharification of many cellulosic materials with Trichoderma cellulase. Cellulase production has been increased by mutation of the strain and optimization of the fermentation. The efficiency of the cellulase has been increased by addition of supplemental β glucosidase from Aspergillus phoenicis. Various pretreatments including chemical, high pressure steam, and various types of milling increase substrate susceptibility. Enzyme has been produced at pilot plant scale (400 liters) at high titre (15g soluble protein per liter of broth) and productivity (85 filter paper cellulase units per liter per hour). Saccharification by Trichoderma cellulase plus Aspergillus β glucosidase has been advantageously coupled with yeast fermentation to ethanol, resulting in greater conversion than in the uncoupled system and preventing contamination of the saccharification reactor without addition of toxic or inhibitory chemicals.

or this process to be economically viable we must e able to produce glucose for 15 cents per kg in rude 10% syrup. To achieve this we must solve he following problems (Fig. 1).

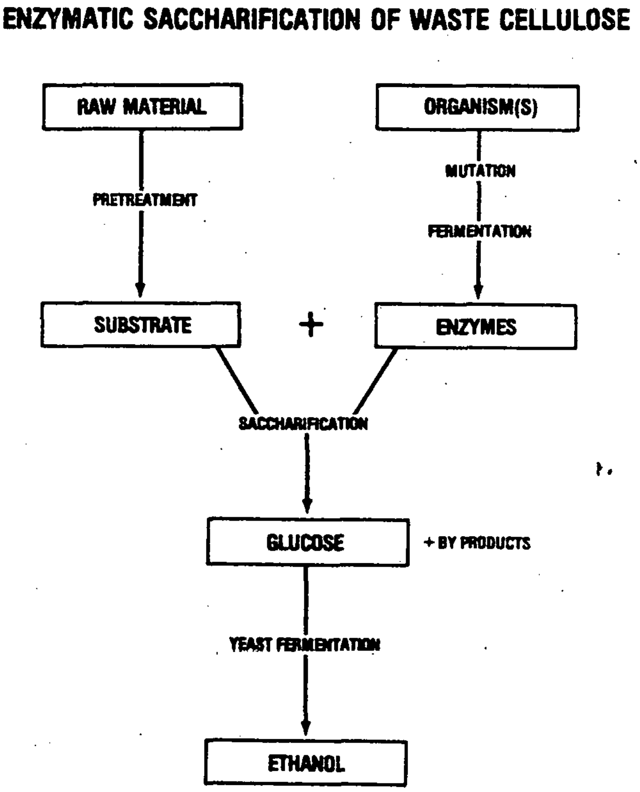

Fig. 1.

SUBSTRATE SELECTION

SUBSTRATE PRETREATMENT	CHARACTERISTICS 50% CELLULOSE	ALTERNATIVE USE
MUNICIPAL WASTE (FIBER FRACTION) 2 ROLL MILL (REMOVE PLASTICS) BALL MILL	LOW COST AVAILABLE YEAR-ROUND ALREADY COLLECTED	BURN DIRECTLY SHINGLES
PAPER MILL WASTE (FINES, SLUDGES) NO PRETREATMENT REQUIRED	LOW COST AVAILABLE YEAR-ROUND LIMITED QUANTITY VARIABLE	NONE
AGRICULTURAL WASTE (STRAW, BAGASSE) 2 ROLL MILL NaOH SWELLING	LOW COST SEASONAL (MUST BE STORED)	SOIL ENRICHMENT ANIMAL FEED BURN DIRECTLY
BIOMASS FROM ENERGY PLANTATION (POPLAR) STEAM EXPLOSION	HIGHER COST POTENTIALLY GREATER QUANTITY	THERMOCHEMICAL PROCESS BURN DIRECTLY

Fig. 2.

PRETREATMENT TO ENHANCE ENZYME SUSCEPTIBILITY

TYPE	ADVANTAGE	DISADVANTAGE
PHYSICAL		
HAMMER MILL	LESS COSTLY	INEFFECTIVE
COLLOID MILL	IMPROVES HANDLING	
BALL MILL	EFFECTIVE	ENERGY INTENSIVE
TWO ROLL MILL	HIGH BULK DENSITY	EXPENSIVE
STEAM EXPLOSION 220-240° 600 PSI	EFFECTIVE LESS COSTLY	LOW BULK DENSITY INEFFECTIVE SOFTWOODS
ALKALI SWELLING NaOH	MODERATELY EFFECTIVE LESS COSTLY	LOW BULK DENSITY MUST WASH OR NEUTRALIZE
DELIGNIFICATION CHEMICAL PULPING BIOLOGICAL	MODERATELY EFFECTIVE CONSERVE ENZYME, REACTOR SPACE	EXPENSIVE LOW BULK DENSITY POLLUTION
DISSOLVE, REPRECIPITATE CADOXEN (TOXIC) HYDRAZINE (TOXIC) FERRIC TARTRATE NaOH CUPRAMMONIUM VISCOSE CONCENTRATED ACID	EFFECTIVE	LOW BULK DENSITY EXPENSIVE DIFFICULT HANDLING SOLVENT RECOVERY SIDE REACTIONS POLLUTION

Fig. 3.

.. SUBSTRATE SELECTION. The ideal substrate, f existed, would be cheap, available in a _ quantity the year round, easy to collect, igh in cellulose and would require little or o pretreatment. At present municipal waste, aper mill waste, and biomass from energy lantations are most promising (Fig. 2). At atick most of our effort is devoted to municial waste.

. PRETREATMENT. The ideal pretreatment would e cheap and have a low energy requirement, reate no pollution problems and result in an ncreased cellulose content (i.e., removal of on hydrolyzable materials), increased eactivity, and increased bulk density. At resent the most promising pretreatments are all milling, two roll milling and steam xplosion (Fig. 3). The pretreatment of choice ill depend on the substrate selection (Fig. 4). valuation of pretreatments must be based on he increase in percent conversion over the ntreated control calculated for the total riginal substrate. Cost effectiveness should e based on all costs of pretreatment (preiminary grinding, steam, electricity, chemials, etc.) and be related to the increased ield of sugar. This calculation will not only ield a means of comparing pretreatments, it ill also (since we know the value of the ugar) tell whether a pretreatment is worth arrying out at all.

retreatment is a high priority problem which erits increased research support. If public ressure forces the diversion of grains to lcohol production, pretreated cellulosic astes could supplement animal feeds. The ypes of pretreatment that enhance enzyme usceptibility also enhance digestibility of ellulosic residues by animals. Experience ained in this use would be valuable when ellulose conversion plants are set up.

. PRODUCTION OF ACTIVE CELLULASE. At Natick nd several other laboratories Trichoderma eesei has been selected as a source of celluase because this organism produces high evels of a complete cellulase complex capable f hydrolyzing crystalline cellulose. Economic nalyses indicate that enzyme is the major cost actor in producing alcohol from cellulose (14). nerefore much of our research effort has been evoted to increasing cellulase productivity y mutation and by optimization of fermentation. ne problems revolve around the low specific ctivity of cellulase, the fact that cellulase s an induced enzyme, and the inverse relation etween growth and enzyme production.

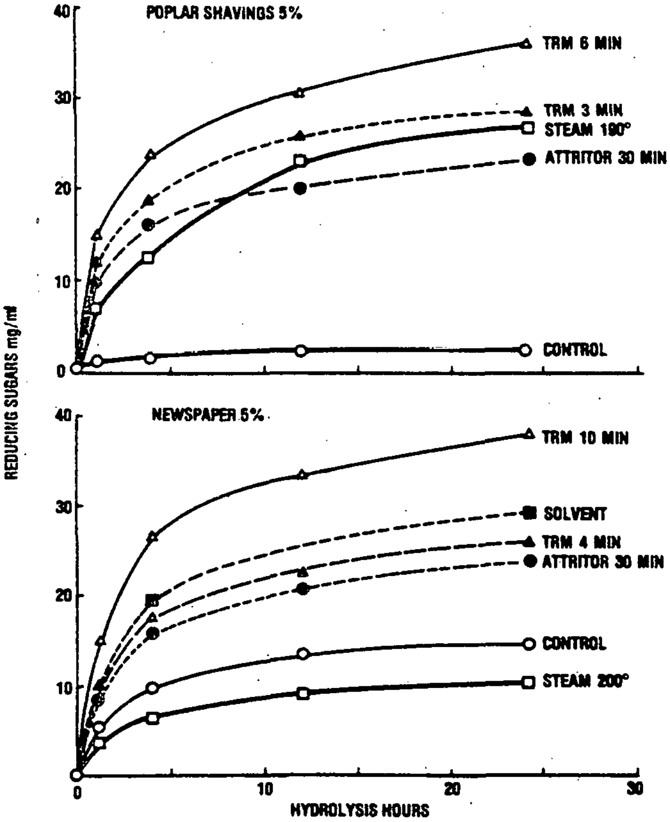

Fig. 4. Pretreatment of Hardwood (Poplar) Compared to Softwood (Newspaper). Hydrolysis at 50°, pH 4.8 with T. reesei Cellulase at 1 Filter Paper unit/ml.

TRM = 2 Roll Milled
Solvent = Dissolved in Cuprammonium - Reprecipitated
Data of Thomas Tassinari and Charles Macy.

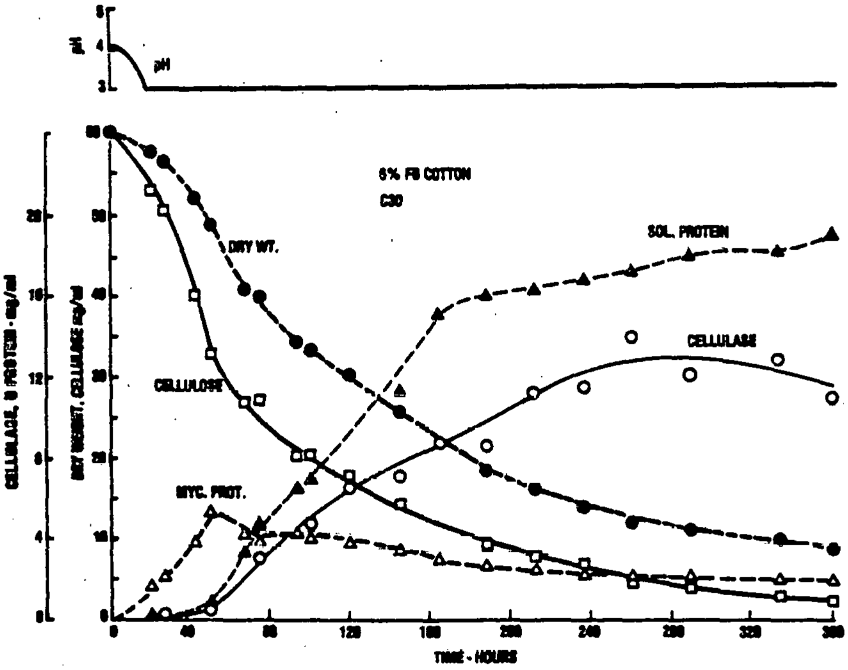

Fig. 5. Fermentation Profile of Trichoderma reesei Rutgers Strain C30 Growing on 6% 2 Roll Milled Cotto

FB = Farrel Birmingham Mill
Cellulase = International Filter Paper Cellulase units/ml (2)
Data of Raymond Andreotti.

Rutgers mutant C30 (9) is the best strain for production of cellulase on cellulose. High cellulose concentrations are required to achieve high enzyme titres since the specific activity of the enzyme is only about 0.6 filter paper cellulase units per mg of extracellular (soluble) protein. Significant quantities of cellulase are never found in the mycelium (16). On 6% two roll milled cotton growth dominates for the first 50 hours (Fig. 5). During this period half of the cellulose is consumed and pH falls rapidly to 3.0 where it is controlled by addition of 2N NH_4OH which also supplies nitrogen as required. Mycelial protein is synthesized rapidly reaching a peak of 5 mg/ml. Assuming 40% protein, this represents 12.5 g/liter of young cells, a 40% yield from the cellulose consumed. Very little extracellular protein or cellulase has been synthesized. The residual cellulose is more resistant and is consumed very slowly over the next 200 hour period of active enzyme production. The pH holds at 3.0 as cellulose falls to 5g per liter, and mycelial protein falls to 2g per liter representing about 7g per liter of older cells containing about 30% protein. Cellulase rises to 14 units per ml and extracellular protein to 18g per liter, 9 times the level of mycelial protein. Eventually when cellulose is depleted pH will rise as cells autolyze.

Ideally we would like high enzyme concentration and a more rapid fermentation. With a more susce tible substrate, ball milled cellulose, and pH control at 4.0, growth and enzyme synthesis are more rapid, but cellulose is consumed and the fermentation is complete by 100 hours yielding only 7 filter paper cellulase units per ml (Fig. 6). On the same substrate, but with pH control a 3.0, growth and enzyme synthesis are slower, but the fermentation lasts longer and cellulase reach 12 units per ml at 150 hours. This fermentation gave the highest productivity, 73 filter paper cellulase units per liter per hour. When the mor resistant substrate, two roll milled cotton, was used and pH controlled at 3.0 a longer fermentation, 250 hours, and higher titre, 14 units/ml, was achieved but overall productivity 54 units pe liter per hour was lower than for the intermediat fermentation.

A number of excellent mutant strains of Trichoder reesei are now available (Fig. 7). The old Natic strain, QM9414, has been supplanted by newer strains, Rutgers NG14 (9), highest titre on cellulose; Rutgers C30 (9), high titre and rapid growt on cellulose; and Natick MCG77 (5), rapid growth and high productivity on lactose. Strains C30 an MCG77 show less severe post glucose repression of

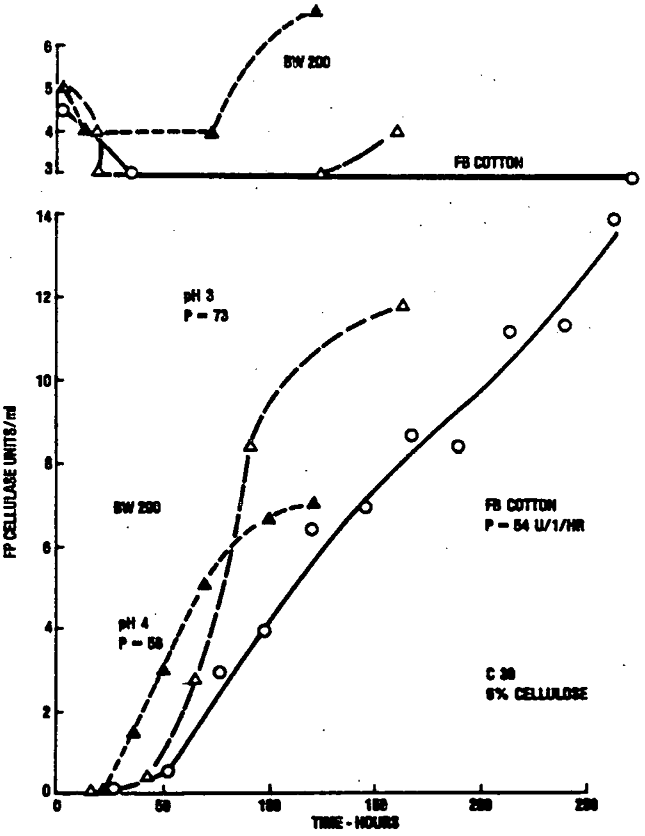

Fig. 6. Effect of Substrate Type and pH Control on Cellulase Productivity of *Trichoderma reesei* Rutgers Strain C30 Growing on 6% Cellulose.

FB = Farrel Birmingham 2 Roll Mill
BW200 = Ball Milled Pulp
P = Productivity - Filter Paper Cellulase units/liter/hr
Data of Raymond Andreotti.

cellulase production although none of these strains are truly derepressed and all still require a cellulase inducer such as cellulose or lactose. When the strains are compared under identical conditions the differences appear to be related to rapidity of growth, secretion of enzyme, and sensitivity to catabolite repression rather than to any fundamental changes in the cellulase genetics. Specific activities and ratios of activities on CMC, filter paper, and cotton are not markedly different. All strains produce only low levels of cellobiase under conditions favoring high cellulase productivity. The ideal ratio for saccharification is 1.5-2.0 units of cellobiase per filter paper cellulase unit. Dr. Frank Bissett at Natick has developed a procedure for separation of the proteins in crude cellulase preparations by an HPLC procedure using a DEAE coated glass beads column (4). The profiles produced show marked changes dependant on growth substrate and handling of enzyme preparations, but when grown and handled alike the different mutants give very similar protein profiles with 64-80% of the total protein represented by exo-β glucanase peaks, 20-36% of the total protein by endo-β glucanase peaks.

The Natick prepilot plant produces fairly large quantities (400 liter scale) of cellulase and β glucosidase for use in saccharification experiments. A typical fermentation run is 100 hours. Strain C30 grown on 6% BW40 (unbleached pulp) yields 8.5 filter paper cellulase units/ml, productivity 85 units per liter per hour, in such a run. Yields are higher for β glucosidase which has a much higher specific activity (150 units/mg protein) than cellulase. *Aspergillus phoenicis* grown on 3.5% hydrolysis sugars yields 18.5 cellobiase units/ml, productivity 185 units per liter per hour.

4. UTILIZATION OF THE CELLULASE TO SACCHARIFY CELLULOSE. Some of the problems involved in saccharification are shown in Fig. 8. We have already discussed strain selection and production of the enzyme. The rate and extent of saccharification depend on the nature and pretreatment of the substrate, enzyme and substrate concentration (i.e., ratio of enzyme to substrate), product inhibition, and enzyme stability. All of these interact to cause the rate of hydrolysis to fall off rapidly with time (7). The most important factor affecting rate and extent of hydrolysis is substrate multiplicity (Fig. 9). Even when syrups are removed and replaced with fresh enzyme to eliminate product inhibition and enzyme inactivation as factors, the time required for total hydrolysis varies from less than 4 days for roll milled pulp to 35 days for untreated cotton. The most important function of pretreatment is to reduce crystallinity and so increase enzyme susceptibility. As the more susceptible portions of the cellulose are hydrolyzed, the residue is increasingly crystalline and resistant, products accumulate, and they competitively inhibit the enzymes. Since cellulose is simultaneously

CELLULASE PRODUCTION BY MUTANT STRAINS OF TRICHODERMA REESEI

STRAIN	CMC u/ml	FILTER PAPER u/ml	β GLUCOSIDASE u/ml	PRODUCTIVITY FPU/LITER/HR	SOLUBLE PROTEIN mg/ml
QM6a (PARENT)	88	5	0.3	15	7
QM9414 (NATICK)	109	10	0.6	30	14
MCG77 (NATICK)	104	11	0.9	33	16
C30 (RUTGERS)	150	14	0.3	42	19
NG14 (RUTGERS)	133	15	0.6	45	21

CULTURES GROWN 14 DAYS IN 10 LITER FERMENTORS ON 6% 2 ROLL MILLED COTTON
pH CONTROL > 3.0 USING 2 N NH_4OH

ENZYME UNITS = μ MOLES GLUCOSE PRODUCED PER MINUTE IN STANDARD ASSAY

Fig. 7. Data of Raymond Andreotti and Charles Roche.

SACCHARIFICATION

PROBLEMS	SOLUTIONS
A. REQUIRE COMPLETE CELLULASE ENDO β GLUCANASES EXO β GLUCANASES β GLUCOSIDASE OTHER (XYLANASE, ETC.)	A. STRAIN SELECTION TRICHODERMA REESEI FOR CELLULASE ASPERGILLUS PHOENICIS FOR SUPPLEMENTAL β GLUCOSIDASE
B. LOW SPECIFIC ACTIVITY OF CELLULASE 1. MUST PRODUCE HIGH SOLUBLE PROTEIN 2. GLUCOSE REPRESSION 3. CELLULOSE INDUCTION	B. INCREASE ENZYME PRODUCTIVITY 1. OPTIMIZE FERMENTATION 2. MUTATION FOR DEREPRESSION 3. CONSTITUTIVE MUTANT?
C. DECLINE OF HYDROLYSIS WITH TIME 1. INCREASING SUBSTRATE RESISTANCE 2. PRODUCT INHIBITION 3. ENZYME INACTIVATION	C. STUDY ENZYME SUBSTRATE INTERACTIONS 1. PRETREATMENT 2. ADD SUPPLEMENTAL β GLUCOSIDASE SIMULTANEOUS SACCHARIFICATION AND ETHANOL FERMENTATION 3. MUTATION - CHEMICAL STABILIZATION

Fig. 8.

decreasing the inhibitor to substrate ratio rises. Furthermore cellobiose inhibition is much more se[illegible] for crystalline substrates. Supplemental β glucosidase can be added to remove cellobiose, but high levels are required because β glucosidase is competitively inhibited by its product, glucose, and as saccharification proceeds the glucose to cellobiose ratio becomes very high. Finally enzymes are inactivated under reaction conditions, and those enzymes, or complexes of enzymes acting on crystalline cellulose are most rapidly inactivated (12).

As a result of these factors the cellulase unit is only 10-15% efficient in a 24 hour saccharification; i.e., a unit produces not 259 but 26-39 mg of glucose in 24 hours depending on enzyme and substrate concentration and nature of the substrate, or it will require 26,000 to 38,000 cellulase units to produce one kg of glucose in 24 hours. Up to about 2/3 of the total enzyme, one β glucosidase unit can replace one cellulase unit (Fig. 10). With high concentrations of a susceptible substrate 10% glucose solutions can be produced in 14 hours in pilot scale (200 liter) runs.

5. FERMENTATION OF THE SACCHARIFICATION SYRUPS TO ETHANOL. The fermentation of syrups to ethanol in the saccharification reactor has been carried out with moderate success (8). Both *Saccharomyces* and *Candida* yeasts are compatible with the cellulase enzymes and the saccharification syrups. Ethanol is less inhibitory to the cellulase enzymes than equivalent quantities of glucose or cellobiose so that under equal conditions saccharification is greater in the coupled system. This also solves the difficult problem of preventing contamination of the saccharification reactor without addition of toxic or inhibitory chemicals. The coupling of the two systems also reduces capital costs. Disadvantages exist. The coupled system must be optimized for the yeast fermentation which requires temperatures to be reduced well below the optimum for enzyme hydrolysis. Although *Candida* yeasts can grow aerobically on xylose, *Saccharomyces* will not, and neither yeast can ferment xylose to ethanol.

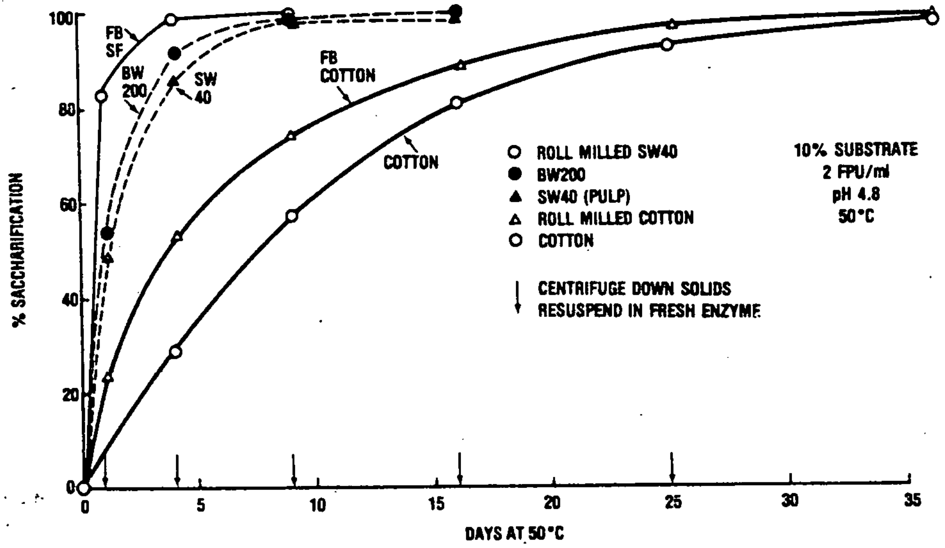

Fig. 9. Total Hydrolysis of Pure Cellulose Substrates of Varying Resistance.

FB = Farrel Birmingham 2 Roll Mill
SW40 = SF, Solka Floc - Pulp
BW200 = Ball Milled Pulp
Data of John Medeiros.

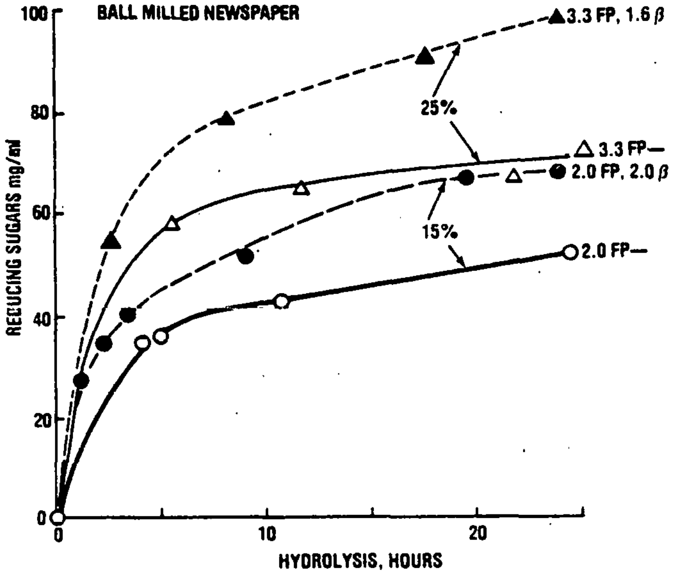

Fig. 10. Stimulation of Pilot Plant Saccharification of Newspaper by Added β Glucosidase.

▲, △ 25% Ball Milled Newspaper
●, ○ 15% Ball Milled Newspaper
FP = Filter Paper Cellulase units/ml
β = Added β Glucosidase (*Aspergillus phoenicis*) units/ml
50°, pH 4.8 200 liter STR
Data of Curtis Blodgett.

ACKNOWLEDGMENTS

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NOTES :

EXPERIMENTAL ANAEROBIC FERMENTATION FACILITY

Daniel J. Lizdas, Program Manager
Hamilton Standard
Division of United Technologies Corporation

Warren B. Coe, Project Engineer
Hamilton Standard
Division of United Technologies Corporation

ABSTRACT

The U.S. Department of Energy awarded Hamilton Standard a contract (EG-77-C-01-4015) on September 30, 1977, for the design, construction and experimental operation of an anaerobic fermentation facility to evaluate the technical and economic feasibility of producing a fuel gas from the residue of an environmental cattle feedlot.

Design began in October 1977. Construction was initiated in Bartow, Florida at the Kaplan ndustries feedlot in February 1978 and was completed in December 1978. Hamilton Standard will conduct an experimental evaluation through February 1980 to determine the commercial feasbility of the process.

OBJECTIVES

n order to meet the program objective of dentifying the technical and economic feasibility f producing fuel gas from environmental beef attle feedlot residues, the following design equirements were established for the system:

- Capacity to process 50,000 pounds of organic material (dry weight basis) per day.
- Utilization of the produced fuel gas in a commercial environment.
- Recovery of solids from the fermentor contents for use as cattle feed ingredient.
- Flexibility to explore a wide range of fermentor operating conditions.

he system design point was established at 55°C, ith a solids retention time of 10 days, utilizing he "as-received" manure slurry at a volatile olids concentration of approximately 10%. The esign flexibility gives the system the capability f:

- Operating at retention times as low as three days and fermentation temperatures up to 65°C simultaneously.
- Modifying the residue slurry by the addition of chemicals to control quality or to supplement the biological nutrients required, and by the addition of water to control solids concentration.
- Evaluating the thermodynamic, hydraulic and mechanical performance of a slurry-to-slurry waste heat exchanger for the recovery of heat from the fermentor discharge.
- Evaluating alternative effluent dewatering or product harvesting equipment and flocculation materials.
- Operating with a variety of loading schedules including continuous feed and withdrawal.

APPROACH

In the anaerobic fermentation process, organic materials are acted upon by microorganisms to produce organic acids. A second group of microorganisms utilizes the organic acids to produce a mixture of methane and carbon dioxide which is suitable as a fuel gas.

This biological process is carried out in tanks equipped with axial turbine-type mixers. The mixing insures that the operating conditions are uniform throughout the tank and provides contact between the microorganisms and the organic materials. Residue from feedlot cattle is continuously scraped and collected as a slurry in a holding tank. This raw material is then pumped into the fermentation tanks. A heat exchanger is used to transfer system waste heat into the incoming residue slurry, and steam is used to bring the slurry up to the process operating temperature prior to entering the tanks. Each tank is also equipped with steam injection which can be used for maintaining the process temperature. The system is capable of operating at temperatures of up to 65°C.

A portion of the tanks' contents are removed each day by a second pump to make room for the incoming raw materials. A centrifuge is used to harvest solids from the discharge flow, and the remaining liquid is sent to the lagoon.

Fuel gas produced by the process flows from the fermentor tanks to the system boiler and through an 800 foot pipeline to a boiler at the Kaplan Industries meat packing plant. The pressure, temperature and moisture content of the gas are controlled and safety functions are provided to prevent damage to the system. If the system and packing plant boilers do not require fuel gas, a flare is provided to burn the gas produced.

Conventional systems such as municipal sewage treatment plants using anaerobic fermentation processes produce fuel gas at costs greater than $10 per million Btu. This experimental facility, because of its advanced design concepts, is expected to initially produce fuel gas at less than $5 per million Btu. In addition, the facility will be used to explore two unique approaches, each of which has the potential to further decrease the cost of fuel gas production.

The first approach involves increasing the amount of fuel gas produced by the facility. This can be achieved by both increasing the quantity of organic material processed and increasing the fraction of this material which is converted into fuel gas. The use of increased operating temperatures, which increase the rate of biological conversion and increase the biological stability at high throughputs, is expected to achieve this improvement.

The facility, as designed and constructed, can process up to 25 tons of organic material each day, and produce a gross yield of fuel energy equivalent to 35 barrels of oil. 23% of this energy, in the form of delivered steam and electric power, will be required to operate the system. The flexibility built into the system and the planned experimental program will allow the exploration of improved process operating conditions and the evaluation of unit operations with the potential to increase the gross yield from a system of this size to 78 barrels per day, and to reduce the required energy for system operation to 7% of the gross production. If this potential is realized, the process practiced on this scale can be expected to produce a fuel gas for "close coupled" use (such as firing in the meat packing plant boiler) for as little as $2.50 per million Btu without any credit from byproducts or coproducts. This price is economically competitive with commercially available fuel.

The second approach to reduce the cost of fuel gas will be to demonstrate an economically valuable coproduct generated by the process. The microbial populations which grow in the fermentors and produce the fuel gas are rich in protein and hold promise for use as a dietary ingredient in cattle feed. The system includes equipment for harvesting this material, and the experimental program will utilize this protein in feeding trials with up to 6,000 head of cattle. Experiments to date indicate that feeding cattle with this product is only successful when the material is obtained from fermentors operated at elevated temperatures. The value of this material has previously been shown on a small scale to be about 60% of that of cottonseed meal. If substantiated on a larger scale by this program, coproduct credit could reduce the fuel gas cost to below $2.50, even without the improved process operating conditions discussed above.

ACCOMPLISHMENTS TO DATE

System design was completed in February 1978, and construction was initiated in March. Construction was completed in December 1978, and after system checkout, startup was initiated in February 1979.

Design

The system is comprised of five operating subsystems (loading, fermentation, thermal control, gas management and solids recovery) and an instrumentation and operational control subsystem. A greatly simplified system schematic is shown in Figure 1. The major equipment contained in the loading subsystem includes a centrifugal pump for maintaining circulation in the holding tank, a comminutor for particle size reduction, a pump for metering residue to the fermentation system, nutrient addition equipment, and the automatic valving required for fermentor loading.

In the fermentation subsystem the residue is held at an elevated temperature in two continuously mixed 320,000 gallon fermentors operating in parallel where the major portion of the volatile content of the residue is converted to a fuel gas. Each fermentor is equipped with a mixer consisting of three, four-bladed axial turbines which can be operated at speeds up to about 9 RPM. This range of speeds will allow evaluation of the effects of mixing rates as high as one tank turnover every 2.5 minutes.

The gas management subsystem measures, regulates, and distributes gas to the system boiler, the meat packing plant boiler, and the engine/generator. This subsystem consists of pressure regulators, flame traps, sediment traps, drip traps, a blower, gas meters, a pipe line to the Kaplan boilers, safety relief valves, and flares for burning excess gas.

The liquid effluent from the fermentation subsystem is dewatered for use as a high protein cattle feed supplement by a centrifuge in the solids recovery subsystem. Centrate is sent to the Kaplan lagoons. This subsystem includes the withdrawal metering pump which alternately withdraws slurry from each fermentor, automatic valving associated with the pump, valving which controls flow to and from the regenerative he exchanger, a centrifuge metering pump, a centrifuge and flocculent preparation equipment.

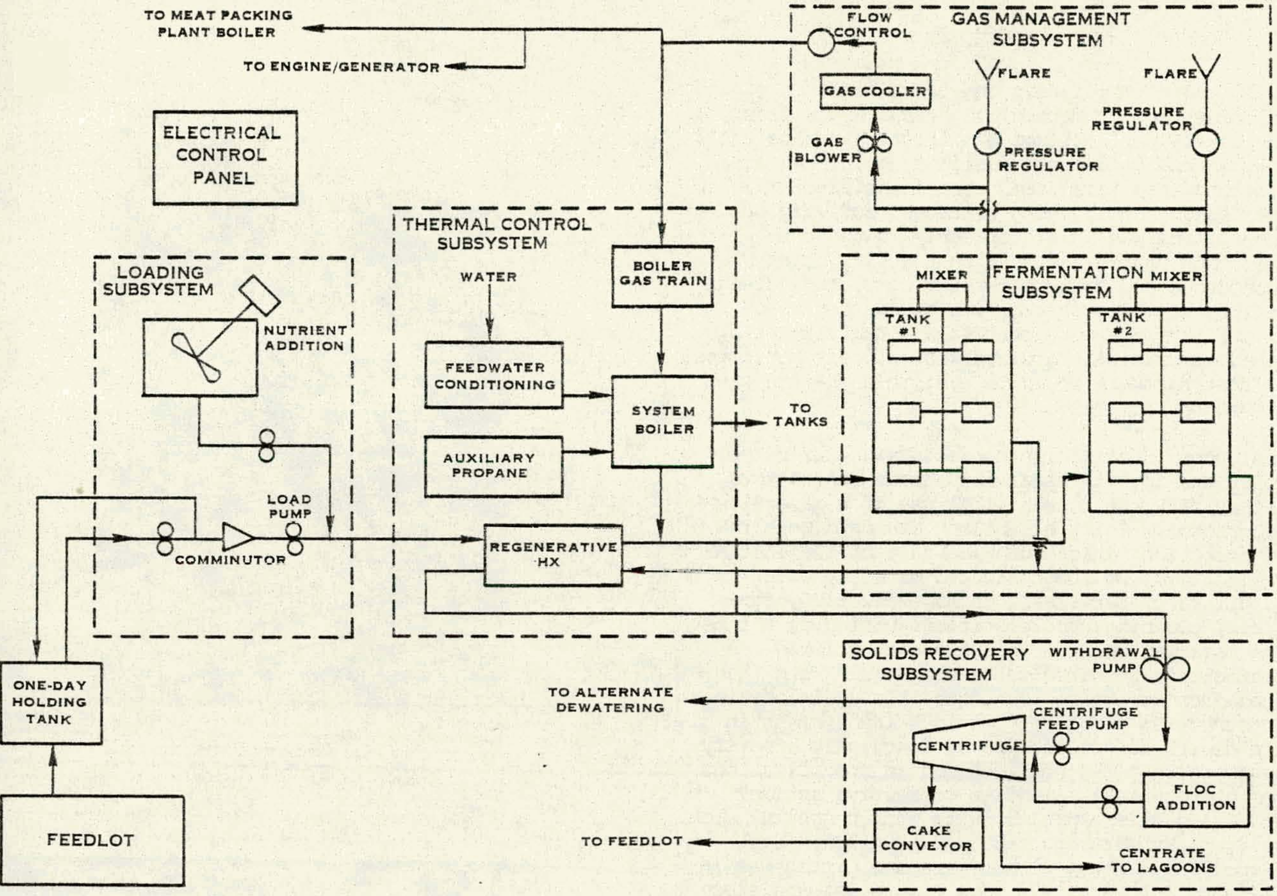


Fig. 1. Simplified System Schematic

he thermal control subsystem consists of a oiler, a heat exchanger, steam injectors and ssociated equipment. A regenerative heat xchanger transfers heat from the fermentor ffluent to the incoming residue thereby conerving system heat. Any additional heat equired for residue heating is provided by the team injector downstream of the heat exchanger. he fermentors are maintained at the proper emperature by automatic temperature controllers nd steam injection directly into the tanks. The team boiler can operate on either fermentor gas r propane. A feedwater heater and feed water reatment equipment are provided for boiler peration.

he instrumentation and operational control ubsystem distributes power to electrical comonents and provides the process operational rogramming for loading and discharging the ermentors through the use of time clocks, motorzed valves and the tank level gages. In addiion, the system monitors 16 different operaional parameters, compares their values with reset limits, sounds an alarm, and takes corrective action for six of these parameters when their limits are violated. An example of these corrective control actions is blower shutdown to prevent negative pressure in the tanks in the event of low gas production. Although primary tank structural protection against negative pressure is provided by dual vacuum breaker/flame arrestors, the blower shutdown feature avoids the entrance of air that might establish a combustible mixture in the tank.

Construction

Due to the stringent time constraints imposed by the contract, a "fast track" construction management approach was utilized. This approach had two distinguishing characteristics that made it quite different from conventional construction techniques. First, phased construction was conducted. At the completion of the formal design period, the design was separated into separate packages, each with different levels of design detail completion. The design of tanks, foundations, and associated site work was completely detailed at this time, so that site work could begin immediately and tanks could be placed on order. Other mechanical hardware was less completely defined, and electrical hardware was only at the conceptual design stage. However, all long lead time items were identified early in the design phase, and procurement activity on these items was initiated early in the construction phase. As work on the site, tank foundation and tanks was being completed, design details of mechanical hardware and then electrical hardware were completed. Delivery lead times of critical hardware were closely monitored so that hardware would be available on-site when needed for assembly.

The second distinguishing factor was our direct relationship with speciality trade contractors, rather than the conventional use of a general contractor. Using a Hamilton Standard on-site construction manager who was part of the initial design team, we subcontracted directly with suppliers of services such as excavation, concrete, plumbing, and electrical wiring on a time and material basis. This approach allowed the flexibility required to be compatible with the phased construction technique. In addition, the construction manager's previous involvement in the design effort allowed us to quickly identify hardware that did not meet design specifications and thereby take immediate corrective action. Utilizing this approach, site work began in March 1978, tanks were erected by June, 1978, (See Figure 2), mechanical hardware and piping was in place by October, 1978 (See Figure 3), and electrical hardware and wiring were in place with the system ready for dedication by December 1978 (See Figure 4).

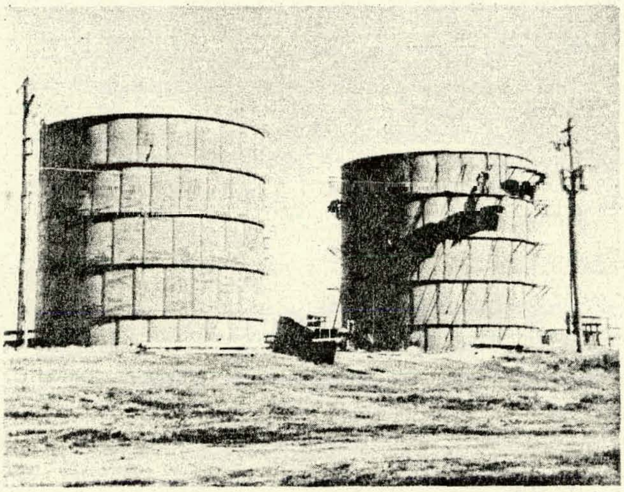


Fig. 2. Erection of Fermentation Tanks

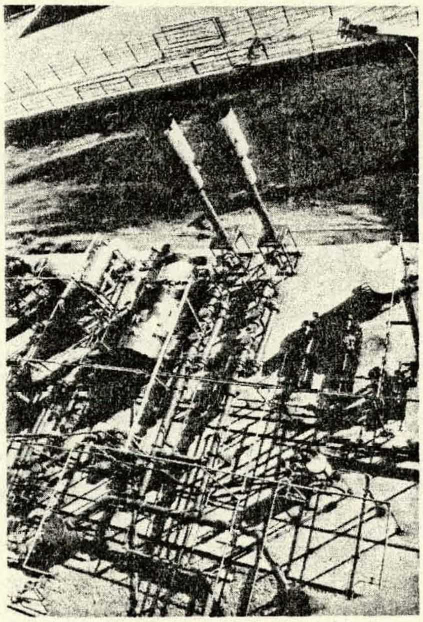


Fig. 3. System Piping

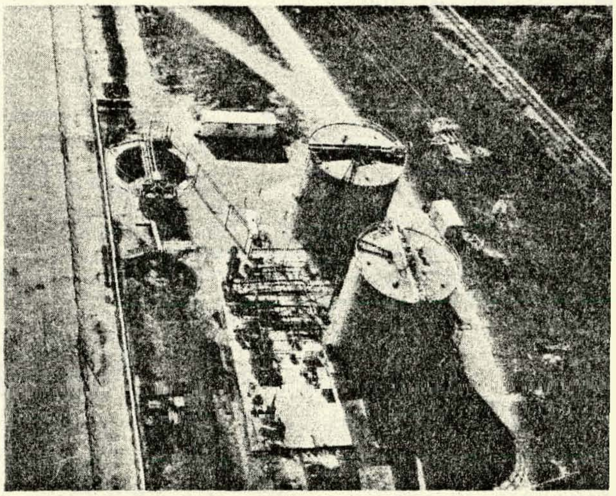


Fig. 4. System Construction Complete

Checkout and Startup

During December 1978 and January 1979 the system was functionally checked out. This effort consisted of activities such as calibration of variable speed load and withdrawal pumps, flow check and adjustment of pressure regulators and relief valves in the gas management subsystem start up and adjustment of the system boiler,

p .on of the main and auxiliary steam de-
i __ system. In addition, all of the elec-
rical power delivery systems were operated and
erified and the entire instrumentation and
ontrol system was subjected to trouble shooting
nd correction.

roblems were encountered with the boiler elec-
ronic safety programmer and with the main steam
rive proportioning motor due to moisture pene-
ration. These items are rated for service in
he open environment, but the occurence of very
trong winds and driving rains resulted in
orrosion. Modifications were made to eliminate
his deficiency. Additionally, the temperature
ontrol and recording equipment for the fermentor
anks was found to be excessively influenced by
ariations in ambient temperatures and was modi-
ied to correct the problem.

he system start up was initiated in February
979. Installation of the fermentation tank
nsulation was intentionally delayed in order to
llow maximum capability to correct leakage in
he bolted plate construction of the tanks under
perational conditions. Lower than normal
mbient temperatures and extremely high fuel
onsumption caused a reassessment of this action,
nd in March the system was shut down in order to
llow the addition of insulation to both tanks.
tartup was reinitiated and progressed slower
han had been anticipated. Residual pharma-
euticals in the manure appeared to be the
rimary cause.

UTURE PLANS

nce steady state operating conditions have been
chieved, activities will begin in two major
reas; utilization of the fuel gas and refeeding
f dewatered fermentor contents.

uel Gas Utilization

dual fuel 500 boiler horsepower Industrial
ombustion boiler, presently existing in the
aplan Industries meat packing plant is being
odified to burn the fuel gas produced by the
naerobic fermentation system.

he performance of the boiler while being operated
ith fermentor gas will be monitored. The
xperimental anaerobic fermentation system
as meters and gas analysis equipment will be
sed to determine the daily quantity of methane
uel gas consumed by the boiler, and the feed
ater meter will monitor the steam production.
his data will provide the daily boiler effi-
iency while operating on fermentor gas. The
ontinuation of the boiler operational log will
rovide data relating to maintenance.

comparison of the boiler efficiency while
perating on fuel oil and fermentor gas, and the
xisting cost of fuel oil will establish the
economic worth of the fermentor gas. The comparison of the operational history of the boiler while operating with both fuels will provide information regarding maintenance costs.

After the boiler tests are completed, the fuel gas will be utilized in a 450KW internal combustion engine/generator set in cooperation with Florida Power Corporation. The electricity generated will be fed to an existing electrical grid system, and a technical and economic evaluation of producing grid-connected electrical power from a small biomass source will be evaluated.

Fermentor Residue

Large scale feeding trials will be conducted to evaluate the dietary value and economic worth of dewatered fermentor contents (called Protein Fermentation Produced, or PFP) as a substitute for conventional supplemental protein sources. The amount of PFP produced by the experimental system will be verified by periodic weighing over a measured period of time while the quality of the PFP will be determined from composite samples. These samples will be used to determine the proximate analysis of the PFP for an evaluation of the material as a feed ingredient.

Kaplan Industries plans to use the PFP to conduct full scale experimental cattle feeding trials with up to 6,000 head of feedlot cattle. During these trials, the PFP will be evaluated at two different levels in the cattle diet. One group of the cattle will have all of their supplemental protein supplied by PFP while another will have only 50% supplied by PFP. A third group of cattle will have a control fattening ration. Based upon the data from the feeding trials and the marketplace value of supplemental protein, an effective value can be established for the PFP as a source of protein in feedlot cattle diets.

NOTES

ENGINEERING ANALYSIS OF ANAEROBIC DIGESTER CONCEPTS

by E. Ashare and E.H. Wilson
Dynatech R/D Company
Cambridge, MA 02139

BSTRACT

ngineering economic analyses were performed on vaious digester design concepts to determine the reative performance for different biomass feedstocks. he digester designs included in the analyses are STR, plug flow, and batch, and CSTR in series, ulti-stage digestion, and biomethanation were reiewed. Other process options investigated inclued pretreatment processes such as shredding, deritting, and chemical and thermal pretreatment, nd postdigestion processes such as dewatering, and as purification. The biomass sources considered nclude feedlot manure, rice straw, and bagasse, ith feed rates ranging from 50 to 1000 tons soids/day (4.5×10^4 to 9.1×10^5 kg solids/day).

he results of the analyses indicate that the most conomical (on a unit gas cost basis) digester deign concept is the plug flow reactor. This onclusion results from this system providing a igh gas production rate combined with a low capial "hole-in-the-ground" digester design concept. he costs determined in this analysis do not inlude any credits or penalties for feedstock or y-products, but present the costs only for conersion of biomass to methane. The batch land-fill ype digester design was shown to have a unit gas ost comparable to that for a conventional stirred ank digester, with the potential of reducing the ost if a land-fill site were available for a ower cost per unit volume.

he use of chemical pretreatment resulted in a igher unit gas cost, primarily due to the cost of retreatment chemical. A sensitivity analysis ndicated that the use of chemical pretreatment ould improve the economics provided a process ould be developed which utilized either less retreatment chemical or a less costly chemical.

NTRODUCTION

s a result of the energy crisis, a considerable ffort has been expended over the last few years o develop alternate renewable sources of energy. ne such source is biomass, which can be considered s solar energy being collected and stored by lants. Biomass can be grown specifically for nimals. Studies have been undertaken by various rganizations under contract with the U.S. Department of Energy (and its predecessor, ERDA) to nvestigate the feasibility of various biomass as potential alternate renewable energy s. These include grains and grasses, silviculture, aquatic plants, animal residues, and other crops such as corn and sugar cane.

Once there is an available source of biomass, the next step is to convert the biomass to a more easily utilized form of energy. Both thermochemical and biological processes are being considered as conversion processes for production of liquid and/or gaseous fuels. One such process is the biological anaerobic digestion of biomass which is used for production of methane gas.

There have been many approaches suggested for operation of an anaerobic digester, some of which are still in the laboratory or small pilot-scale stage. It is the purpose of this work to present an engineering and economic analysis of some of the conceptual designs for the anaerobic digestion of biomass, including the advanced as well as the new technology. Table 1 presents some of these processes, with an indication of some of the advantages of each. A schematic flow sheet, shown in Fig. 1, describes the process for conversion of biomass to methane.

SYSTEM ANALYSIS

The various digestion processes were analyzed by combining material balances, energy balances, and reaction kinetics in order to size the equipment. Costs for equipment were determined from relationships found in the literature, scaled and updated using appropriate factors and indexes. The calculation of unit gas cost was based on the public utility financing method developed in 1961 by the American Gas Association [1] and modified in 1971 by the Panhandle Eastern Pipeline Company. The procedure was described in a report from Esso Research and Engineering Co. to the Federal Power Commission [2]. It consisted of general bases for calculating total capital requirement, operating cost, and average unit gas cost. A computer program was developed to analyze the many possible combinations. The various digester concepts and other process options which were analyzed using the the computer program are presented in Table 2. The other digester concepts indicated in Table 1 are analyzed by extrapolation and/or comparison with the computer analyzed results. Comparisons are made of the various digester concepts by indicating differences in unit gas costs and the reasons for such differences. The effects of other process options on unit gas cost are presented.

Table 1 ECONOMIC ANALYSIS AND ASSESSMENT OF ANAEROBIC DIGESTER DESIGN CONCEPTS FOR BIOMASS CONVERSION

Digester Concept	Pretreatment (if integral to the digestion concept)	Remarks
Landfill	None	As practiced at Palos Verdes, City of Mountain View, and other sites, methane is withdrawn from an existing landfill of solid waste. This makes this digestion concept the "most simple" for biomass. Note that no pretreatment or admixtures of inoculum/buffer/nutrients are implied.
Controlled Landfill	Admixture of inoculum, nutrient, and buffer essential to process functioning. Pretreatment alternatives: a) None b) Shredding c) Shredding plus mild alkali	This concept implies admixing of the biomass source with nutrients, inoculum and buffer. Note that shredding and alkali pretreatment are additions to the digestion concept, i.e., they are not required, although they may enhance the rate.
Multi-Stage Digestion	Shredding/comminution to some degree appears essential. Chemical pretreatment is not required in that the first digestion "stage" is a microbially induced (acetic acid) treatment.	This system, the separation of acid formers and methane formers, has received the attention of a number of workers. CNG, for example, funded a multi-stage plug flow front end digestion system followed by a CSTR back end. IGT and University of Pennsylvania have also investigated this digestion concept.
"Channel Digester" (addressed by Biogas of Colorado, Ludington, MI group and Cornell University)	Dilution of the biomass and a substantial degree of comminution appears to be required. Mild alkali treatment may be included.	This appears to be a standard type of innovation for groups seeking low capital cost biomass digestion systems.
Packed Bed Digester	Minimal dilution mild alkali pretreatment may be an advantage.	Packed bed concept implies that the digester liquid is regularly drained and recycled for the conversion to occur, rather than mix the solids. Thus higher solids conversion is possible. Note: Development work to make the process continuous will be required.
Conventional CSTR Digestion (Traditional to Sewage Treatment)	Pretreatment alternatives: a) Comminution required b) Comminution and mild alkali treatment	Base-line for comparison. Prof. Pfeffer, for example, is accumulating a data base for biomass conversion.
"McCarty Process"	High temperature/high pressure alkali pretreatment developed recently by McCarty/Gossett, et al.	Organic compounds from pretreatment to be digested in a McCarty type anaerobic filter (extensions of process by Jeris, Converse, and Jewell).
"Porteous Saeman Process"	Acid hydrolysis of largely cellulosic material to wood sugars carried out. Subsequent conversion to ethyl alcohol practiced, but higher conversion efficiency to CH_4 may be expected.	Process used during WWII to produce ethyl alcohol. Initiated again by Porteous at Dartmouth for use with solid waste. Conversion of all products of acid hydrolysis to CH_4 is the objective (for ethyl alcohol production, only the cellulose fraction is utilized).
"Kukharenko Process"	Pretreatment similar to McCarty but other chemicals selected (K_2CO_3). Conversion to benzene-carboxylic acids is target goal for lignaceous fraction.	CSTR digester or anaerobic filter follows the pretreatment step. Suitable for other than annually harvested biomass sources.
Gasification plus Biomethanation	Gasification of total carbon in biomass converted to CO; CO_2 and H_2. Thus, the total lignin and refractory organic matter is utilized.	Anaerobic fermentation at high pressures of the gases CO, CO_2 and H_2 to CH_4 takes place. Perhaps the highest conversion rates possible with this system.

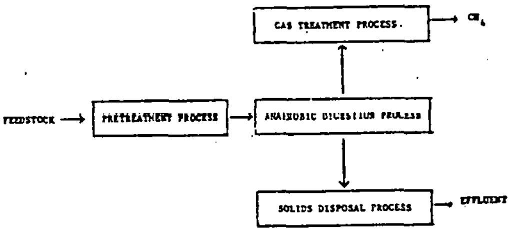


Fig. 1 A Flow Diagram Illustrating the System for Anaerobic Digestion

A sensitivity analysis of the effects of changing important system variables is also performed.

The systems analyses are presented in Figs. 2 and 3 as unit gas cost (\$/GJ) vs feed rate (kg/day) for feedlot manure and rice straw, respectively. These analyses utilize the assumption that the processes are operating without any interruptions. Costs presented do not include any feedstock cost or credits or penalties for digester effluent. Costs are for the processing operation only.

For each type of feedstock, the plug flow concept results in the lowest unit gas cost. A comparison with the CSTR system indicates several reasons for the difference in unit gas cost. The retentio time for a given fractional conversion is high for the CSTR design concept. (Relationships or y pretreatment equipment will be utilized only t-time basis during the loading step. For aw feedstock, with utilization of chemical etreatment, the batch system is about equivalent the CSTR system on a unit gas cost basis. In is case, the increased digester cost is balced by the increased production to result in proximately the same unit gas cost.

e unit gas cost for the batch digester concept s shown to be about the same as for the CSTR conpt with rice straw feedstock. It was assumed at the "landfill" type digester cost was $4/yd^3$ 5.23/m^3). If it were possible to develop a landll digester system for a lower per volume cost, e unit gas cost would be correspondingly reduced. e sensitivity of unit gas cost to batch digester r volume cost is shown in Fig. 4. The limit of ./ft^3 yields a $2.95/MM Btu unit gas cost. This miting case can be considered as the expected it gas cost resulting when no site excavation preparation is necessary.

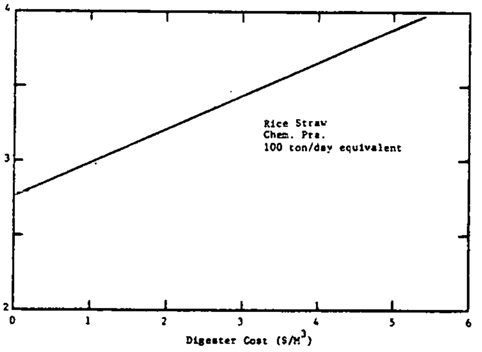


Fig. 4 Effect of Batch Digester on Unit Gas Cost

emical pretreatment is utilized to improve the gestibility of the feedstock. The cost of chemil pretreatment must be compared to the additional oductivity to determine economic feasibility. As n be seen from Fig. 3, the use of chemical preeatment results in over a 50% increase in unit s cost. This increase is due to changes in both erating and capital costs.

r rice straw as feedstock, the contribution to it gas cost due to capital costs decreases th use of chemical pretreatment. This can be plained by the significant increase in gas proction, so that even though the total capital sts increase, the capital contribution to unit s cost decreases from about $1.13/MM Btu to .91/MM Btu or about 20%. On the other hand, nce such a large fraction of the feedstock is nverted to biodegradable solids, a large quantity pretreatment chemical will be required. This sults in significantly greater contributions of erating costs to the unit gas cost, $3.28/MM Btu th chemical pretreatment compared to $1.51/MM Btu thout pretreatment, over an 100% increase. The t result is a 55% increase in unit gas cost.

Since the major contribution to increased unit gas cost associated with chemical pretreatment is due to the cost of pretreatment chemical, the sensitivity of unit cost to chemical cost was analyzed. The results of this sensitivity analysis is presented in Fig. 5 for rice straw. The baseline conditions for use of chemical pretreatment are $300/ton ($0.33/kg) for the chemical cost and 0.2 lb chemical/lb biomass converted to biodegradable material. The unit gas cost for the base-line case is $4.20/MM Btu. As the chemical requirement is decreased, the unit gas cost shows a corresponding decrease, as indicated in Fig. 5. The effect of decreased unit chemical cost is similar. It should be noted that if either the chemical requirement is decreased below 0.07 lb per lb converted or the cost is decreased to approximately $110/ton, the unit gas cost will be lower than the unit gas cost for no chemical pretreatment. This is a consequence of the increased gas production resulting from chemical pretreatment of rice straw which leads to a decrease in the capital contribution to the unit gas cost, as indicated above. These limits correspond to a 65% decrease in total chemical cost for pretreatment.

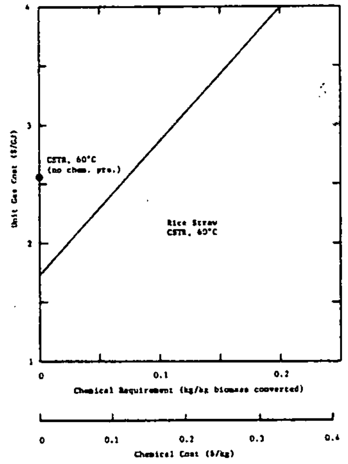


Fig. 5 Effect of Pretreatment Chemical Requirement and Cost on Unit Gas Cost

The use of other system components also will influence the unit gas cost, but not as significantly as chemical pretreatment. These other options should be considered only when it is necessary to ensure proper operation of the digestion process. For example, a degritter should be used for high ash containing feedstocks to eliminate any problem arising from ash build-up in the digester. Also, dewatering should be used only when necessary for handling of the effluent.

The use of a shredder was included for the systems with rice straw. It contributes approximately $0.17/MM Btu ($0.16/GJ) to the unit gas cost for a 100 ton/day (90,800 kg/day) feed rate. This increase in cost is due primarily to the capital cost of the shredder unit.

Table 2 PROCESS OPTIONS ANALYZED

Feedstock	Shredder	Degritter	Chemical Pretreatment	Storage	Digester	Heat Exchanger	Dewatering	Gas Purification
Environmental Feedlot Manure	N	N	N	Y	CSTR	Y	N	N
Dirt Feedlot Manure	N	Y	N	Y	CSTR	Y	N	N
Dirt Feedlot Manure	N	Y	N	Y	Plug	Y	N	N
Dirt Feedlot Manure	N	N	N	N	Batch	N	N	N
Rice Straw	Y	N	N	Y	CSTR	Y	N	N
Rice Straw	Y	N	Y	Y	CSTR	Y	N	N
Rice Straw	Y	N	Y	Y	Plug	Y	N	N
Rice Straw	Y	N	Y	N	Batch	N	N	N

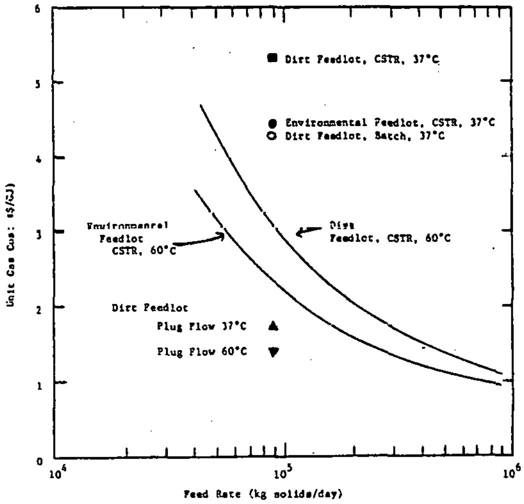


Fig. 2 Unit Gas Cost for Digestion of Feedlot Manure

retention time and fractional coversion are shown in Table 3). Thus, for the same fractional conversion the CSTR system will require a larger digester volume than the plug flow system. Another major factor affecting unit gas cost is digester cost. For the 100 ton/day (90,800 kg/day) feed rate, the cost for the CSTR digester was approximately \$130/yd^3 (\$170/m^3 whereas for the plug flow system, the cost was \$10/yd^3 (\$13/m^3). The CSTR system is a standard high cost reactor while the plug flow system is a low cost "hole-in-the-ground" reactor. The resultant unit gas costs are \$3.21/MM Btu (\$3.04/GJ) and \$1.49/MM Btu (\$1.41/GJ) for the CSTR and plug flow systems, respectively. The difference in unit gas costs is due primarily to the difference in digester costs. This emphasizes the economic advantage of the low capital system over the standard tank digester

For dirt feedlot manure feedstock the unit gas cost for a batch land-fill type digester (\$4.49/MM Btu) is higher than the unit gas cost resulting from a CSTR digester (\$3.21/MM Btu). This cost is due primarily to the increased digester cost for a batch system. Even though the per volume digester cost for a batch system is significantly lower than for the CSTR system, \$4/yd^3 (\$5.22/m^3) vs \$130/yd^3 (\$170/m^3), the retention time is longer and hence the digester volume required is much greater, 5.8 x 10^6 (165,000 m^3) vs 46,000 ft^3 (1600 m^3).

Table 3 RELATIONSHIP BETWEEN RETENTION TIME AND FRACTIONAL CONVERSION OF BIOMASS

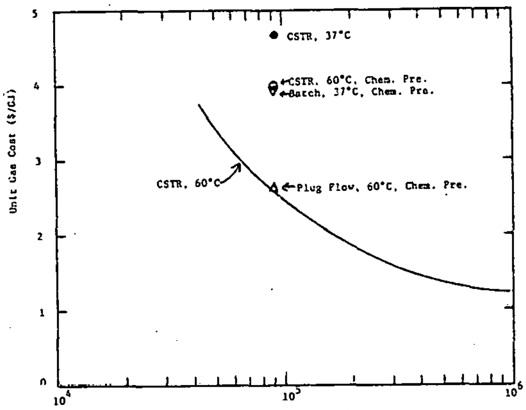
CSTR $$\tau = \left(\frac{1}{1-x} - 1\right) / k$$

CSTR in Series $$\tau = \left[\left(\frac{1}{1-x}\right)^{\frac{1}{N}} - 1\right] N/k$$

Plug Flow and Batch $$\tau = \left[\ln\left(\frac{1}{1-x}\right)\right] / k$$

τ = retention time
k = first order kinetic rate constant
x = fractional conversion of biomass
N = number of CSTR digesters in series

Fig. 3 Unit Gas Cost for Digestion of Rice Straw



This results in a greater digester cost for the batch system compared to the CSTR, \$865,000 vs \$274,000. The greater volume for the batch system is required to contain a one year loading of feedstock (assuming a one year cycle is required for filling, digesting, and emptying)

It should also be noted that for the batch s

d(ter was included in the treatment of dirt
e(manure since this feedstock has a high ash
ntent which could lead to digester inefficiency.
e use of a degritter results in only about 0.5%
crease of the system capital cost, and hence has
ttle effect on the unit gas cost.

e use of a gas scrubbing step will increase the
it gas cost about $1.33/MM Btu ($1.26/GJ). This
similar to the results presented by Ashare et
. [3] Since cost of gas scrubbing is so high
is option should only be used when necessary for
livery to a pipeline.

e use of a dewatering step to provide digester
fluent solids in a more concentrated form and
ssibly to provide recycle water will result
an increased unit gas cost of about $0.86/MM Btu
0.80/GJ). This option should be used when a
edit for the concentrated effluent compensates
r the increased cost, or if it is necessary to
tisfy environmental regulatory requirements.
wever, an evaluation of these by-product credits
requirements should be done on a site specific
case-by-case basis.

NCLUSIONS

e results of the engineering economic analyses
rformed on the various digester design concepts
d processing options have led to the conclusions
esented below. For these analyses, it was as-
med that the process were operating without
terruptions. The actual utilization of some of
ese design concepts to produce methane from bio-
ss should be preceded by preliminary experimental
rk to determine if the actual performance will
equivalent to the performance assumed for these
alyses. The results presented were for proces-
ng costs only, and did not include credits or
nalties for feedstock or effluent stream costs.

The most economical (based on unit gas cost) of the systems analyzed is the plug flow concept. This is a result of the high productivity (VVD) and low capital "hole-in-the-ground" design used for this concept.

The analysis for the batch "land-fill" type digester results in a unit gas cost which is comparable to the conventional stirred tank digester design. However, the cost for the batch system could be lower if the area for the landfill requires little or no preparation prior to loading the biomass.

The use of chemical pretreatment will result in a higher unit gas cost, primarily due to the cost of pretreatment chemical.

Chemical or thermal pretreatment could be potentially more economical, providing the increased gas production more than compensates for the cost of pretreatment equipment and operation.

Other system options such as shredding, detting, dewatering, and gas purification result in increased unit gas cost, and should only be used when necessary for efficient system operation or when necessary to provide purified gas for a pipeline or digester effluent solids as feed or fertilizer.

6. The results of this analysis do not incorporate any credits or penalties for the feedstock or byproducts, but present the costs only for the conversion of biomass to methane. An analysis of credits and penalties should be performed on a case-by-case basis.

ACKNOWLEDGEMENTS

This work was performed for the U.S. Department of Energy Fuels from Biomass Branch under contract No. EY-C-02-2991. *000.

1. Kavanaugh, J.F., "General Accounting Procedures to be Used for Large Scale Production of Gas from Coal and Oil Shale." American Gas Association General Accounting Committee, 1961.

2. Siegel, H.M., T. Kalina, and H.A. Marshall, "Description of Gas Cost Calculation Methods Being Used by the Synthetic Gas-Coal Task Force of the FPC National Gas Survey." EXXON Research and Engineering Co. Report to the Federal Power Commission, 1972.

3. Ashare, E., et al., "Evaluation of Systems for Purification of Fuel Gas From Anaerobic Digesters," Dynatech Report No. 1628, July 30, 1978, submitted to U.S. Department of Energy under contract No. EY-76-C-02-2991. *000.

NOTES

METHANE FROM AGRICULTURE RESIDUES
PROCESS CONVERSION EFFICIENCIES

John T. Pfeffer
Department of Civil Engineering
University of Illinois
Urbana, Illinois 61801

ABSTRACT

A large scale laboratory anaerobic fermentation system has been constructed to evaluate the efficiency of converting various agriculture residues to fuel gas. The system consists of four completely mixed and heated reactors with a working volume of 775 liters (205 gal). The system operates in a semicontinuous mode, receiving feed slurry every hour. Various feed stocks, including beef feed lot manure, corn stover and wheat straw have been tested at various retention times, fermentation temperatures and pretreatment techniques. Only the results of the corn stover and wheat straw studies are presented in this paper. The results show that the biodegradability of the substrates can vary substantially. The volatile solids in corn stover were found to be only 36 percent biodegradable at a fermentation temperature of 58 to 60°C. However, a mild thermochemical pretreatment increased the biodegradable volatile solids to 77 percent. Wheat straw that had been steam treated at 115°C to enhance water absorption was found to have volatile solids that were 49 percent biodegradable.

INTRODUCTION

The studies discussed in this paper have been conducted in conjunction with a research contract with the Biomass Energy Systems Branch of the Department of Energy (Contract No. EY-76-S-02-2917). The purpose of this study is to develop information on the conversion efficiency and kinetics of methane production from crop residues via anaerobic digestion. Crop residues as collected from the fields were used in these studies. Both corn stover and wheat straw have been processed in this system.

Corn stover is the residue remaining in the fields after the grain has been harvested. Because the grain is shelled from the cob, stover contains the cob, stalk and leaves of the mature corn plant. Stover was mechanically collected from the corn fields adjacent to the laboratory, and consisted of the above ground plant material. The dry stover was stock piled under cover prior to use. Baled wheat straw was obtained from a commercial supplier. This was stored inside the crain bay prior to use.

The processing system is described in detail elsewhere (Pfeffer and Quindry, [2]). All substrates processed through this system were subjected to the following processing:

- Particle size reduction
- Feed slurry storage
- Anaerobic fermentation
- Effluent storage
- Centrifugation

Because of the constraints of the processing system, it was necessary to reduce the particle size to prevent blockage of the feed and effluent piping. The dry milled residue was mixed with fresh water or recycle water from the centrifuge and stored in the feed slurry storage tank. This slurry was then pumped on an hourly cycle to heated and mixed anaerobic fermenters. The operating capacity of the fermenters was 775 liters (205 gal). The system could be operated at temperatures between ambient and 65°C and retention times ranging from 2 to 20 days. A level control in the fermenter activated effluent pumps that pumped the fermented slurry to effluent holding tanks. The slurry in these tanks was dewatered daily with a centrifuge. Under certain test conditions, the liquid was recycled for slurry water for the incoming residue. When this was not practiced, the water was discharged. The cake from the centrifuge was collected and subjected to various evaluations.

CONVERSION EFFICIENCIES

Residue Particle Size

It was necessary to pass the residues through a feed mill prior to preparing the feed slurry. Physical limitations of the fermentation system required relatively small particle size. Consequently, the residues were milled to pass through either 6.4 mm (0.25 in.) or 3.2 mm (0.125 in.) screens. The resultant particle size distribution (dry sieve analysis) is given in Table 1. The size of screen employed in the mill as well as the type of residue definitely influence the particle size distribution. For the corn stover milled through the 3.2 mm screen, 50 percent by weight of the particles were equal to or greater than 0.62 mm. Stover milled through the 6.4 mm screen had

50 percent by weight of the particles equal to or greater than 1.1 mm. Wheat straw milled through the 3.2 mm screen also had a medium particle size of 1.1 mm even though it was milled through a smaller screen.

Table 1. Particle Size Distribution of Milled Residue

Screen Size - mm	Percent by Weight Passing Screen		
	Corn Stover		Wheat Straw
	6.4 mm*	3.2 mm*	3.2 mm*
1.98	85	99	96
0.84	37	81	37
0.30	11	26	4
0.15	5	12	1.5
0.07	1.5	4	0.5

*Refers to screen size in the mill.

These particle size analyses clearly show that there is a significant difference in the response of the materials to size reduction. Corn stover yielded a more uniform size distribution than wheat straw. With the straw, 95 percent by weight of the particles were between 0.3 and 2.0 mm in size. With this relatively uniform size, straw exhibited some unique handling characteristics when slurried with water.

Gas Production Rates - Corn Stover

One primary objective of this contract was to determine the efficiency of the methane fermentation system in converting corn stover into methane gas. The four fermentation tanks were operated at different retention times at 59 ± 1°C on corn stover milled through the 3.2 mm screen. Previous studies [1] have shown that in this range of particle size, gas production was not a function of particle size. Since the slurry was easier to process with the smaller particle size, all of the results reported herein were obtained by milling the residue through a 3.2 mm screen.

The effect of retention time on conversion efficiency is shown in Table 2. The corn stover slurry was fed to the reactor at a continuous rate over a period of 12 to 16 hould daily (4 to 16 liters per hour). Reactor 3, the control reactor, had been operating for approximately 15 weeks at this retention time prior to collecting the date shown in Table 2.

Reactor 1 was operated at approximately 7.8 day retention time for 5 weeks. The data in Table 2 were taken only during the last 3 weeks of this run. This would represent a perior when the reactor could be expected to have reached equilibrium. The retention time for Reactor 2 was set at 5 days and operated for 4 weeks. Reactor 4 was operated at 200 liters per day feed rate, resulting in 3.8 day retention time. This feed rate was utilized for a 3 week period.

Table 2. Effect of Retention Time on Gas Production

Reactor No.	Θ Days	Measured CH_4 m^3/Kg V.S. Fed	Adj. CH_4 m^3/Kg V.S. Fed
4	3.8	0.047	0.070 (1.12)*
2	5.0	0.056	0.083 (1.33)
1	7.8	0.073	0.101 (1.62)
3	13.7	0.090	0.129 (2.07)

*( ) Data expressed as CF/lb V.S. fed.

During this period, the reactor effluents were screened to remove the coarse solids. Fine suspended solids passed through the screen and were incorporated in the recycle liquor used to slurry the fresh corn stover being fed to the fermenters. The organic solids contained in this liquor had been fermented and were essentially biologically inert. They added to the quality of volatile solids fed to the fermentation tanks, resulting in lower methane production rates as shown by the column labeled Measured CH_4. This methane production was calculated from the total volatile solids fed to the fermenters.

In order to determine the impact of the recycle of these solids on gas production, sampling of the recycle liquor was initiated. The total volatile solids of the recycled liquor varied significantly, ranging from low values of 5 g/l to high values of 15 g/l. This variation was due to a number of factors, including how well the recycle storage tank was mixed prior to using the recycle liquor to slurry more corn stover. By measuring the recycle solids concentration and the volume of recycle added, the mass of the recycle solids added to the feed was calculated. This allowed for the calculation of the correction factor. When this factor is applied to the measured methane production, an adjusted methane production rate is calculated. This is the gas production rate based on the volatile solids added as new corn stover.

The adjusted gas data can be used to determine the kinetics of the fermentation. The first step is to determing the biodegradability of the substrate. This must be done in order to establish S_o, the initial substrate level. S_o can be determined graphically by linearizing a plot of S_o-S (m^3 CH_4/kg V.S. fed) against the reciprocal of the retention time, Θ. When Θ approaches infinity, $Θ^{-1}$ approaches zero. Also, the substrate remaining, S, will also approach zero. Consequently, S_o-S will approach S_o.

A plot of the log of S_o-S, expressed as methane production, vs. the reciprocal of the retention time provide a reasonably good fit to a straight line, with a y-intercept of 0.16 m^3 CH_4/kg vol. solids fed. If one assumes that corn stover is a carbohydrate composed primarily of hexoses, the methane production at 100 percent conversion of substrate would be 0.45 m^3/kg (20°C gas temperature). Based on this maximum possible methane roduction and the observed S_o value of 0.16 m^3/kg, he corn stover appears to be only 36 percent biodegradable for the test conditions employed (59 ± °C fermentation temperature).

sing the above biodegradability factor and the djusted gas production in Table 2, a rate constant an be determined on the basis that the gas producion rate is first order with respect to biodegradble substrate. The following equation applies:

$$S = \frac{S_o}{1+K\Theta}$$

his equation can be linearized in the following orm:

$$\frac{S_o}{S} = K\Theta+1$$

he slope of the line will be K. A least square omputation for the data in Table 2, including a alue of $S_o/S = 1$ when $\Theta = 0$, yield a slope of .25 day^{-1}. Therefore, K is equal to 0.25 day^{-1}. his constant can be used to predict gas production ates and solids destruction at various retention imes.

hermochemical Pretreatment of Corn Stover

he milled dry corn stover was mixed with sodium ydroxide at the rate shown in Table 3. Water was dded at the rate of 4 kg per kg of dry stover. ith a caustic addition of 5 g/100 g of dry stover, he caustic concentration in this mixture was aproximately 0.3 molar. After thorough mixing of his paste in the pressure reactor, the pH at the bove caustic addition was 12.1 to 12.2. This mixure was heated for 4 hours at 115°C. After the eat treatment, the pH decreased to approximately .5. This paste was then diluted to about 6 perent solids and pumped to the mixing tank for feed o the reactors.

he results of this pretreatment are given in able 3. A caustic dosage of 5 g/100 g of dry olids resulted in a significant increase in ethane production rate of 0.25 m^3 per kg of volatile solids fed with an average retention time in he fermenters of 13.7 days. This pretreatment as repeated again in Reactor 1 during the period rom week 8 through week 11. With an average reention time of 14.2 days, the methane production as 0.25 m^3 per kg of volatile solids fed. If the ate constant for treated and untreated stover emains the same, the increase in biodegradability an be estimated from these data and the following quation:

$$S_o - S = 0.25\ \frac{m^3}{kg} = S_o\left(\frac{K\Theta}{1+K\Theta}\right)$$

he calculated value for S_o is 0.32 m^3 CH_4 per kg olatile solids fed. Based on an ultimate gas roduction of 0.45 m^3 CH_4 per kg volatile solids estroyed, the biodegradability of the corn stover increased from 36 percent to 71 percent.

Table 3. Effect of NaOH Pretreatment of Corn Stover

Week	Θ Days	CH_4 Prod.* m^3/Kg V.S.	Loading** Kg V.S./m^3-D	NaOH % Dry Solids
		REACTOR 1		
1	8.3	0.07	5.2	0
2	12.9	0.24	3.4	5.0
3	14.1	0.26	2.9	5.0
4	14.2	0.25	2.6	5.0
5	14.2	0.17	2.6	3.0
6	13.5	0.15	2.6	3.0
7	14.4	0.18	2.3	3.0
8	14.2	0.23	2.0	5.0
9	13.9	0.28	1.7	5.0
10	14.0	0.24	1.8	5.0
11	14.8	0.25	1.9	5.0
		REACTOR 2		
6	8.5	0.14	4.2	3.0
7	8.2	0.15	4.0	3.0
8	7.2	0.23	4.0	5.0
9	8.1	0.24	3.1	5.0
10	7.3	0.18	3.4	5.0

*Multiply m^3/kg by 16 to obtain cf/lb.
**Divide kg/m^3 by 16 to obtain lb/cf.

The above calculation applied to the data collected from Reactor 2 during the period from week 8 through week 10 yield an S_o value of 0.33 m^3 CH_4 per kg volatile solids fed. This would result in a 73 percent biodegradability. There was some variation in the three weeks of gas production.

Lower caustic dosages were investigated. The pretreatment for Reactors 1 and 2 during week 5 through week 7 was 3 g NaOH per 100 g dry solids. The initial pH of the mixed stover paste was in the range of 11.8 to 12.0. After heating, the pH was approximately 9.0. While there did not appear to be a major change in the initial and final pH, there was a substantial reduction in the gas production. The methane production in Reactor 1 operating at a 14.0 day retention time was 0.17 m^3 CH_4 per kg volatile solids fed. Reactor 2, operating at a retention time of 8.4 days, had a methane production rate of 0.145 m^3 per kg volatile solids fed. These data clearly show that significant caustic dosages are required to enhance the biodegradability of the corn stover.

Gas Production Rates - Wheat Straw

The data presented in Table 4 show the gas production obtained at various retention times. Each reactor was operated at the specified condition for at least 3 retention times before these data were collected. These data are for a fermentation temperature of 59 ± 1°C (138 ± 2°F). The straw slurry had been steam treated (115°C for 4 hours) prior to addition to the slurry holding tanks. This treatment was necessary to alter the characteristics of slurry so that it was possible to pump it.

Problems in pumping of this slurry were encountered. It was not possible to start pumping without first flushing the pipes with water. The time cycle pump activation system could not be used. Approximately 50 liters of slurry were added each time the pumps were activated. At the short retention time (3.8 days), four such pump periods were used daily. Unfortunately, this resulted in adding a substantial feed volume (200 liters into 775 liters) over a period of 6 to 8 hours. This mode of feeding resulted in an ustable system at low retention times.

The gas production in Table 4 is expressed in terms of the gas (methane) generated per unit of volatile solids added. When processing straw, a liquid recycle stream was employed. The fermented slurry was passed over a screen to remove most of the suspended solids. The total solids of this stream was reduced from approximately 20 g/l to 6 g/l by this simple screening process. The liquid fraction was used as make-up water for the new feed slurry. Consequently, a measurable quantity of solids was recycled with this system. The gas production in Table 4 is expressed in terms of the total volatile solids fed (including recycle solids) and in terms of the volatile solids added with the straw only, the latter being the data in the adjusted CH_4 production column.

Table 4. Gas Production Data for Straw

Θ Days	CH_4 Production m^3/Kg V.S. Fed	Adjusted CH_4 Production m^3/Kg V.S. Fed
3.8	0.083 (1.31)*	0.100 (1.58)
5.0	0.089 (1.41)	0.109 (1.72)
7.4	0.110 (1.74)	0.134 (2.12)
13.7	0.138 (2.18)	0.167 (2.64)

*( ) Data expressed as CF/lb V.S. fed.

The adjusted methane production was used to determine the percentage of volatile solids that are degradable. This is accomplished by extrapolating the gas production data to an infinite retention time. A semi-log plot of methane production against the reciprocal of the retention time yields a straight line. The y-intercept is the methane production at an infinite retention time, or the biodegradability of the organic material. Based on these data, the maximum possible methane production would be 0.22 m^3/kg volatile solids added (3.52 cf/lb). If one assumes that the organic material being fermented is cellulose, the methane production per kg of cellulose fermented would be 0.45 m^3 at a gas temperature of 20°C. Based on this maximum gas production, the volatile solids in the straw would be 49 percent biodegradable.

The rate of conversion is determined from a simple first order kinetic relationship in which the substrate removal rate (dS/dt) is a function of the biodegradable substrate remaining.

Using a value for S_o of 0.22 m^3/kg V.S. fed, and the values of S_o-S (methane produced per kg of volatile solids) from Table 4, a least squar t including the y-intercept value (0,1), yields a line with a slope of 0.23 day^{-1}. This rate constant (temperature of 59 ± 1°C) corresponds closely with the values found for manure (0.25 day^{-1}) [2] and corn stover (0.25 day^{-1}).

Discussion

During the past 10 years, several substrates have been studied as possible raw materials for the production of methane via the anaerobic fermentation mechanism. These substrates are in various stages of development, ranging from the laboratory stage as with the crop residues, to an operating demonstration plant at Pompano Beach, Florida, that is processing urban refuse. A comparison of these various substrates with present information can be useful to better understand some of the problems that must yet be overcome. For this purpose, the following matrix has been constructed. The + symbol represents a favorable factor that improves either the cost or energy conversion efficiency. Conversely, the - symbol represents unfavorable factors.

	Substrate Acquisition	Substrate Biodegrad.	Chemicals Required	Mixing Power	Net Energy Production	Heat Required
Urban Refuse	++	-	0	-	++	+
Manure Env. Lot	+	+	0	0	++	-
Manure Open Lot	0	=	0	0	0	-
Corn Stover	=	=	0	-	=	=
Wheat Straw	=	=	0	=	=	=
Pretreat Crop Residue	=	+	=	0	+	-

Substrate Availability. Economically, this facto can be highly significant. Urban refuse is a was material and the associated dump fee adds a substantial income to the process. Manure from the feed lots must be removed from the lots for prope lot management. Therefore, as with an open feed lot, there may be little or no economic gain or loss associated with the aquisition of this material. Frequent manure removal from the environmental lot generally required a continuous manure management program. Therefore, there may be an income, i.e. a reduced operating cost for the fe lot operator, associated with this substrate.

Crop residues have a significant aquisition cost This cost will be a function of many factors ranging from per acre production rate to harvesting techniques to cropping intensity of the regi Factors that may be equally important in det ing the availability of this material relate f g practices and the perceived or real value o residue when left on the land. Failure to consider these questions in developing a crop residue utilization system will lead to failure, either for economic reasons or because of accepted agricultural practices.

Substrate Biodegradability. Of all factors, this is the most important. The degree to which the substrate can be converted to gas has a significant impact on several economic and energy efficiency factors. In addition to greater quantities of product gas, a more biodegradable substrate allows for a more concentrated feed slurry. This reduces the reactor size and associated heat loss and it also reduces the quantity of water, and associated heat requirements, that pass through the process. Also, the fermented slurry processing systems are reduced in size.

Urban refuse is marginal as a biodegradable substrate. The volatile solids are only about 55 percent biodegradable. In addition, there is a substantial ash component. Manure from an open feed lot has been subjected to considerable natural stabilization on the lot and frequently contains considerable amounts of soil that contributes to the ash content. Environmental lots produce manure that is fresh, containing volatile solids that may be 65 percent biodegradable [2]. Consequently, this manure is the preferred substrate.

Corn stover and wheat straw are relatively stable organic materials. Approximately 36 percent of the volatile solids in the stover were found to be biodegradable. Heat treated straw volatile solids were only 49 percent biodegradable. Consequently, severe economic and energy penalties are applied to these substrates as a result of the requirement for larger reactors and reactor slurry processing systems as well as the heat loss associated with the large quantities of water passing through the system.

Thermochemical pretreatment of the crop residue significantly alters the process. Corn stover volatile solids increase to 71 percent with a mild thermochemical pretreatment. With gas priced at \$1.90/GJ (\$2.00/10^6 BTU) Pfeffer and Quindry [3] reported that thermochemically pretreated corn stover could not be economically processed. Depending upon methods of financing, gas priced at \$3.80/GJ (\$4.00/10^6 BTU), and excluding substrate acquisition costs, would be economically attractive. Therefore, it appears that thermochemical pretreatment may offer sufficient improvement in the crop residue fermentation to warrant further development.

Chemical Requirements. Chemical requirements in all cases except for thermochemical pretreatment are minor. In this system, the cost of sodium hydroxide is substantial. Effective use of this system will require careful optimization of the caustic consumption in the pretreatment step.

Heat Requirements. Except for urban refuse, all ns require an external heat source. An in- itor is included as part of the refuse pro- g system and, consequently, excess heat energy is available. Because of the poor biodegradability of both stover and straw, excess water is processed through the system. As a result, significant additional quantities of heat are required. In fact, this excess heat in conjunction with the low gas yield, will result in net energy production of zero or even negative. A relatively high temperature is required for thermochemical pretreatment, but because of the low water content used in this cooking, only a limited quantity of heat is required. Most of this heat energy is recovered as heat for the feed slurry.

Mixing Power. The fibrous nature of urban refuse required a somewhat higher power input for mixing than required by the manure or pretreated crop residue. The poor conversion efficiency of stover requires a larger reactor per unit of energy produced. This large volume increases the mixer power significantly. In addition to poor biodegradability, the straw possess unique properties when mixed with water. At low solids concentration (less than 10 percent) the slurry is a semi-solid and very difficult to mix or pump. Consequently, additional power is required.

Net Energy Production. Economics are important, but the economic feasibility will change as the price of energy increases. However, net energy production should always be positive for any long term application of this system. Clearly, more useful energy must be produced than is input to the system in the form of heat and electrical power. For the urban refuse system, approximately one-third of the total output energy is required to operate the system [4]. For an environmental feed lot using heat recovery on the fermenter effluent, less than 15 percent of the energy produced by the process is required for process operation. However, for an open lot, as much as two-thirds of the energy production is required to operate the system [2]. For processing corn stover, essentially all of the energy produced by the fermentation is required to operate the system [3].

Crop residue that has been thermochemically pretreated presents a much better energy efficiency. The improved biodegradability significantly increases the gas production. It also greatly lowers the fermenter size and the mass of water processed through the fermenter. The reduced quantity of fermenter residue greatly reduces the dewatering costs and energy use. Consequently, less than 30 percent of the energy produced by stover fermentation is required for process energy, both heat and mechanical [3].

SUMMARY

Results to date have shown that the production of methane from urban refuse and manure from environmental feed lots are economically viable sources of fuel gases. Demonstration plants currently in operation will establish the commercial feasibility of these processes. Studies on crop residues have shown that both corn stover and wheat straw are not viable raw materials unless the biodegradability of the material can be substantially increased. In fact, without pretreatment of these materials, the process energy requirements exceed the energy produced in the form of methane.

A mild thermochemical pretreatment of corn stover completely reverses the above situation. A substantial quantity of energy in the form of a fuel gas in excess of the process energy requirements is produced. A gas price of \$1.91/GJ (\$2.00/10^6 BTU) does not cover processing capital, operating and maintenance costs. Considerably more work on the optimization of the thermochemical pretreatment is needed. Also, an indepth analysis of crop residue availability is necessary to support this application.

ACKNOWLEDGEMENTS

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Session IV C

BIOMASS ENERGY SYSTEMS PROGRAM
AN OVERVIEW OF THERMOCHEMICAL CONVERSION ACTIVITIES

G. F. Schiefelbein
L. J. Sealock, Jr.

Pacific Northwest Laboratories*

STRACT

minishing supplies of low cost oil, natural gas,
d petrochemical feedstocks in the U. S. has
ompted government and industry to pursue the
velopment of alternative sources of clean energy.
ergy sources that are constantly renewed by the
ction of solar radiation offer a promising alter-
tive to conventional fossil fuels. The U. S.
partment of Energy (DOE) is actively involved in
e development of renewable energy sources through
search and development programs sponsored by the
omass Energy Systems Program. The objective of
is program is to develop the capability for con-
rting renewable biomass resources such as wood,
ricultural residues, animal manure, and energy
ops into clean fuels, substitute petrochemical
edstocks and other energy intensive products that
n displace fossil fuels.

cific Northwest Laboratory (PNL) has recently
en selected to provide program management serv-
es to DOE's Biomass Energy Systems Program. PNL
responsible for the technical management of 14
omass conversion projects funded by DOE's Divi-
on of Distributed Solar Technology. These pro-
cts are located throughout the United States and
ve an annual funding level of about 8 million
llars.

is presentation will provide an overview of the
omass Energy Systems Program with particular
phasis on the thermochemical conversion of bio-
ss to fuels and feedstocks via direct combustion,
rect and catalytic gasification, and indirect
quefaction. It will include a delineation of the
omass Energy Systems Program's objectives and
als, organizational structure, implementation
an and program budget.

TRODUCTION

is paper is intended to provide an overview of
ermochemical conversion technology development
tivities within the Biomass Energy Systems Pro-
am. Particular emphasis will be placed on con-
rsion technology development activities in the
eas of direct combustion, gasification and indi-
ct liquefaction via synthesis gas.

Pacific Northwest Laboratory (PNL) has recently been selected to provide program management services to the U. S. Department of Energy's (DOE) Biomass Energy Systems program. The Biomass Energy Systems Branch office is a part of DOE's Division of Distributed Solar Technology. PNL is responsible for the technical management of technology development projects directed toward the thermochemical conversion of biomass by direct combustion, gasification and indirect liquefaction via synthesis gas. Papers on specific projects within these areas of technology development will follow this introduction. Lawrence Berkeley Laboratory is responsible for the technical management of development activities on the direct liquefaction of biomass feedstocks. Direct liquefaction activities have been discussed in a previous session of this conference and will only be mentioned incidentally here.

Biomass comprises all organisms, both terrestrial and aquatic and includes renewable resources such as forests and forest residues, agricultural crop residues, animal manures, and crops grown on energy farms specifically for their energy content. Biomass production and conversion is considered a solar technology because living plants absorb solar energy and convert it to biomass through photosynthesis.

PROGRAM OBJECTIVE

The objective of the thermochemical conversion technology development activities of the Biomass Energy Systems Program is to provide a technology base for the utilization of biomass feedstocks which will result in the displacement of traditional liquid and gaseous fuels and chemical feedstocks. This objective can be accomplished by the direct combustion of biomass materials and the substitution of biomass derived fuels and chemical feedstocks for those produced from conventional sources.

Thermochemical conversion processess employ elevated temperatures to convert biomass materials to more useful energy forms. Examples include:

- Combustion to produce heat, steam, electricity, or combinations of these;

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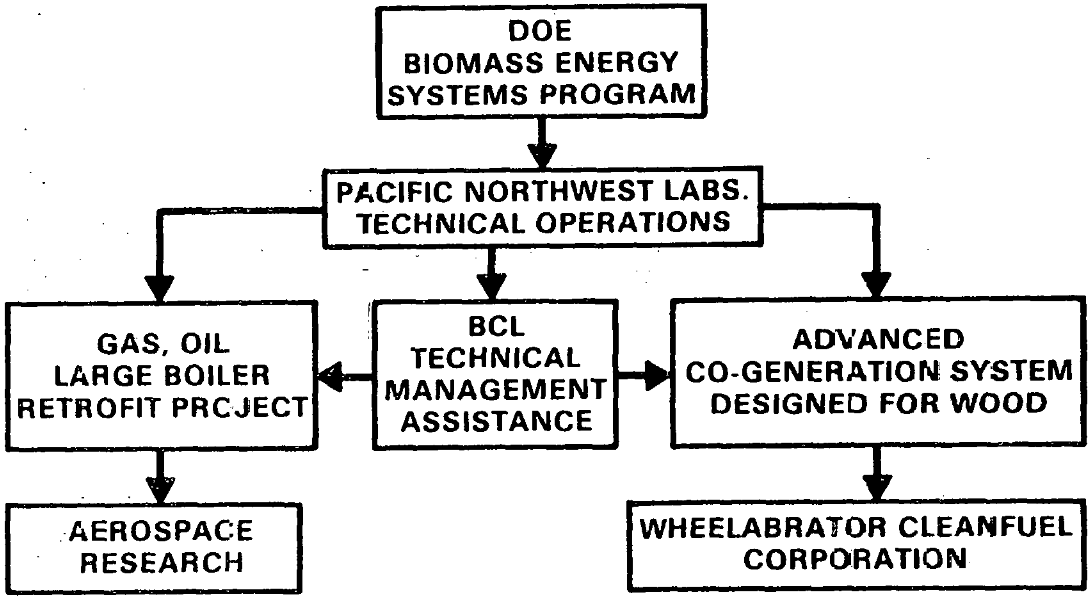

FIGURE 1. Direct Combustion Systems (Near Term)

yrolysis to produce gases (low or intermediate BTU), pyrolytic liquids and char;

- Gasification to produce low or intermediate BTU gas, synthetic natural gas (SNG), ammonia, methanol, or Fischer-Tropsch liquids and gasoline via catalytic processes; and
- Liquefaction to produce heavy oils, or with upgrading, lighter boiling liquid products such as distillates, light fuel oils and gasoline.

ROGRAM ORGANIZATION AND IMPLEMENTATION

hermochemical conversion technology development ctivities funded by the Biomass Energy Systems rogram can be divided into the following four ategories:

- Direct Liquefaction
- Direct Combustion
- Gasification
- Indirect Liquefaction Via Synthesis Gas

n the remainder of this paper we will address ainly the latter three categories as direct liqefaction activities have been discussed in a preious session.

ear Term Systems

he direct combustion of biomass feedstocks is aleady widely practiced by several industries, specially the forest products industry. Many ypes of direct combustion equipment are commerially available for this purpose. New developents in direct combustion technology are expected o have a near term impact on energy supplies hrough the utilization forest residues and other eadily available biomass feedstocks. Therefore, irect combustion technology development projects eing funded by the Biomass Energy Systems Program re categorized as near term systems development ctivities.

wo projects in the area of direct combustion techology are currently being funded by the Biomass nergy Systems Program. These projects are shown n the organization chart illustrated in Figure 1. he Aerospace Research Corporation project is irected toward the development of a wood fueled ombustor whith can be directly retrofitted to xisting oil-fired boilers. The Wheelabrator leanfuel Corporation project is a demonstration of arge scale co-generation based on wood feedstock. he scope of this project includes the design of he plant plus additional tasks such as preparation nvironmental impact statement, demonstration e tree harvesting equipment and determinaion of feedstock availability for a large facility.

Mid Term Systems

Development of biomass gasification and indirect liquefaction technologies are mid term development activities because these technologies are not expected to have a substantial impact on U. S. energy supplies for 10 to 20 years. Biomass gasification technologies can be divided into processes which produce a low BTU gas and those which produce a medium BTU gas.

Low BTU gasification technology is commercially available for most types of biomass feedstocks. Many of these commercial processes are based on low BTU coal gasification technologies and the gas produced can best be used as fuel for supplying process heat, process steam or for electrical power generation. A discussion of commercially available low BTU gasification technology will be presented in a later paper by Mr. Abu Talib of the Mitre Corporation.

The versatility of low BTU gas is limited and its use is subject to the following limitations:

- The low heating value of the gas usually requires that it be consumed on or near the production site in a close coupled process.
- Substitution of low BTU gas for natural gas as a boiler fuel usually requires boiler derating and/or extensive retrofit modifications.
- The high nitrogen content of low BTU gas precludes its use as a synthesis gas for most chemical commodities which can be produced from synthesis gas.

Medium BTU gas (MBG) offers the following advantages over low BTU gas:

- Boiler derating is usually less severe when substituting MBG for natural gas than when substituting low BTU gas for natural gas.
- MBG can be transported moderate distances by pipeline at a reasonable cost.
- MBG is required for the synthesis of derived fuels and most chemical feedstocks and commodities which can be produced from synthesis gas. The versatility of MBG is illustrated in Figure 2.

The major disadvantage of MBG is that its production by conventional means requires the use of an oxygen blown gasifier which is expensive to operate due to the cost of the oxygen.

If thermochemical conversion of biomass is to achieve its maximum potential for augmenting existing U. S. energy supplies in the mid term, the following two points will have to be considered.

1. Barring serious coal production constraints, biomass conversion will have to be cost competitive with synthetic fuels produced from coal.

2. Thermochemical biomass conversion must have an impact on the availability of liquid fuels and chemical feedstock supplies as well as supplementing gas for heating purposes.

Biomass has two potential advantages over coal. First, biomass is a renewable resource and coal is not. Second, and more important from a thermochemical conversion standpoint, biomass is more reactive than coal. It has the potential for gasification at lower temperatures, without the addition of oxygen, to produce MBG. The gasification process development activities sponsored by the Biomass Energy Systems Program are attempting to exploit this advantage. These activities are also directed toward improving the competitiveness of biomass gasification through the use of catalysts and unique gasification reactors to produce directly specific synthesis gases for the production of ammonia, methanol, hydrogen, and SNG. Success in these efforts could eliminate the necessity for external water gas shift or methanation reactors when producing these commodities. The potential elimination of the oxygen requirement and the water gas shift step are indicated by the dashed lines in Figure 2.

The gasification technology development activities of the Biomass Energy Systems Program include processes for the production of both low and medium BTU gases. However, the major thrust is directed toward improved processes for the production of medium BTU heating and synthesis gases. This appears to be the area where biomass gasification could have the greatest impact on U. S. energy requirements.

We have pointed out that the thermochemical conversion of biomass to fuels and chemical feedstocks has some potential advantages over competing coal conversion technologies. It would be inappropriate not to point out some disadvantages of biomass feedstocks. Because biomass feedstocks are usually disperse and may contain 50 to 60% moisture, there is a substantial cost associated with the collection and transportation of biomass to a central processing point. One concept which may offer a partial solution to these economic disadvantages would be to locate biomass gasification units in a disperse manner within a large forest or energy farm. The medium BTU synthesis gases produced by these units could be transported 25 to 50 miles by pipeline to a large, centrally located chemical or fuels synthesis plant. This concept offers the advantages of shorter hauling distances, eliminates transporting the moisture long distances and offers the opportunity for a large scale central synthesis plant. However, this concept requires more economic evaluation before its merit can be determined.

Gasification and indirect liquefaction technology development projects currently sponsored by the Biomass Energy Systems Program are dipicted on the organization chart shown in Figure 3. A project breakdown includes operational process development units (PDUs) at Garrett Energy Research, the University of Arkansas, Texas Tech University, Pacific Northwest Laboratory and West Virginia University.

A large fluidized bed gasification PDU is nearly operational at the University of Missouri at Rolla. PDUs are in the design stage at Wright-Malta and Battelle Memorial Institute. Supporting systems studies are being conducted by Gilbert/Commonwealth, Inc., Gorham International, Inc., Science Applications, Inc., Catalytica Associates, Inc., and Mitre Corporation/Metrek Division.

Included on the organization chart is a proposal for a large experimental facility (LEF) for the gasification of biomass. This facility, which would have a projected capacity of 300 oven dried tons per day of biomass, is still in the planning stage and has not been formally approved.

The LEF would serve as a demonstration unit for the gasification processes currently being developed at the PDU stage and would provide process information that could be used for the design of a commercial sized facility.

The geographical distribution of current biomass thermochemical conversion projects, including direct liquefaction, is shown in Figure 4.

THERMOCHEMICAL CONVERSION BUDGET

The projected Fiscal Year 1979 budget for those thermochemical conversion activities managed by Lawrence Berkeley and Pacific Northwest Laboratory is $12.5 million. The distribution of funding by major activity subelement is shown in Figure 5. This does not represent the total thermochemical conversion budget for the Biomass Energy Systems Program. A few additional thermochemical conversion activities, which are not technically managed by PNL and LBL, are funded by the Biomass Energy Systems Program.

The projected Fiscal Year 1980 Biomass Energy Systems Program budget for thermochemical conversion activities (direct combustion, direct liquefaction, gasification and indirect liquefaction) is approximately $17 million.

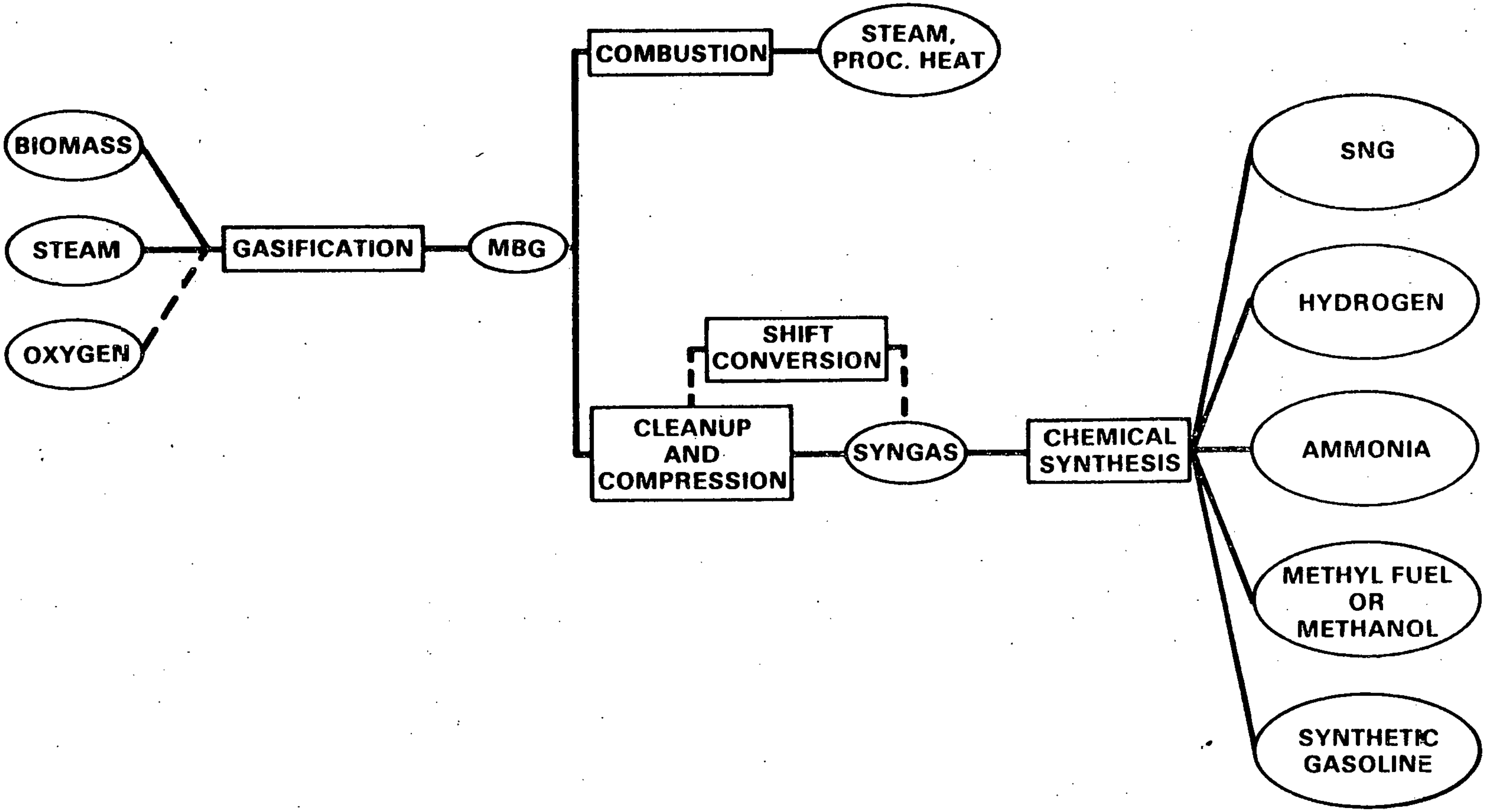

FIGURE 2. Medium BTU Gasification (MBG) - Indirect Liquefaction Technology

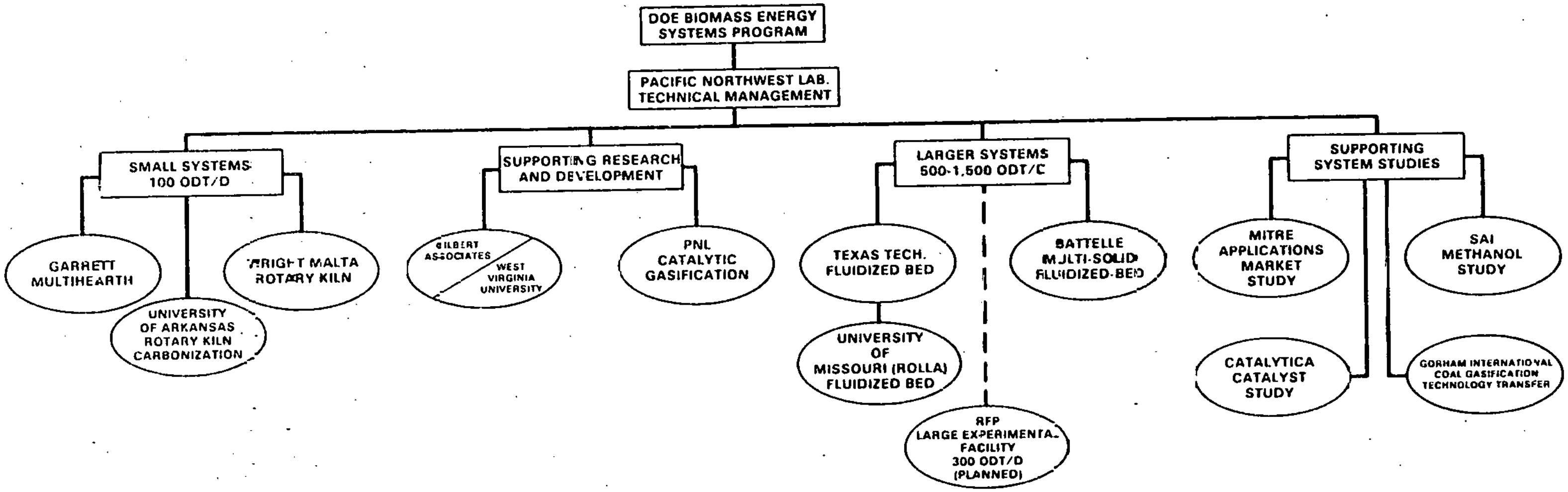

FIGURE 3. Gasification - Indirect Liquefaction Systems
(Mid Term)

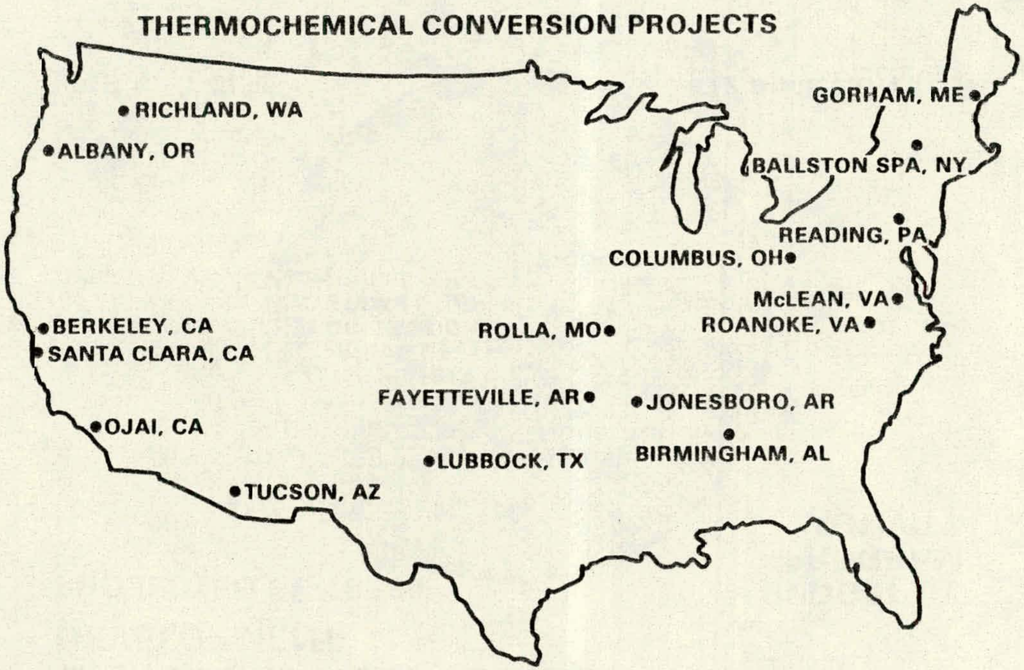

FIGURE 4. Geographic Distribution of Thermochemical Conversion Projects

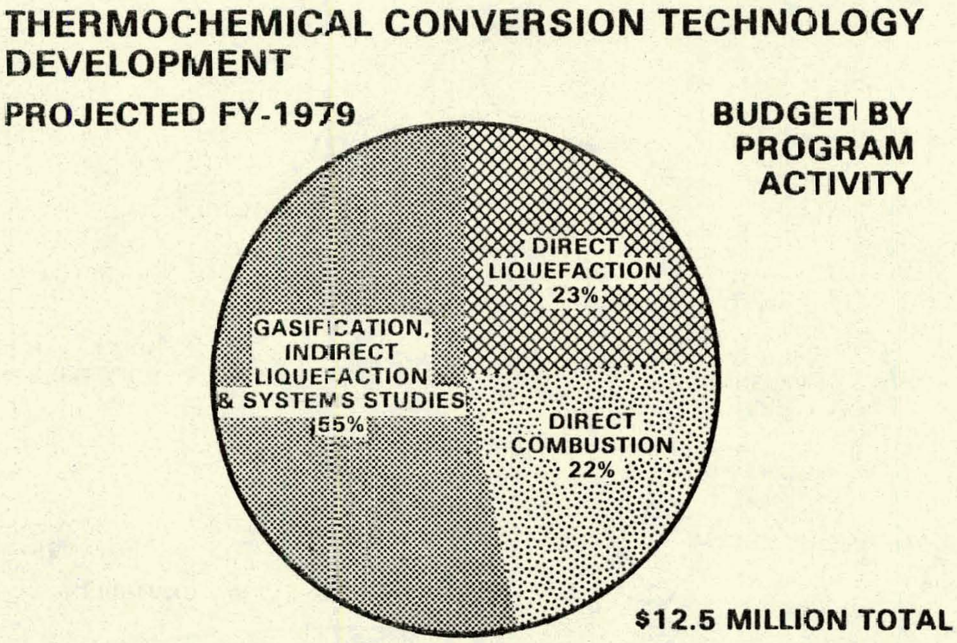

FIGURE 5. Projected Fiscal Year 1979 Budget - Thermochemical Conversion

DEVELOPMENT OF WOOD AS AN ALTERNATIVE FUEL FOR LARGE POWER GENERATING SYSTEMS

Joseph T. Hamrick
Aerospace Research Corporation

ABSTRACT

The work was carried out under DOE Contract ET-78--05-5682 starting January 3, 1978. The major task was to construct a small combustion module to demonstrate burning of wood in such a manner and at such a rate as to simulate the burning of oil or natural gas in a central station boiler and to analyze the data.

The objectives of the program were to feed pulverized green wood into a swirling stream of high temperature combustion air and determine the effect of changes in particle size and air temperature on burning rate and completeness of combustion. The projected cost of the first phase was $130,000 to build and test a small combustion module that would burn 1000 lbs (453.5 kg) of green wood per hour. The module employed a regenerative heat exchanger to preheat the combustion air to approximately 1000°F (537.78°C). The performance target was to completely burn green wood pulverized to pass a 1/2 inch (1.27 cm) screen while suspended in 1000°F air within a distance of ten feet.

A small combustion module 8 ft (2.44 m) high, by 6 ft (1.83 m) wide by 10 ft (3.05 m) long was built and tested with pine and oak wood particles which had passed a half inch screen. The pine particles were pulverized from whole tree chips of Virginia pine which were purchased from a local vendor. The moisture content was 43 percent at time of burning. The oak particles were pulverized from chips derived from trunk and limbs only. The moisture content was 35 percent at the time of burning. The combustion module was preheated with propane gas until the primary combustion air reached 500°F (260°C) and the secondary air reached 1000°F (537.8°C). The wood feed was then turned on and within 15 seconds the gas was turned off. Immediately the primary air temperature began to climb and was allowed to reach 1000°F (537.8°C) while maintaining the secondary air at 1000°F (537.8°C). Complete combustion was achieved in suspension with both oak and pine. Results of an analysis of the burning in the module indicate that a 60,000 Btu/ft^3/hr heat release rate can be achieved.

Burning experiments in the second phase of the program with larger wood particles and higher moisture content is continuing while a furnace is being built to burn wood at rates of 15,000 to 30,000 pounds (6,802.7 to 13,605.4 kg) per hour. The objective of the tests in the furnace is to determine scale effects. Knowledge of scale effects is necessary to design retrofit systems for central stations that now burn oil or gas.

INTRODUCTION

There are many steam plants in the United States that were designed specifically for and limited to the use of oil or gas. These include packaged boilers of shop erected construction as well as central power station boilers that are erected on site. In either case, the furnaces are usually characterized by lower volume and more closely spaced tubes than is normal for coal fired furnaces. There is normally no provision for ash accumulation or removal. Coal, which has only 36 to 43 percent volatile matter and an ash content from 3 to 18 percent has proven unsuitable for use in these furnaces even in the powdered form because of ash accumulation and deposits on the boiler tubes. Whole tree wood which has 79 to 82 percent volatile matter and an ash content of 1 to 2 percent appeared to have a better chance of working in the application, if it could be burned in suspension.

Dry wood in particle sizes comparable to powdered coal burns well in suspension, but the cost of drying and grinding wood to such sizes for burning is prohibitive. Calculations indicated that pulverized wood particles of a size that would pass a screen with half inch perforations would burn in 1000°F combustion air at approximately the same rate as 74 micron size powdered coal in 650°F air, and bench scale experiments with wood particles tended to confirm the indications. The dry powdery nature of wood ash throughout the burning process suggested that unlike coal, the burning wood particles would not stick to the boiler tubes even if they failed to burn completely in the furnace.

Since packaged or central station systems designed for use of oil or gas cannot currently function with solid fuels, the prospect of developing a method to retrofit them to burn wood, a renewable resource, appeared adequately attractive to design and build a small combustion module to evaluate the possibilities. Tests in the small module showed that pulverized green wood particles having passed a half inch screen can be burned efficiently in the same manner as gas or oil and the results of the first phase of the program are reported herein. The cost of pulverizing green wood with a hammermill to that size may vary from 1 to 3 percent of the electrical output, which is on the order of 2 to 3 times that for pulverizing coal. However, when the clean up costs for emissions from coal and wood burning are taken into account, the lower particulate content of wood smoke and almost complete absence of sulfur more than make up for the greater cost for pulverizing green wood. The cost of pulverizing green wood by presently available means is much greater than for dry wood, so processing and burning of green wood presents the most difficult case. While the final objective is to retrofit central station systems which may burn five to ten thousand tons of green wood per day, a nearer term goal will be to retrofit a packaged boiler which may burn 150 to 300 tons per day. To that end a furnace is being constructed in which burning at a rate of 180 to 360 tons per day can be evaluated.

DESCRIPTION OF EQUIPMENT

The basic furnace consists of heat exchangers, burner, and combustion chamber. The heat exchangers use the heat from the combustion products to preheat the combustion air to 800°F to 1000°F, the required temperature being dependent upon wood particle size and moisture content. A schematic diagram of the furnace is shown in figure 1. The primary and secondary combustion air supplies are heated as they pass over the outside of 2 3/8 inch diameter tubes through which the combustion products are ducted. Dampers in the main, primary, and secondary exhaust stacks can be adjusted from the control room to separately regulate the primary and secondary air temperatures. The flow rates of the primary and secondary air supplies are controlled by butterfly valves at the blower outlets. Wood particles from the wood bin are fed to the burner by means of a screw conveyor driven by a variable speed motor. Both air and wood feed rates may be adjusted from the control room.

Figure 2 shows a clearer view of the wood-air feed arrangement and the starting system. During start up propane gas is fed to the starting gas ring to preheat the system. When the primary air reaches 500°F and the secondary air reaches 1000°F the wood feed is turned on. The swirling primary air picks up and partially pyrolyzes the wood particles as they leave the screw conveyor moving them out into the swirling secondary air stream where combustion commences. Fifteen seconds after the wood feed is turned on, the gas is turned off. If the primary air temperature begins a sustained climb, it is a signal that a stable start has been achieved.

The wood used in the experiments was first reduced to chips and the pulverized with a Schutte Pulverizer Company, Inc. Model 1300 series wood grinder hammer mill. For the pine wood the mill operated at 1800 rpm and for the oak wood at 3600 rpm. The wood chips before grinding varied in size from approximately 1/8 in X 1/4 in X 3/4 in to 3/16 in X 1 in X 1 in (3.175 mm X 6.35 mm X 19.05 mm to 4.76 mm X 25.4 mm X 25.4 mm). After pulverizing, most particles were 1/2 inch (12.7 mm) in length but varied in the other dimension from approximately 1/32 in X 1/32 in to 1/16 in X 3/16 in (.81 mm X .81 mm to 1.59 mm X 4.76 mm). The oak particles which were pulverized at 3600 rpm were an exception. All oak particles were 1/32 in X 1/32 in X 1/2 in (.81 mm X .81 mm X 12.7 mm) or less.

INSTRUMENTATION

Temperatures at various locations in the system were measured by thermocouple and read out on millivolt strip chart recorders. Both chromel-alumel and tungsten 5 percent rhenium-tungsten 25 percent rhenium type thermocouples were used.

Air flow was measured by means of pitot tubes which measured center pipe velocity. Graphs of air flow versus center location pitot tube readings for 10°F steps and for barometric pressures ranging from 28 to 30 inches (711 to 762 mm) of mercury were prepared for use in setting air flows. Inclined manometers located in the control room were used to measure pressure differences across the pitot tubes.

The rotative speed of the wood screw conveyor was read by means of a magnetic pick up at the conveyor

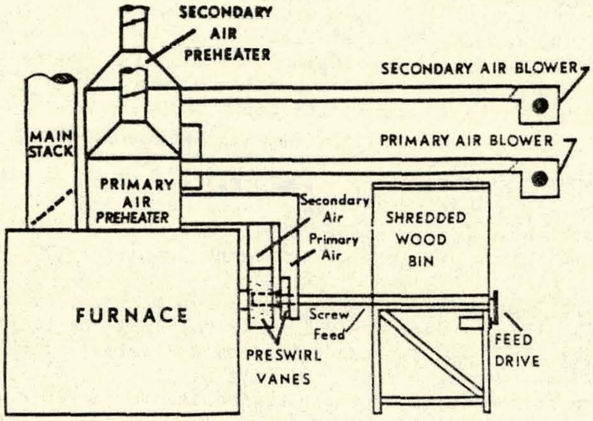

Fig. 1 - Schematic - Small Combustion Module

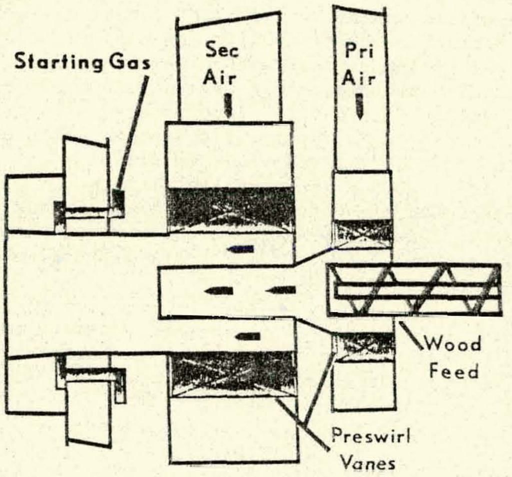

Fig. 2 - Schematic Of Wood-Air Feed

hε nd a digital read out in the control room.
oι ettings of speed were made by means of the
rive motor speed control prior to start up. In
rder to assure a more accurate average of wood
eed rate for some tests, the wood was weighed and
imed for a constant feed rate into the furnace.

n Orsat type apparatus was used to analyze flue
ases. This type analyzer uses selective absorption
f the components, except nitrogen, in liquid re-
gents.

oisture content of the wood was determined by mea-
urement of weight loss in an oven at 200°F (93.3°C)
ver a 24 hour period. It was determined that no
urther weight loss was experienced by exposing a
ample for an additional two 24 hour periods.

he heat value determinations of the wood specimens
ere made using a Parr adiabatic bomb calorimeter
n accordance with ASTM D20k5-66 by personnel of
roehling & Robertson, Inc. in their Richmond,
irginia laboratory. That laboratory also furnish-
d ultimate analyses of wood samples.

TABLE 1 - RESULTS OF TESTS WITH GREEN OAK AND GREEN PINE WOOD

	OAK	VA. PINE	VA. PINE
REEN PERFORATION DIAMETER, IN.	1/2	1/2	3/8
OUNT OF EXCESS AIR, %	0	9	37
IMARY PRESWIRL VANE SETTING	11°	11°	11°
CONDARY PRESWIRL VANE SETTING	9°	9°	9°
IMARY AIR FLOW RATE, CFM	395	390	383
CONDARY AIR FLOW RATE, CFM	325	380	476
OD FEED RATE LBS/HR	811	994	900
R TEMPERATURE, °F	88	74	66
OD TEMPERATURE, °F	68	71	66
IMARY PREHEATER STACK, °F	1686	1660	404
CONDARY PREHEATER STACK, °F	972	931	786
IMARY SWIRL CHAMBER, °F	965	1020	692
CONDARY SWIRL CHAMBER, °F	1000	1027	1002
LL BELOW BURNER, °F	2083	2031	2069
IMARY AIR-WOOD MIXTURE, °F	628	736	550
ILING OF FURNACE, °F	2260	2180	2160
LLULAR MOISTURE CONTENT, %	35	43	49
OD HEAT VALUE, BTU/LB	8096	8604	8600
SIDUE, AFTER COOL DOWN, %	.8	.4	3
$_2$ IN EXHAUST, %	15.9	15	15.5
IN EXHAUST, %	1.4	3.9	3.9
IN EXHAUST, %	3	1.5	.9
IN EXHAUST, %	79.7	79.6	79.7
OKE COLOR	BLACK	BLACK	GRAY
SE RATE BTU/FT^3/HR	10,930	12,203	9407

EXPERIMENTAL RESULTS

A total of fifteen experimental runs were made in which an average of 500 lb (226.8 kg) lots of wood, pulverized and screened were burned in periods varying from 27 to 46 minutes after the system was brought up to starting temperature by propane gas. One run was made in which a 1500 lb (680.4 kg) lot of wood was burned in 90 minutes. Two of the runs were made with red oak wood with the remaining runs being made with Virginia Pine. The primary and secondary air temperatures varied from an average of 650 to 1020°F (343.3 to 548.9°C). Typical tests will be described for pine and oak and for the 90 minute run with pine in which a range of combustion air temperatures were used.

Thirty Minute Tests With Virginia Pine With 43% Cellular Moisture

For a typical test for which results are shown in Table 1, the primary air temperature was brought up to 500°F (260°C) and the secondary air to 1000°F (537.8°C) with propane gas before starting the wood feed. Fifteen seconds after the wood feed was started, the propane gas supply was shut off. Approximately a ten minute period was required to stabilize with a constant wood-air feed. In a 30 minute test using 66 percent excess air with 826°F (441.1°C) primary air and 897°F (480.6°C) secondary air there was an approximate 36 inch (914.4 mm) diameter erosion pattern on the refractory opposite the burner ten feet (2.54 m) away. The refractory was eroded to an average depth of approximately 1/4 inch (6.35 mm). For runs in which the combustion air was 900 to 1000°F (482.2 to 537.8°C) and there was zero to thirty percent excess air, there was no significant erosion and there was very little ash residue in the furnace. The quantity varied from 2 to 4 pounds (.907 to 1.81 kg) depending upon the amount of excess air. The greater the amount of excess air the smaller the amount of residue. Most of the residue was found at the bottom of the door which formed the wall opposite the burner. Determination of burning behavior was difficult because of inability to view the flame. In an attempt to determine if particles had fallen to the floor and burned, liquid carbon dioxide was injected directly into the furnace after shut down at the end of one test. The objective was to quench any particles that may have been in the process of burning and to maintain a blanket of carbon dioxide in the furnace until it was cool enough to remove the end door. The results were that there appeared to have been no wood residue on the floor at the end of the test. The usual amount of ash residue was found at the bottom of the door.

Ninety Minute Test With Virginia Pine With 43% Cellular Moisture

The objective of this test was to start off with high temperature combustion air and reduce the temperature in successive steps to a point where combustion became erratic. Although the wood particles used in the test had a cellular moisture content of 43%, added surface moisture increased the total to 51.8%. The test was carried out with 10 percent excess air. Because of the high moisture content, starting failed with the primary air temperature at 500°F (260°C) and secondary air at 1000°F (520°C). After the primary air temperature was raised to 700°F (371.1°C) and the secondary to 1125°F (607.2°C), the start was made in a routine manner. The primary air was then adjusted to 968°F (520°C) and the secondary to 956°F (513.3°C). These temperatures were stepped downward at approximately fifteen minute intervals until the final temperature of the primary air was 615°F (323.9°C) and the secondary air 679°F (359.4°C). In the tests at these and the higher combustion air temperatures combustion was stable. After the furnace had cooled and the door was opened, there were approximately 2.5 cubic feet (.0707 cubic meters) of charred wood and insulation debris piled up at the door and a crater approximately 2 feet (.609 m) in diameter had been eroded away in the refractory on the door. The center of the crater was approximately 5 inches (127 mm) below the centerline of the burner and the refractory was eroded away to the sheet metal base. The temperatures on the wall above and below the burner remained approximately constant across the range of combustion air temperatures.

Thirty Seven Minute Test With Red Oak With 35% Cellular Moisture

Two tests were made with pulverized red oak. As the oak particles were shredded to a smaller size than for the pine, it was fluffier and less dense. Therefore, the feed rate was proportionately lower for the same speed of the screw conveyor. The starting procedure was the same as with the pine. The data for one of the red oak tests with 0 percent excess air is shown in Table 1. Test results for the two tests with oak wood were not significantly different from those with pine wood, and the erosion of the refractory material on the door was not measurable.

Tests With Virginia Pine With 49% Cellular Moisture

Tests were made with particle sizes passing 3/8, 1/2, and 3/4 inch (9.53 X 12.70 X 19.05 mm) screens. With a normal start, the particles passing a 3/4 inch (19.05 mm) screen failed to provide sustained combustion. Those passing a 1/2 inch (12.70 mm) screen provided sustained combustion after a normal start and burned in the same manner as the propane gas. However, only approximately 85 percent of the particles burned in suspension before striking the opposite wall 9.3 feet (2.835 m) away from the burner. Of the particles passing a 3/8 inch (9.525 mm) screen, approximately 97 percent burned in suspension. The data for this test are provided in column three of Table 1. One test was made with the pine with 49% cellular moisture after pulverizing to pass a 1/2 inch (12.70 mm) screen and exposure to the open air for several days. The average moisture content was reduced to 31 percent, and the particles were burned with 10 percent excess air at 1000°F (537.8°C). The most significant result of the test was that approximately the same quantity of particles failed to burn in suspension after air drying as for equal size particles with no drying.

DISCUSSION OF RESULTS

It has been shown that, with the techniques being used, green wood can be burned rapidly and in much the same manner as gas or oil. It can be economically pulverized to a particle size that burns read ly in 800°F (426.7°C) to 1000°F (537.8°C) combustic air. The tests to date have provided information for modification and further testing of the small module and for design of a larger furnace to evaluate scale effects.

Factors which merit further consideration and evaluation include particle size optimization, cellular moisture content, combustion air temperature, combustion efficiency, maximum temperature, and heat release rate.

Particle Size Optimization

While it has been demonstrated that particles passing a screen with half inch diameter perforations will burn and combustion is easily sustained, for maximum economy it will be necessary to establish combustibility for various particle sizes with varying moisture content. Completion of combustion within a given trajectory appears to depend primar ly upon the cellular moisture. Water was added to the half inch pine particles of Table 1 which contained 43 percent cellular moisture, to raise the moisture content of the wood to 49 percent. This moisture addition did not result in the adverse effects experienced with half inch particles containing 49 percent cellular moisture. Even in the case of partially drying the particles with 49 percent cellular moisture to a 31 percent overall moi ure content, the center core of the larger particl apparently retained most of the cellular moisture and failed to burn more completely than the undrie particles in the 9.3 ft (2.835 m) distance. In a working situation, a rapid method of determining moisture content could result in the ability to practice selective pulverization and accurate programming of feed rate to maintain a given Btu inpu to the furnace. For example, with a 9.3 ft (2.835 path length the oak wood of column 1, Table 1, whi was pulverized much finer than the pine of column exhibited no better burning characteristics than t pine from standpoint of complete combustion, but i would have shown up better in a chamber short enou for some of the longer burning pine particles to impact before completely burning.

Cellular Moisture Content

The amount of cellular moisture in green wood vari with age of the tree, species, location, ground mo ure, and the time of year. For the wood used in t program it varied from 35 percent for red oak to 5 percent for Virginia pine. Taking into account, t latent heat of vaporization of the water in the wo the cost per million Btu was estimated as shown in Table 2. It can be seen that the purchase of gree wood on a weight basis leaves much to be desired, and that purchase on a basis of Btu per pound as d livered may be more equitable. Air drying of

TABLE 2 - COST OF WOOD ON BTU BASIS
(980 BTU PENALTY PER LB OF MOISTURE)

WOOD SPECIES	MOISTURE CONTENT GREEN %	BTU PER LB	COST PER MM BTU AT $18.50/TON,$
VA. PINE	50	3815	2.42
VA. PINE	43	4481	2.06
RED OAK	35	4919	1.88

can present a significant storage problem as the moisture content in wood reduces to only 18 to 25 percent after a year of air drying. Therefore, burning the wood green appears to be the most practical approach. In the planting of trees for fuel, consideration of the maximum cellular moisture for the various species in a given location may be an important factor in species selection.

Combustion Air Temperature

Chemical reaction rate as noted by Arrhenius in 1889 varies with absolute temperature to the 6th to 8th power, and is the key to rapid burning of green wood particles. A comparison of wood pulverized to pass a half inch screen with coal pulverized to 74 microns indicated that the preheated secondary air temperature for wood should be 1000°F (537.8°C) as compared to 650°F (343.3°C) for coal and that the preheated primary air for wood after mixing with the wood should be 500°F (260°C) compared to 200°F (93.3°C) for the primary air and coal mixture. The conventional mixing process for pulverized coal and air is to mix the two far enough upstream of the burner to raise the temperature of the coal particles to that of the air. With the wood in this application, the mixing length ahead of the combustion zone was set at approximately fourteen inches, the objective being to partially pyrolyze the wood particles at the surface and keep the products of pyrolysis above 500°F (260°C) as they move into the combustion zone. A longer mixing length would result in greater heat transfer to the wood and a cooler mixture. It is assumed that the products of pyrolysis ignite immediately upon reaching the combustion zone, and that with heat radiation to the remaining portion of the particle, continued rapid combustion occurs. The thermocouple used to sense the temperature of the air-wood mixture was located three inches downstream of the wood feed outlet. At that point, the measured temperatures as shown on Table 1 were 428°F (331.1°C) for the oak, 736°F (391.1°C) for the 1/2 inch (12.7 mm) pine, and 550°F (287.8°C) for the 3/8 inch (9.5 mm) pine. For all three, suspension burning was successful.

Combustion Efficiency

ount of CO in the exhaust products, which in on to unburned char is a measure of combustion efficiency, appears to be at a minimum with 30 to 40 percent excess air. The amount and density of the smoke is also less. The amount of smoke could also be affected by vane setting, but so far, the effect of the vane setting and direction of swirl in the preswirl chambers has not been evaluated. With the settings as shown in Table 1 the primary and secondary air swirl in opposing directions. With zero excess air for the oak wood as shown in Table 1, the CO content of 3% is unacceptable as is the content of 1.5 percent CO for the 1/2 inch (12.7 mm) pine with 9 percent excess air. With 37 percent excess air and 3/8 inch (9.5 mm) pine the .9 percent CO may be as low as can be obtained with 49 percent cellular moisture and the primary air temperature of 692°F (366.7°C). Further experiments will be necessary to determine the effects of both moisture and primary air temperature on combustion efficiency for a given particle size.

Maximum Temperature

Wall temperatures as high as 2200°F (1204.4°C)and ceiling temperatures as high as 2500°F (1371.1°C) were measured with thermocouples. Difficulty was encountered in measuring flame temperatures. Theoretical temperatures with 25 percent excess air range from 3000°F (1648.9°C) with 600°F (315.6°C) combustion air to 3170°F (1743.3°C) with 980°F (526.7°C) combustion air. With no excess air the theoretical temperatures are approximately 400°F (222.2°C) higher. It is indicated by the erosion of the refractory that the actual flame temperatures approach the theoretical. The computed temperatures bracket the 3200°F (1760°C) melting point of the refractory, and it is conjectured that the flaming char, upon impacting, melts the refractory and is quenched at the same time. The quenched char and melted refractory then fall to the floor.

Heat Release Rate

The erosion patterns on the refractory indicate that the wood air mixture leaves the 10 inch (254 mm) diameter burner with a vortex motion and expands to approximately a diameter of two feet (700 mm) in the 9.3 ft (2.83 m) length, dropping down only approximately 5 inches(127 mm) in the process. If it is assumed that the burning takes place in a 2 foot (700 mm) diameter cylindrical zone 9.3 feet (2.83 m) long the heat release rate in the zone at the maximum wood flow rate of 1272 pounds (577 kg) per hour would be greater than 150,000 Btu/ft^3/hr. Based on these observations it is expected that even though the maximum heat release rate based upon total chamber volume was 13,445 Btu/ft^3/hr, the 30,000 to 60,000 Btu/ft^3/hr needed to retrofit existing oil and natural gas boilers can be achieved.

SUMMARY OF RESULTS

1. Tests have shown that by using 1000°F (537.8°C) primary and secondary air, pulverized green wood particles which have passed a half inch (12.7 mm) screen can be burned in suspension in the same manner as natural gas.

2. For green wood delivered by a 4 inch (101.6 mm) diameter screw conveyor at a rate of 994 lbs/hr (451 kg/hr) into 1020°F (548.9°C) combustion air with 9 percent excess air, combustion was completed in suspension within a distance of 9.3 ft (2.83 m).

3. The overall heat release rate in the small combustion module was 13,445 Btu/ft^3/hr. Results of an analysis of the burning in the module indicate that a 60,000 Btu/ft^3/hr heat release rate can be achieved.

FUTURE PROGRAM

Tests are currently being carried out in the small combustion module which has been modified to produce higher heat release rates. Both tulip poplar and pine woods are being burned. It is planned to make further modifications to the small module to increase the path length by inserting a 2 ft (609.6 mm) diameter 10 ft (3 m) long metal cylinder between the burner and the existing furnace. An annulus around the metal cylinder will convey the secondary combustion air from the existing heat exchanger to the swirl chamber at the burner. This arrangement will more realistically approach one that is likely to be used in a retrofit situation where 550 to 650°F (287.8 to 343.3°F) preheated combustion air will exchange heat with the combustion products to be raised to 800 to 1000°F (426.7 to 537.8°C). A large furnace which has a combustion air cooled cylindrical shell around the combustion chamber is being constructed to evaluate burning of green wood at rates of 15,000 to 30,000 pounds/hr (6,802.7 to 13,605.4 kg/hr). The primary objective is to determine scale effects. Several species of wood of different particle sizes will be evaluated.

ADVANCED SYSTEM DEMONSTRATION FOR THE UTILIZATION OF BIOMASS AS AN ENERGY SOURCE

S. Brown Milligan

and

Otis J. Manar, III

I. INTRODUCTION

A. Purpose

This report summarizes progress achieved and the results of our project study for a biomass conversion facility. It has been prepared for the Department of Energy, Fuels from Biomass Systems Branch, Washington, D.C., by the Wheelabrator Cleanfuel Corporation of Washington, D.C., and The Rust Engineering Company of Birmingham, Alabama. This report has been prepared under DOE Contract EG-77-R-06-1036 and presents the results of our extensive investigation into the technological and economic viability of a biomass fueled plant which generates steam and/or electricity. It includes the results of several special studies which appraised factors relating to biomass availability, environmental impact, and site selection.

B. Summary

Under this contract, Wheelabrator Cleanfuel Corporation served as the Contractor, and The Rust Engineering Company performed the Program Management, Site Selection, and Engineering Development. Other portions of the contracted work were assigned to the Wood Energy Institute for a Biomass Study, and to Dartmouth College for an Environmental Impact Study.

The objective of this contract was to study the feasibility of a demonstration facility to be built by a non-government entity for the conversion of biomass to energy. It was to show that a significant portion of the present industrial need for other sources of energy could be accommodated through the use of biomass fuel and an integrated conversion facility. The facility was to demonstrate that biomass could produce steam and electricity at a competitive cost. As a result of this successful demonstration, the construction of additional biomass conversion units for commercial utilization was to be encouraged.

The basic requirements of Wheelabrator's contract with the DOE include the following:

1. Find a constant source of biomass fuel to provide 1,000 tons (dry basis) of fuel daily to a plant site.
2. Select a system for biomass harvesting, collection, and transportation to assure a constant source of fuel to the plant.
3. Estimate costs for biomass harvesting and determine the delivered fuel cost at the site.
4. Select a suitable site for a biomass conversion facility to produce steam and/or electricity.
5. Engineer and design a biomass conversion facility plan.
6. Estimate capital cost, operating costs, and product prices.
7. Prepare an Environmental Impact Statement for the demonstration system.

Wheelabrator Cleanfuel was selected to perform this work by the DOE in September, 1977. A letter contract was issued October 1, 1977, and definitive contract was signed on March 16, 1978. The project as conceived by the DOE consists of three phases to be completed in 64 months. The overall project schedule is as follows:

Overall Project Schedule

PHASE I. System Design. 19 months. Completion by June 31, 1979.

EVALUATION PERIOD. 4 Months after Phase I.

PHASE II. Construction and Start-Up. 29 months.

PHASE III. Operation and Evaluation. 12 months after Phase II.

The Phase I work has now been completed and the documents are being submitted to the DOE by Wheelabrator Cleanfuel Corporation. Further work has not been authorized.

This design package is based upon a specific site in the State of Maine. However, the design could serve as a model for other biomass conversion facilities located anywhere biomass is abundantly available.

C. Biomass Energy Flow Concepts

1. Energy Source

The energy source is biomass--which includes forest residues and forest products mill wastes. The forest residues consist largely of standing timber that has no commercial value at present--rough-and-rotten wood, non-marketable species, dead and diseased trees, thinnings and culls, overage stands, and other biomass presently unused by the forestry industry. The second source is mill residue, which includes bark from paper mills; trimmings and sawdust from sawmills; shavings and trimmings from furniture plants; and other industrial and commercial wood waste materials.

All of these materials represent energy that is presently underutilized. Additionally, the forests continue to produce an abundant amount of this material each year. If this material were removed, it would improve the remaining growing stock. Our forests are renewable energy sources that can provide a continual supply of fuel. Thus, the use of biomass as an energy source is a viable way to help us deal with the energy crisis that threatens our society.

2. Harvesting and Collection

The existence of the biomass supply is beyond question. The methods available for economical harvesting and collection of biomass from the woods are under investigation and development. Mechanical equipment is now available that reduces manual labor and increases productivity such that biomass can be competitive. Many types of machines knows as "fellers", "feller-bunchers", and "grapple skidders" and "forwarders" have come on the market to assist the logger in his tasks and greatly increase daily output.

These methods have been studied and reported upon as a major effort of this project, and are covered in detail in a separate report.

3. In-Woods Processing

The economics of delivering biomass to the plant requires consideration of every available approach that can reduce costs and improve efficiency. The concept of in-woods processing can now be implemented by machines that take whole trees and reduce them rapidly to chips. Some forestry operations now utilize in-woods chippers along with mechanical fellers, feller-bunchers, skidders, or combinations of this type equipment to provide a complete "chip-making" operation on wheels that can be re-located from place to place in the forest. The chips produced are then loaded into a truck or van for transport to the plant, where they can be handled more easily than logs and can be routed directly to a furnace for combustion.

Other approaches to in-woods processing include topping and de-limbing to produce shortwood (lengths up to 8 or 10 feet) and longwood (tr length logs, with tops removed. These approaches require transport as roundwood with appropriate unloading, handling, and chipping at the power generation facility. Economics favor in-woods chipping and this approach has been incorporated into the design of the power generation facility.

4. Transportation

Two basic approaches to the transport system have been explored--truck and rail. Trucks will invariably be used for the trip out of the woods. Railroads offer possibilities for low-cost transport, especially if distances are long enough to warrant an extra loading/unloading operation. Mechanical means of loading and unloading are essential to the transportation system and have been incorporated into the design based on study results.

D. Site Selection

A comprehensive site selection study revealed a number of locations in the State of Maine that offered desirable characteristics for selection as the prime site for the proposed facility. The initial list of potential sites included 22 localities where a large industrial concern could become a customer for energy produced by the facility. This list was reduced to five prime sites by careful analysis of the most viable elements required to meet the precise conditions set forth in the contract. The five final sites were evaluated and ranked in numerical order.

As a result of this investigation, the site designated as Site 1--Westbrook, Maine, was selected as the site to be used for the basic study.

E. Description of Plant and Process

The plant is designed as a nominal 50-megawatt power plant cogenerating steam and electricity and using biomass as the principal fuelstock. Oil is available as a back-up fuel and a small amount of oil is burned continuously to keep the lines, pumps, and control system ready for immediate switchover if necessary.

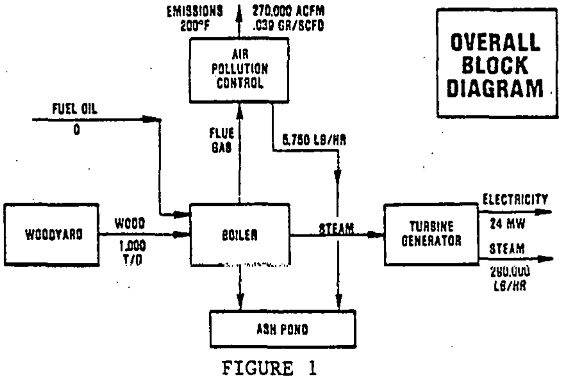

FIGURE 1

Js: the cogeneration principle, the plant produ[illegible] both steam and electricity for sale to ppropriate customers. High pressure steam is enerated and passed through a turbine-generator or electrical generation, with extraction of team at lower pressure levels for process and eating uses. The boiler and turbine-generator ave been matched to the industrial process steam equirements; electrical generation will vary lightly with steam demand. This design approach chieves a high level of efficiency.

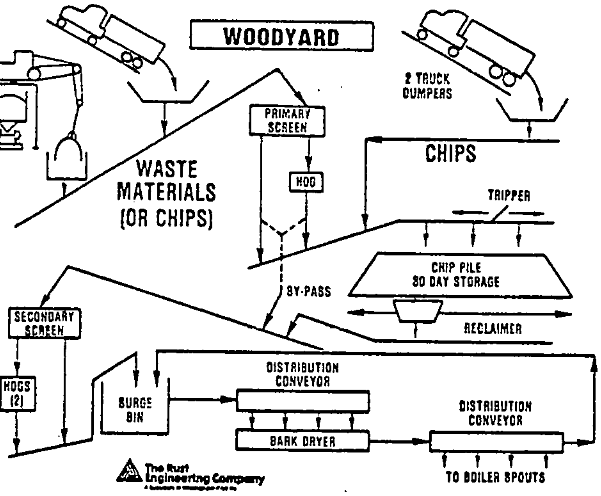

FIGURE 2

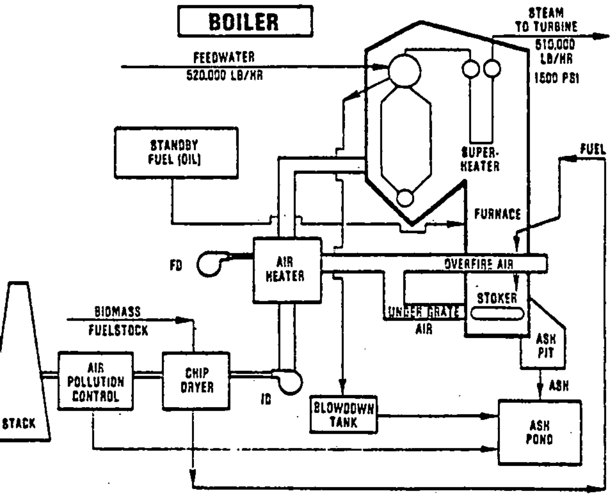

FIGURE 3

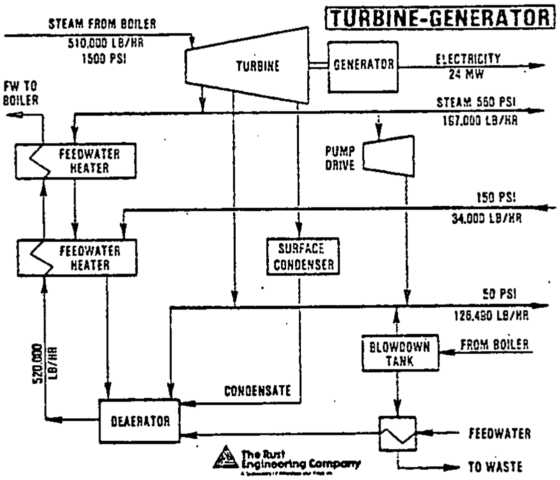

FIGURE 4

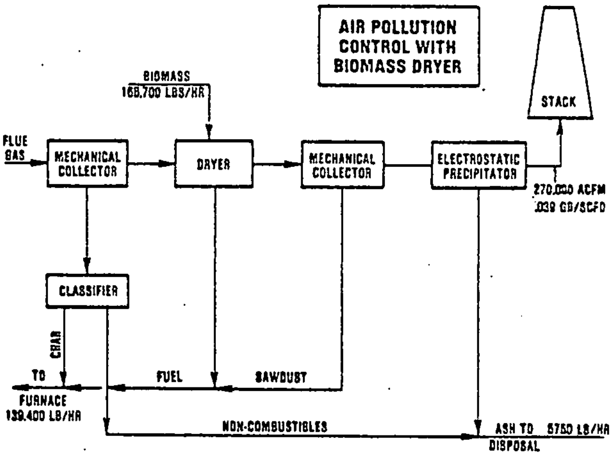

FIGURE 5

The plant consists essentially of biomass receiving, handling, and storage facilities, and a boilerhouse with its associated facilities. The biomass handling facilities include truck dumpers for unloading trailers loaded with whole-tree chips and mill residue, along with the necessary conveyors and processing equipment for size reduction and storage of 30-day supply of fuelstock. Fuel reclaimed from storage is conveyed to the boilerhouse, where it is dried and delivered to the furnace. The boiler generates steam at 1500 psig and 925°F for delivery to the turbine-generator. The hot exhaust gases are used as a drying medium for the incoming fuelstock as well as to pre-heat the combustion air. Exhaust gases are processed for air pollution control to meet all applicable regulations. A double-basin ash pond is provided for receiving ash sluiced from the furnace. At any given time, one section of the pond is being used to receive ash, while the other section is drying and settling, which will be disposed of in a landfill.

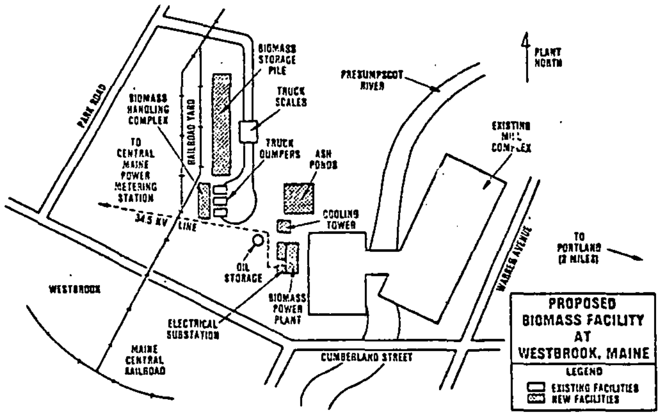

FIGURE 6

The overall external appearance of the facilities will be that of an industrial complex designed in keeping with modern industrial architectural practice.

F. Environmental Impact Statement

An Environmental Impact Statement has been prepared for the proposed facility as required by applicable laws and regulations. The study revealed that the biomass conversion plant will have environmental impacts, both beneficial and adverse. These impacts fall into four general categories, effects upon the environment due to (1) plant construction, (2) plant operation, (3) biomass harvesting, and (4) the transportation of biomass.

Plant construction would provide approximately 1,000 jobs, direct and indirect, over the two-year construction period. There would also be a significant increase in the ambient noise levels in the area.

Plant operation would change the overall emissions to the atmosphere from the Westbrook complex, decreasing particulates and sulfur dioxide considerably, and increasing the emission of nitrogen oxides and carbon monoxide. The water quality in the Penobscot River would be improved, Landfilling of the residual ash should not pose any hazards to the environment.

The environmental effects of harvesting are dependent upon forestry management. With the use of sound harvesting systems and well-managed silvicultural practices, both the economic and aesthetic values of the residual forestry stands can be improved.

The transportation of biomass supplies to the plant will increase the traffic in the area, and will tend to increase noise levels. With good dispersal of traffic throughout the working hours, this effect will not be significant.

The project will contribute to several goals our National Energy Plan, including the repla ment of approximately 766,000 barrels of oil annually. It will also provide a significant demonstration of appropriate technologies and environmental standards for the development of wood as a sustainable, renewable energy source for industry.

G. Concept of Operation

The facility is designed for continuous operation, 24 hours a day, 7 days a week, for 340 days a year. The shut-down period is for scheduled maintenance and repair to assure continuous operation for the following scheduled period.

It will operate upon a fixed fuel supply rate of 1,000 oven dry tons per day of biomass fuelstock, converting the energy in the fuel by direct combustion into steam for process use and electrical generation.

The operation will be controlled to balance the production of electrical energy such that the steam user will receive his needs first, and the remainder of the energy will be converted to electricity for the utility company. Under average conditions, approximately 258,000 lb/hr of steam will be sold to the industrial user while approximately 20 megawatts of electricity will be sold to the utility grid. This balance will vary from season to season and at various times due to the fluctuations in steam requirements.

II. KEY RESULTS AND ECONOMIC EVALUATIONS

The work under this contract has been completed, and the report for the program is now finalized fo presentation to the DOE in early June of this year, ahead of the target date, and within budget.

A. Key Results

1. The Biomass Availability Study has concluded that within a 50-mile radius of the model site selected for the conversion facility that there is sufficient biomass fuelstock not presently utilized to sustain the operation over a long-term period.

2. The Environmental Study completed for the model site indicates that no significant major impacts will result either in the harvesting and collection of fuelstocks in the forest or at the conversion site providing the harvesting procedures as recommended are observed, and the plant is constructed in accordance with the conceptual design.

3. Based on the harvesting and collection system recommended and the conceptual design for the conversion facility, the cost of the energy products-process steam and electric power, the use of biomass as a feedstock is a competitive, viable alternative fuel source.

B. Economic Evaluations

Tl broad areas were evaluated in determining the overall costs for the facility. This basis for evaluation was divided between the total investment costs and the total operating costs.

The total investment costs for the proposed biomass conversion facility are $64,500,000. The components of this cost are:

Total installed capital, with engineering	$55,000,000
Start-Up, Checkout and Testing	800,000
Land	300,000
Working Capital	1,140,000
Interest during construction	6,860,000
Bond sale cost	400,000

The annual operating costs totaled the following information:

Fuel cost	
Biomass	$ 8,485,600
Oil	446,500
Labor	
Operating	570,400
Maintenance	119,700
Maintenance services	2,017,200
Local Tax	1,552,300
G&A expenses	289,300
Total	$12,790,900

Other evaluations considered the economics of the Composite Fuel Cost, Annual Average Revenue Required, Product Sales Price Range, and the Product Price Combinations That Would Generate The Required Annual Revenue.

One additional evaluation compared the cost of generating electricity and steam based on a 16.1 percent return on equity. This produced the following data:

Electricity $/Kwh	Steam $/Million Btu	
	Biomass	Coal
.055	4.38	4.33
.040	5.29	5.24

This clearly demonstrates the economic viability of biomass conversion costs as compared to those costs associated with the use of coal.

III. FUTURE PLANS

The work completed under this contract to date needs to be continued. A widespread interest is becoming apparent in the industrial sector to finance and construct new plants utilizing biomass as a feedstock. An extension of our work on this contract is needed to develop the final detail design and to nurture and promote industrial involvement in the program. Wheelabrator Cleanfuels Corporation and The Rust Engineering Company are prepared to assist DOE in this important effort.

NOTES

THERMAL CONVERSION OF BIOMASS
IN A
ROTARY KILN SYSTEM

Henry H. Hicks
Project Director

Jim L. Turpin

James R. Kimzey
Associate Director

DOE Contract No. EG-77-5-04-4096

Starting Date: October 1, 1977

UMMARY OF PRESENTATION

escription of Task

he task involves the justification for the use of rotary kiln as a device for biomass conversion, he goals to be achieved by a rotary kiln biomass rocessing system, and the justification for the xpenditure of funds for the study of such a sysem.

bjectives/Cost and Performance Targets

he objectives of the project were:

1) to develop engineering design data for rotary kiln pyrolytic converters.
2) to make an engineering evaluation of the patented rotary kiln processing concept.
3) to obtain pyrolytic conversion data for the biomass materials available in Arkansas and the Ozark Plateau Region.

he total cost of the project was $440,000. Obective (1), the scale model kiln phase, cost oughly $119,000 and is about 90% complete. Obective (2) of the project required about $216,000 nd is about 60% completed. Approximately $79,000 as expended on objective (3) and this task is bout 50% complete. The remaining $26,000 of the roject funds were for a Washington-based manageent consultant firm.

pproach

he approach to the problem was to divide the proect into three phases to address the three respective objectives. Phase (1) was to be accomplished by designing, constructing, and operating scale model kiln with various forest and agricultural waste feedstocks. Phase (2) was to be ccomplished by utilizing an existing prototype iln for testing and data acquisition. A laboratory was developed for the analysis of feedstocks nd products and for basic pyrolytic studies to ccomplish phase (3).

Key Results/Accomplishments

Phase (1) of the project is essentially complete. The design and constuction of the experimental kiln was successfully completed, demonstrating a minimum of design scaling problems.

Roughly 100 experimental kiln runs were made with wood waste feedstocks from several industries, with agricultural waste feedstocks including rice hulls, cotton gin waste, and nut hulls, and with industrial waste feedstocks from television and air cleaner manufacturing concerns. Successful kiln operation and data collection were realized with each of the feedstocks.

Phase (2) of the project was hampered by mechanical difficulties. These problems have been addressed and the evaluation of the 40tpd prototype unit is continuing.

Phase (3) provided analytical support for phases (1) and (2) of the project, but essentially no basic pyrolytic studies were made.

Future Plans

It appears that the economics are very favorable for the use of a rotary kiln biomass processing system for disposal of agricultural and forest waste products with the production of charcoal and process heat and/or steam. This is especially true in Arkansas and the Ozark Plateau Region where there are many small industries producing significant quantities of biomass waste materials while at the same time requiring process heat and/or steam in their processing scheme.

Plans are to disseminate the project results to potential users. It would be desirable to continue the investigation in order to study other variables of the system, to encourage the development of charcoal as a fuel, and to further develop the rotary kiln process.

THERMAL CONVERSION OF BIOMASS
IN A
ROTARY KILN SYSTEM

Henry H. Hicks
Project Director

Jim L. Turpin

James R. Kimzey
Associate Director

DOE Contract No. EG-77-5-04-4096

Starting Date: October 1, 1977

DESCRIPTION OF TASK

A rotary kiln system is a viable processing alternative for the thermal conversion of biomass materials. It is a flexible system, accepting a wide variety of agricultural, forestry, and municipal wastes as feedstocks. The rotary kiln can be operated in various modes, ranging from complete incineration to pyrolysis of the feedstocks. Kiln products may be completely oxidized flue gases, gaseous and liquid fuels, solid char fuel, or a combination of these.

Economical pollution control is achieved by these systems in a processing scheme which provides a "clean" atmospheric discharge of gases as the solid waste materials are eliminated. Finally, rotary kilns are designed and constructed in a wide range of sizes with only minimal scale-up problems.

The goals of the processing system to be discussed are (1) the elimination of solid waste materials in a non-polluting manner, (2) the utilization of off-gas process heat, and (3) the production of $75 per ton charcoal with an acceptable rate of return on the investment.

The immediate background for the project to be presented here is that a small Arkansas wood products company (A&P Coop Company) had developed and patented a process for making charcoal in a rotary kiln converter. The company approached the University of Arkansas for assistance in evaluation of their invention. A preliminary investigation of the process made it evident that an adequate engineering evaluation would require external financial support.

There were several incentives for the proposed study. The investigation should result in an integrated processing scheme which is compatible with the needs of the region. There are numerous small industries, and some not so small, in the region which process agricultural or forest products with large quantities of biomass waste. Disposal of the waste materials in an environmentally acceptable manner poses a considerable problem. Also, many of these industries require quantities of steam and/or process heat. Examples are furniture factories, sawmills, and lumber yards which require process heat for drying lumber, cotton gins producing about 200 pounds of waste per bale ginned and requiring process heat for drying, and rice mills producing copious quantities of rice hulls while requiring process steam.

Finally, the processing scheme would produce charcoal, a fuel which has been largely overlooked in the search for new energy sources, but which is not an undesirable fuel. It is storable, readily transportable, has a relatively high heating value and presents no great problems in firing.

The overall process would consume large quantities of solid waste materials while producing process heat and an alternate fuel, all in an environmentally acceptable manner. Thus, a proposal was made to the Department of Energy, and the project was subsequently funded.

OBJECTIVES/COST AND PERFORMANCE TARGETS

The objectives of the proposed project were:

(1) to develop engineering design data for rotary kiln pyrolytic converters.

(2) to make an engineering evaluation of the patented rotary kiln processing concept.

(3) to obtain pyrolytic conversion data for the biomass materials available in Arkansas and the Ozark Plateau Region.

The total cost of the project was $440,000. Objective (1), the scale model kiln phase, cost roughly $119,000 and is 90% complete. Objective (2) of the project required $216,000 and is about 60% complete in terms of goal achievement. Additional expenditures to complete this phase of the project would be minimal. Approximately $79,000 were expended on objective (3) and this task is about 50% complete. The remaining $26,000 of project funds were for a Washington-based management consultant firm.

APPROACH

The project was divided into three phases to coincide with the three primary objectives as listed above. These phases were (1) development of a scale model rotary kiln biomass converter, (utilization of the existing prototype kiln f

st and data acquisition, and (3) development
.a laboratory for the analysis of feedstocks and
oducts and for basic pyrolytic studies.

ecific objectives of phase (1) were:

) to provide a facility which could be used to evaluate the process at less cost and with greater operating flexibility than by using the full-scale converter.

.) to design and construct the scale model rotary kiln.

) to operate and collect data for material and energy balances using a wood waste feed.

.) to optimize the converter operation with the wood waste feed.

) to operate with feedstocks other than wood waste materials.

ie scale model kiln was designed to duplicate the
tention time and the bed agitation of the full-
ale rotary converter. Overall kiln dimensions
e 35 inches long by 26-inch o.d. by 14-inch i.d.
iese dimensions duplicate the length/diameter
itio of the prototype kiln. Kiln feed rates to
i0 lb/hr and kiln speeds to 16 rpm are possible.
ie angle of inclination can be varied from zero
roughly 10 degrees. Air can be introduced into
ie kiln at six different locations, and five ther-
couples are located within the kiln for measure-
ent of temperature profiles. Bed depth is con-
rolled by the angle of inclination and by a re-
ovable discharge dam.

iase (2) of the project was the testing and acqui-
ition of operating data for the prototype unit
ith various feedstocks. Complete details of this
iit have been given previously [1].

nase (3) of the project was development of an ana-
ytical laboratory for support of the other two
hases of the project. Basic batch pyrolytic data
ere also to be obtained for the various feedstocks
nder carefully controlled conditions.

ESULTS/ACCOMPLISHMENTS

t the inception of the project it was thought that
he extent of phase (2) of the project would con-
ist of the design, purchase, and installation of
n instrumentation system for the acquisition of
ata for performing material and energy balances.
he need for repair of portions of the existing
onverter was evident, but this was thought to be
secondary problem which could be handled as the
nstrumentation proceeded. However, the reverse
ituation has been the case. Repair of the exist-
ng unit has become the primary task in order for
he unit to remain operational, and instrumenta-
ion has been relegated to a secondary task. Cur-
ently, the mechanical problems have been addres-
ed, instrumentation has been installed, and the
valuation is continuing.

The laboratory, phase (3) of the project, has been developed and analytical data has been provided for the other phases of the project, but essentially no basic pyrolytic data has been obtained.

Phase (1) of the project is essentially complete. Data collection was completed for wood waste feed rates of 30, 60, and 90 lb/hr, and data analysis is now underway.

In addition to the wood waste feed, data was collected for several other feedstocks, including rice hulls, cotton gin waste, and scraps of partical board material.

Operation with the rice hull feed was very successful. Char was produced at relatively low pyrolysis temperatures of 800-900°F. These low temperatures eliminate the problems of silica fusion which are encountered at the higher combustion temperatures of a stationary boiler. Copious quantities of rice hulls are available in Eastern Arkansas and in other rice producing states.

The cotton gin waste was also successfully processed. The fibrous fraction was completely incinerated while the woody fraction was converted to char. This waste has been processed with other types of incinerators, but less than completely satisfactory results have been obtained to this point. It is estimated that roughly 70,000 tons of gin trash are generated annually in Arkansas.

Scraps of particle board from a television manufacturing operation were also processed. Good kiln operation was achieved with this feedstock and excellent char was produced. There are large quantities of this scrap material, on the order of 50 to 100 tons per day from this single manufacturing location, available for processing.

Roughly 100 experimental kiln runs have been made to this point. Several sets of typical data are attached as Table I. Analysis of the data is continuing and significant results will be presented as they become available.

FUTURE PLANS

It appears that the economics are very favorable for the use of a rotary kiln biomass processing system for disposal of agricultural and forest waste products with the production of charcoal and process heat and/or steam. This is especially true in Arkansas and the Ozark Plateau Region where there are many small industries producing significant quantities of biomass waste materials while at the same time requiring process heat and/or steam in their processing scheme.

It is estimated that the capital investment for a nominal 50 TPD processing unit, including the auxiliary equipment is $126,000. Annual operating expense for the unit is estimated at $105,000. Annual income from the unit would amount to

$195,000 in charcoal sales at current prices. A saving of $20,000 per year would also be realized from waste disposal costs. This translates to a return on original investment exceeding 40%. No dollar value is put on the process heat and/or steam produced because this would depend on whether or not the particular industry would be able to utilize it.

Plans are to disseminate the project results to potential user industries. It would also be very desirable to continue the investigation in order to study other variables of the system, to encourage the development of charcoal as a fuel, and to further develop the rotary kiln biomass processing concept.

TABLE I - SELECTED DATA

Feed Type	Feed Rate (lb/hr)	RPM	Air Rate to Kiln (lb/hr)	Feed Conversion (lb char/100 lb feed)	Average Bed Temperature (°F)
Wood Waste	30	2	12	14	900
	60	2	29	14	900
	90	2	35	7	1000
	30	8	12	10	900
	60	8	29	14	900
	90	8	35	7	1100
	30	14	13	10	1000
	60	14	29	6	1350
	90	14	35	5	1100
Rice Hulls	60	2	25	17	850
Pecan Shells	30	2	7	27	800
Macadamia Nut Shells	30	2	31	42	1000
Gin Waste	25	6	90	10	1100
Particle Board	45	2	120	5	1050

LITERATURE CITED

(1) J. R. Kimzey, "Utilization of Existing Converter," Proceedings of Sixth Biomass Thermochemical Conversion Contractors' Meeting, Tucson, Arizona, January 16-17, 1979.

SYNTHETIC FUELS FROM A LARGE GASIFIER

DOE Contract No. DE AC02 79 ET23029
Starting Date March 1, 1979

V. J. Flanigan
Y. Omurtag
J. E. Halligan
University of Missouri-Rolla
Rolla, Missouri

N. E. Welch
Sverdrup/Aro, Inc.
Tullahoma, Tennessee

BSTRACT

he University of Missouri-Rolla's Large Gasifier roject is a development program supported by the epartment of Energy's Biomass Energy Systems rogram. The UMR program is just beginning and at he present time is in the construction phase. ts purpose is to determine scale-up factors for ow, medium, and high Btu gasification processes or transition from medium sized to commercial perations. The resource recovery system donated o the University of Missouri-Rolla by the Adolph oors Company will be modified and used to obtain he necessary development data. The program will nclude the operation of a fluidized bed reactor esigned to handle mass flow rates of 200 to 2000 bs/hr with and without a catalyst to determine ptimum design and operating parameters for wood esidue gasification. The modification and development will include the design and use of an inserted leeve, which will provide an unusual two-reactors-n-one vessel arrangement. This sleeve will also ake possible a wide range of operating conditions. revious data collected by Coors and others are resented to support the system's design and onfiguration.

NTRODUCTION (DESCRIPTION OF TASK)

he University of Missouri-Rolla Wood Gasification GROW) project is a development program that is eing funded by the Department of Energy. The urpose of the program is to determine scale-up actors for low, medium, and high Btu gasification rocesses for transition from medium sized to ommercial operations. The program is a four-phase rogram. The first three phases relate to the Btu ontent of the gas to be produced, and the final hase to a specific industries' utilization of the as. The Coors resource recovery system donated o the University of Missouri-Rolla by the Adolph oors Company will be modified and used to obtain he necessary development data. The modifications ill include the insertion of a reactor sleeve for ome of the runs as well as many other equipment nd operational changes. The development task ill be focused on the operation of a fluidized ctor designed to handle mass flow rates of 2000 lbs/hr with and without a catalyst so that the design and operating parameters for wood residue gasification can be determined for this type of reactor. It is expected that reliable scale-up data for design and operating conditions will have been established. The program will also provide information on the economics of operating such systems, the market potential for the products, and the data base needed for technology transfer and commercialization.

The program is in its initial stages of development, and the modified Coors process is presently being refabricated. The flow diagram shown in Fig. 1 is the flow scheme for the first phase (Low Btu Gas Production) of the project. Funds received from the Department of Energy in April have allowed the research team to order the necessary additional equipment and to contract for the services of sub-contractors. With the purpose of the project being the production of three different Btu content gases, the objectives as they relate to the purpose are listed below.

OBJECTIVES

The specific objectives of the program are:

1. To demonstrate the operation of a large-scale reactor and develop scaling data. A key advantage of the Coors system is that it was designed to process up to 2000 lbs/hr of biomass. A sleeve will be placed in the reactor during the initial runs to reduce the diameter of the reacting section from 40 to 22.5 in. (Fig. 2). After experience has been gained by running the reactor with the sleeve in place, it will be removed to obtain data from the full-scale system.

2. To obtain higher thermal efficiency. The reactor's heat economy will be improved by using a top feed and by gasifying and producing methanation within the reactor.

3. To minimize the reactor's size per unit volume of throughput by using catalysts.

4. To develop data on the characteristics of feed

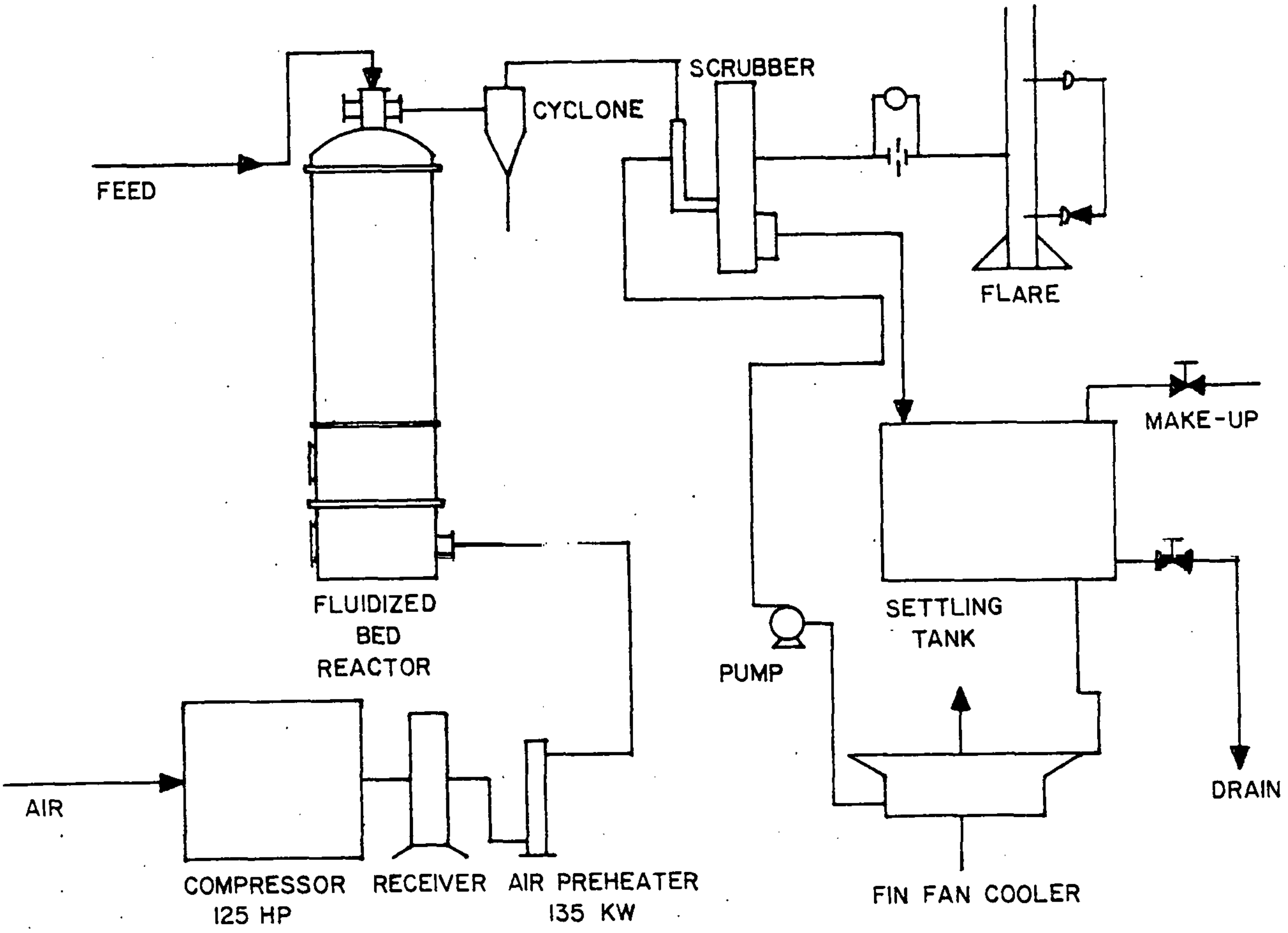

Fig. 1. Phase 1, Flow Diagram

stock and on their effect on gasification.

5. To develop economic data for the design and operation of a comparably sized unit for the production of low, middle, and high Btu gas.

6. To establish an environment in which the transfer of technology to industry can be achieved effectively. There will be active industrial participation in all phases of the project.

UMR-COORS RESEARCH FACILITY HISTORY

In the early 1970's, the Adolph Coors Company launched a research and development project to determine the feasibility of meeting their energy needs by converting municipal solid wastes to a low Btu gas. They designed and constructed a pilot unit and conducted experiments over a period of several months. However, as a result of changing company priorities, they abandoned the project and later, upon inquiry, agreed to donate the system to the University of Missouri-Rolla.

After much investigation, it was decided that there were many problems requiring additional research in this area and that the establishment of the Coors system as a research facility that could be used by engineers and scientists would be a more effective way of using the equipment.

Subsequent contacts along these lines with various interested parties within the University, the Sverdrup and Parcel Corporation, the St. Louis Regional Energy Council, various wood industry representatives, and the Missouri State Department of Natural Resources produced enthusiastic support for the establishment of a Wood Energy Research Program and the installation of the Coors system as a research facility at the University of Missouri-Rolla. Various projects were formulated within the program to investigate the desirability of using wood and other agricultural by-products as renewable energy sources. It was also proposed that the Coors facility be used to determine the design and operating parameters of energy conversion systems that use wood and other renewable resources for industrial and commercial use.

PROJECT DESCRIPTION (APPROACH)

The project as previously stated will consist of

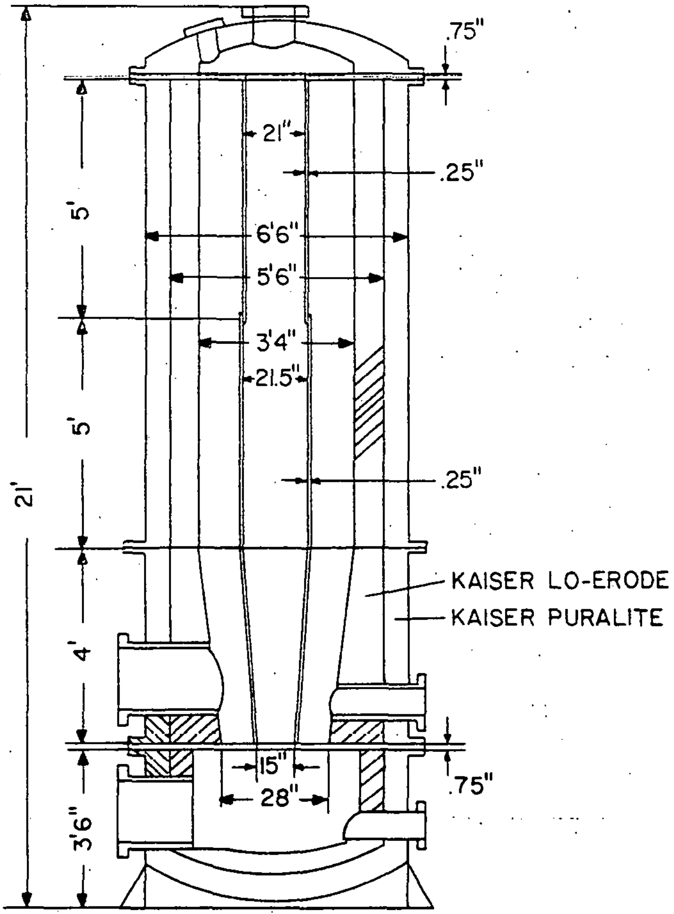

Fig. 2. Reactor Sketch (Sleeve Inserted)

our phases: 1) Low Btu Gasifier Operation,
) Medium Catalytic Btu Gasifier Operation, 3) High
tu Catalytic Gasifier Operation, and 4) Assess-
ient of the Potential to Specific Industries. The
ocus of the effort at the present time is Low and
edium Btu operation.

hase I: Low Btu Gasifier Operation

hase I of the program will involve a preliminary
'shake-down" of the UMR-Coors low Btu gasifier that
ill include cold studies of the reactor. This will
rovide some test data with the modified equipment.
he hot studies will make it possible to compare
he original Coors data with the checkout of the
ata acquisition system. This will be accomplished
y operating the reactor with a 22.5-inch diameter
sleeve) inserted and with the full 40-inch diameter
f the reactor open. The reactor will be air blown.

ith the air preheaters, the heating value of the
roduced gas is expected to be in the 100 to 150
tu/SCF range. The economics of producing the gas
ill be determined, and the feed system will be
hecked for pressure operation. The feed charac-
ics and the reactor operating conditions will
ited to the composition of the gaseous product
attempt to locate an optimum operating point.

The data will also be reviewed relative to other small-scale studies to determine scaling correlations. Figure 1 shows the system configuration in this phase.

Phase II: Medium Btu Catalytic Gasifier Operation

In Phase II of the project, the operating and scaling data for the medium Btu wood gasifier will be developed. During the phase, data will be obtained with a 22.5-inch sleeve inserted in the reactor and with the 40-inch diameter of the reactor open. Recycles with and without CO_2 removal and catalytic methanation will be incorporated. These data will be utilized to develop scaling correlations.

Figure 3 shows the system modifications that will be necessary for this phase. These modifications are needed to allow for catalytic methanation, CO_2 stripping, and recycling of the product gases as well as the catalyst material. An improved feed handling system will also be needed to allow for better control of the feed rates and catalyst blending.

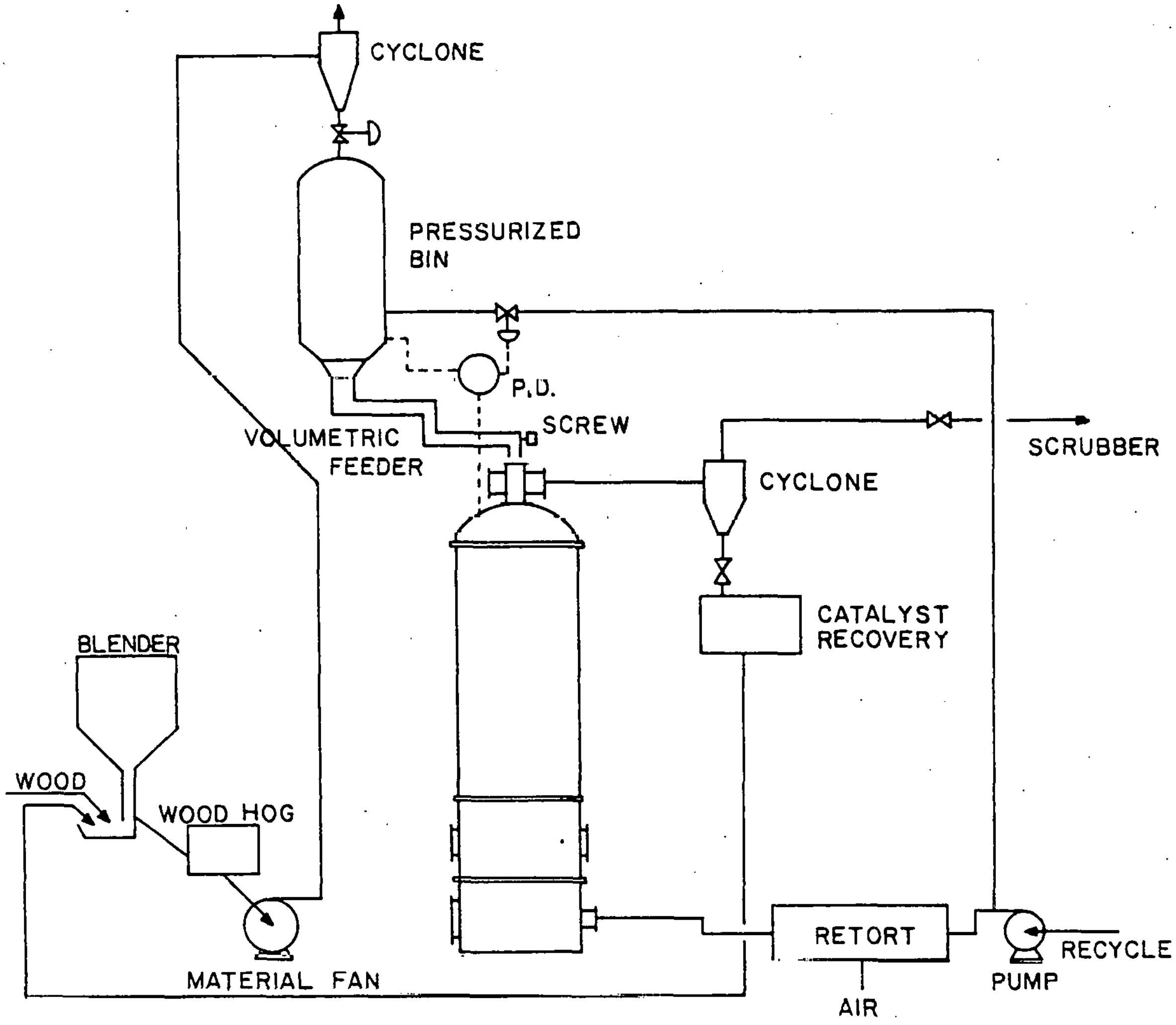

Fig. 3. Phase II, Flow Diagram

EXPERIMENTAL JUSTIFICATION (KEY RESULTS)

Although Phases I and II of the program are intended to be stages of development for a high-Btu (pipeline grade) gasifier, they will provide equipment, operational, and environmental data for low and middle Btu gasifiers that will have a reasonable probability of being economically productive and having an acceptable environmental impact.

Phase I

The equipment to be used in Phase I of the program is basically the original Coor's air-blown system with a top rather than a bottom feed. The feed will be introduced at the top through the gas outlet so that it will quench the gas prior to leaving the reactor. This will allow for a minimum heat loss in the quenching process. In the cooling process, the flashing of water in the green wood should occur at a sufficiently high rate to produce in excess of 90% of the enthalpy decrease in the gas within two to three feet of the top of the reactor. The disadvantage of a top feed in an air-blown, low Btu gasifier, without recycle, is that traces of pyroligneous acids may appear in the overhead and considerable reductions in the operating range for fluidization. The problem with the bottom feed is that the residence time in the high temperature region of the reactor where therma cracking of the acids occurs is too low. The problem with the top feed is that one must keep the upper velocity limit low enough so that the feed will not blow out with the overhead.

The operating conditions used by Coor's personnel [1] for the air gasification of paper-cardboard feed were as follows:

Feed Rate	500 (lb/hr)
Air Temperature	1000 (°F)
Air Rate	1100 (lb/hr)
Bed Temperature (sand)	1200 (°F)
Reactor Pressure	3 (psig)

Th erage heating value of the gas produced was ap imately 100 Btu/SCF. It had a yield of 12.2 SCF per pound of feed that corresponded to a feed conversion of 93%. The composition of the feed and of the samples of gas for two "runs" are shown in Table I.

TABLE I

ULTIMATE ANALYSIS OF PAPER-CARDBOARD FEED MATERIAL

Ultimate Analysis (% by wt.)

Component	Percent
Carbon	46.88
Hydrogen	7.40
Nitrogen	0.11
Sulfur	0.27
Ash	3.49
Oxygen (by diff.)	41.85
Total	100.00

PRODUCT GAS ANALYSIS

Mole % of Components

Gas Temp (°F)	CO_2	CO	C_1	C_2	C_2O	H_2	O_2	N_2	BTUS/SCF
1200	8.4	10.8	3.3	---	0.3	5.0	7.9	64.3	87
1200	13.8	17.9	5.5	---	0.5	8.3	---	54.0	145

The thermal efficiency of the system was approximately 75% without recovery of the sensible heat of the overhead. It is reasonable that a thermal efficiency of 90% could be attained by raising the operating temperature and recovering the sensible heat of the overhead. The data from the air-blown Coor's system generally agree with the data from fixed bed reactors.

Further justification for the decision to change to a top feed system was obtained from work done at Texas Tech [2]. The investigators at The University of Missouri-Rolla sent ten barrels of oak sawdust to the Texas Tech SGFM facility for a test run. This facility is a top feed fluidized bed gasifier of approximately 50 lb/hr size. The reactor was operated in the air blown mode for the tests using the oak sawdust. The small amount of steam was also supplied to the reactor. The average feed rate was 20 lb/hr of 4% moist sawdust, and the average higher heating value of the produced gas for the tests was reported to be in the range of 350 Btu/SCF. The test report stated that the steam seemed to have little effect on the reactor ion and on the gas produced. In conclusion, ported Btu content was considered to be very aging and certainly helped to substantiate the top feed system for the biomass material in the gasifier at the University of Missouri-Rolla.

Phase II

A non-air blown system will be used in Phase II of the program. The heat required to sustain the reactions will be supplied by externally heating a recycle stream with a preheater (Fig. 3). Because approximately 38 weight-percent of the reactor overhead is water vapor, steam reforming of the tars will occur. This is critical for three reasons. First, the tars are carcinogenic. Second, they increase the feed conversion and thereby the thermal efficiency. Third, they reduce the pollutants that must be removed from the product gas. Considerable data recently obtained from laboratory and bench-scale systems support these facts. Some runs were made with the reactor by Coor's personnel who injected superheated steam. Although their data do not include the effect of an increased residence time that would be produced by recycling the product gas components, they do serve to illustrate the effect of steam reforming. The operating conditions used for the steam gasification of the paper-cardboard feed were as follows:

Feed Rate	400 (lb/hr)
Steam Temperature	1000-1350 (°F)
Steam Rate	1800 (lb/hr)
Reactor Pressure	3 (psig)

The heating value of the product gas was proportional to the steam temperature and ranged between 200 to 366 (Btus/SCF) for steam temperatures between 1000° and 1350° F.

The two most significant points are that no unpyrolyzed feed material and tars were found in the reactor or sand bed after the runs were made, and very little solid material (0.56 to 0.68%) was detected in the scrubber water.

It is of interest to note that the distribution of components (Table II) are not near equilibrium. This is not surprising, because the residence time is less than seven seconds, and no catalyst was used.

The Pacific Northwest Laboratory [3] has obtained data for catalytic steam gasification by using laboratory apparatus. They have also begun to operate their Pilot Development Unit and have obtained catalytic gasification data from its operation. Their data will be helpful in evaluating Phase II of the University of Missouri-Rolla project.

CONCLUSIONS (FUTURE MILESTONES)

The probability that the system proposed in Phase II will produce an environmentally acceptable middle Btu gas is high, but there are three principal problems that will have to be solved in developing the middle Btu gasifier. These are:

TABLE II

PRODUCT GAS ANALYSIS SUMMARY

Steam Runs

Run Temp(°F)	CO_2	CO	C_1	C_2	C_2	H_2	O_2	N_2	BTU/SCF Total
1350	13.8	58.1	9.7	0.1	1.7	16.6			365
1350	25.1	51.5	8.7	0.3	1.0	13.4			312
1200	37.1	45.3	7.0	0.2	1.2	9.3			268
1200	31.0	50.0	6.0	---	---	13.0			264
1100	41.7	40.7	7.1	0.2	1.2	9.2			254
1100	39.6	42.7	6.1	0.1	1.3	10.2			255
1000	38.8	51.3	3.8	---	---	6.3			227
1000	50.1	39.4	3.6	0.3	0.7	6.0			201

1. Development of the top feed to gain thermal efficiency and the advantages of processing a lower temperature overhead and of "freezing" the gas composition at a high temperature condition.

2. Determination of a set of catalyst and operating conditions that will minimize reactor size and simultaneously reform the tars to gaseous products. The Pacific Northwest experiments should provide the bulk of this information prior to Phase II.

3. Development of hardware reliability data.

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2. Beck, S., "Biomass Masification in SGFM Process", Seventh Biomass Thermochemical Conversion Contractors Meeting, Roanoke, Virginia, April 1979.
3. Robertus, R. J., "Investigation of Gasification of Biomass in the Presence of Catalysts", Seventh Biomass Thermochemical Conversion Contractors Meeting, Roanoke, Virginia, April 1979.

ACKNOWLEDGMENTS

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APPLICATION OF SGFM TECHNOLOGY TO OTHER FEEDSTOCKS

Steven R. Beck
Department of Chemical Engineering
Texas Tech University
Lubbock, Texas 79409

ABSTRACT

The Synthesis Gas From Manure (SGFM) process has been under development at Texas Tech University since 1970. In 1976 a small pilot plant (approximately 1/2 ton per day) was constructed to evaluate the process in a continuous mode of operation. In 1977 a contract was awarded to Texas Tech to evaluate non-manure biomass feedstocks in the SGFM pilot plant. These feedstocks were to include wood as the major emphasis, along with various agricultural residues which might be available. The agricultural residues which have, or will, be looked at are corn stover, cotton gin trash, sunflower stalks, and sorghum stalks.

The SGFM reactor operates as a countercurrent, fluidized bed in which the solid feedstock is fed to the top of the reactor and fluidized by an upward flowing stream of air, steam, and product gases. This top feed system provides some advantages over conventional fluidized beds in that the incoming feedstock is partially dried by direct contact with hot product gases. In addition, the initial pyrolysis gases which are formed quickly exit the reactor zone and are quenched by the incoming cool solids. As a result, the product gas contains significant amounts of methane, ethane, and ethylene along with hydrogen, carbon monoxide, carbon dioxide, and nitrogen.

The major results to date pertain to the gasification of wood. The reactor was operated over the temperature range of 500°-800°C. Under these conditions the yields of product gas varied from about 16 SCF/lb daf to 20 SCF/lb daf. The higher heating value of the raw gas was 300-400 BTU/SCF. Preliminary results from corn stover gasification indicate that the gas yields from corn stalks will be very similar to those observed with wood. In both these cases, the gas yields are significantly higher than was seen with manure. An economic evaluation of the SGFM process using wood as the feedstock, indicates that for a 1000 ton per day plant, a medium-BTU fuel gas can be produced for approximately \$3.30 per million BTU's, assuming wood chips delivered to the plant at \$15 per ton.

DESCRIPTION OF TASK

The original objective of this project was to develop a process that would convert cattle feedlot manure to ammonia synthesis gas. The process which erged is based on a countercurrent, fluidized actor in which manure is gasified by partial oxidation. On August 1, 1976, the U.S. ERDA contracted with Texas Tech to develop a data base for gasification of manure in the SGFM pilot plant and to evaluate other biomass feedstocks using SGFM technology. The results of this Phase I study were presented in October, 1977 at the review held at Wright-Malta in Balston Spa, New York and have been published by Beck [1].

On August 1, 1977, Phase II of this project commenced under a continuation of the previous contract. The principal objective of this phase of the work is to develop a data base for the gasification of wood residues in the SGFM pilot plant and to investigate other feedstocks such as corn stover, sorghum stalks, sunflower stalks, and cotton gin trash in the pilot plant. The other tasks of this project are shown in Table 1.

Table 1
OBJECTIVES - PHASE II

1. Develop a data base for pyrolysis of wood residue in the SGFM pilot plant. The data base will include product yields and characterization, heat and material balances.
2. Develop a kinetic model of the SGFM process.
3. Obtain an independent economic evaluation of the SGFM process on a commercial size plant.
4. Develop correlations necessary to predict the performance of a variable velocity fluidized bed.

One of these tasks is to develop a kinetic model of the process suitable for use in reactor scale-up. A third task of Phase II is to subcontract an independent economic evaluation of the SGFM process utilizing manure as the feedstock. The Bechtel National Corporation was awarded this subcontract in 1978 and has completed this task. The final task is to investigate the fluidization characteristics of biomass feedstocks. This was aimed primarily at cattle feedlot manure due to its rather unique behavior in fluidizing situations.

OBJECTIVES

The goal of this work is to develop an economically competitive process to convert any biomass feedstock to a useful end product. The original objective was to produce ammonia from cattle feedlot manure because ammonia is in high demand in the agricultural areas where cattle feedlots are located.

It was felt that this was an ideal situation in that the supplies of natural gas are diminishing and ammonia manufacturers are searching for alternate sources of feedstock. The problem that has arisen over the past few years is that imported ammonia is starting to displace domestic ammonia at a much lower cost. Consequently, many of the ammonia plants in the United States are being shut down.

With this in mind the SGFM process was re-evaluated for possible options on the desired product. Currently it is felt that this process can produce a medium-BTU gas (<300 BTU/SCF) from biomass feedstocks such as wood residues. It is hoped, and is the goal of this work, to develop the process to the point that this medium-BTU gas can be produced in the range of \$3-\$4 per million BTU's.

A third, and perhaps more important, product that is produced in the SGFM reactor is ethylene. Ethylene is the largest volume petrochemical in the world and is the building block of the whole plastics industry. Due to the current shortages of petroleum in the United States and the world, it is felt that ethylene can be produced from biomass if a reactor can be designed to produce a gas stream containing >10% by volume ethylene. This is a feasible goal that can be reached in the SGFM process but needs additional work.

APPROACH

Pilot Plant Description

A schematic drawing of the SGFM pilot plant is shown in Figure 1. The heart of the pilot plant is the reactor itself. The reactor is 6" ID in the lower 5 ft. and 8" ID in the upper 3 ft. or disengaging zone. The solids are fed to the top of the reactor through a screw feeder, which controls the feed rate, and falls by gravity into the reactor itself. The air-steam mixture which enters the bottom of the reactor is preheated in a 25 ft. length of tubing that serves as a resistance heater. The char is removed from the reactor through a centerport opening in the bottom distributor plate. A hydraulic ram is used to prevent any bridging of the char in the discharge line. The gases exit the top of the reactor and pass into a cyclone that is operated at approximately 350°C. The cyclone is heated to prevent condensation of any of the reaction products and adequately removes most of the entrained solids. The gases leaving the cyclone then pass through a 3-stage impinger sequence which is operated at about 110°-140°C. This serves to condense the tar but maintains the water in a vapor state. Following the impingers, the water is condensed in a double-pipe heat exchanger and collected in the downstream impinger section. The product gases are then passed through a turbine meter for flow rate measurements and vented to the atmosphere. Using this arrangement good material balance data has been obtained, but there are some problems in operation of the pilot plant. With the screw-feeder, very few problems have been encountered in feeding the biomass. The tar collection system is currently the major problem. Tar produced from biomass feedstocks is a very viscous material and tends to condense on all piping and also plugs the impingers. For this reason, the errors that are apparent in material balance are primarily due to the inability to collect and measure all the tar produced. This is a relatively minor error because of the fact that

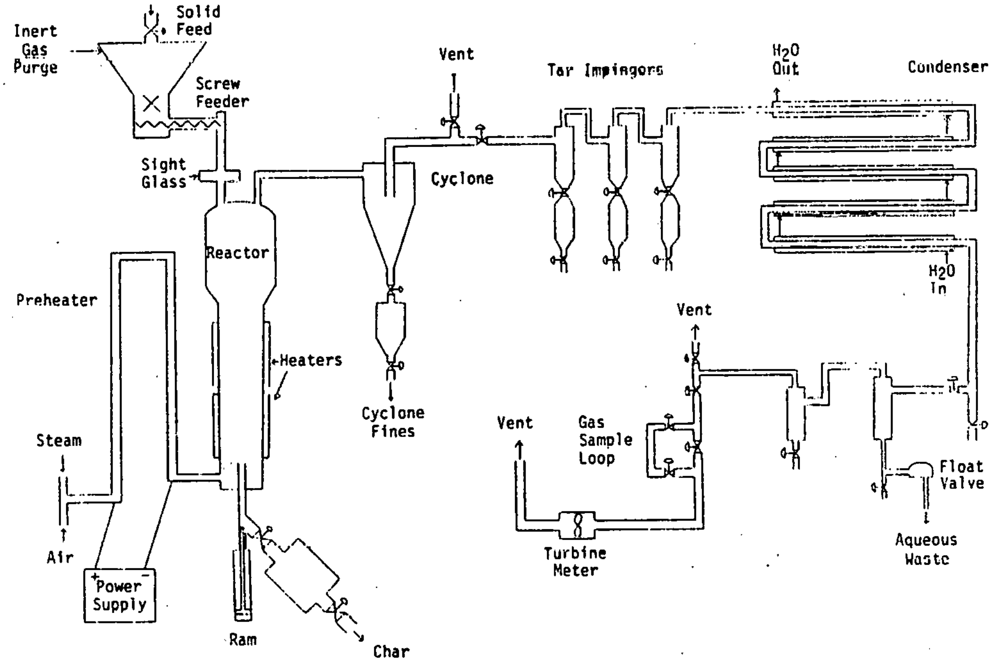

Fig. 1. SGFM Pilot Plant

th : product is only about 5% of the raw feedstock weight.

Material and Energy Balances

The major emphasis in this work has been to operate the pilot plant with various feedstocks and obtain material and energy balances for the reactor. The reactor temperature can be varied by varying the air-to-solid feed ratio as well as controlling the heaters located on the outer wall of the reactor. The major feedstock studied under this contract to date is oak sawdust, obtained from south central Missouri. The sawdust is in the form of small, uniform particles containing about 40% moisture. For many of the runs, the sawdust was air-dried to about 4% moisture and used in this form. Corn stover has also been used in the pilot plant but the results to date are not complete and will not be discussed in detail. During the next two months more data will be obtained on corn stover as well as the other agricultural residues discussed previously.

Economic Assessment

During Phase I of this contract, a detailed design data base was developed for cattle feedlot manure. It was felt that to obtain a realistic and fairly accurate design and economic evaluation for a commercial facility that an independent contractor was required to perform this task. This eliminated any bias which might be present in the minds of those working closely on the process. The Bechtel National Corporation was chosen to perform this assessment because they have had experience working with biomass feedstocks as well as tremendous resources involved in design of chemical processing facilities. The method used to perform this assessment was that Texas Tech provided the raw data to the Bechtel design engineers. A design basis was agreed upon jointly by Texas Tech and Bechtel and from that point on all the work was performed by Bechtel with no input from Texas Tech University. The results of this study were presented by Beck [2] at the Second Annual Fuels From Biomass Symposium in Troy, New York.

Three options were investigated by the Bechtel Corporation. These included the production of a fuel gas, the production of ethylene plus a fuel gas, and the production of ammonia. The major conclusion from the Bechtel study was that ammonia production from cattle manure may be competitive in the late 1980's and that the production of a fuel gas is only marginally competitive. One thing that should be pointed out is that in the fuel gas case, Bechtel included facilities to compress the gas to 500 psia and remove the carbon dioxide. This resulted in a serious economic penalty when looking at the overall process because the syngas compressor employed was a four-stage, 11,000 BHP compressor which cost nearly as much as the rest of the reaction section. This compression is necessary for the synthesis of chemicals from the product gas not necessary if the gas will be burned near cation of the reactor.

Economics of Wood Gasification

The pilot plant results for gasification of oak sawdust were used to evaluate the economic potential of the SGFM process for production of a medium-BTU gas from oak sawdust. The Bechtel design was used as the basis for designing a complete SGFM facility for wood gasification. The capital cost data presented by Bechtel were adjusted for difference in size using a 0.7 exponent and these capital costs were then used to evaluate wood gasification. The material and energy balances for the plant were performed by Texas Tech University as well as the complete evaluation. It is the opinion of this author that this should be a fairly accurate design for this facility. It was assumed in this design that the wastewater could be treated by conventional means rather than by incineration, which Bechtel used in their study. It was also assumed that what little char is produced from the reactor would be used in some type of wood-fired boiler to generate steam required for the facility. This seems to be a fairly reasonable assumption since the boiler required is on the order of 25,000 lbs. per hour of low-pressure steam.

DISCUSSION OF RESULTS

Wood Gasification

The design data base for gasification of wood in the SGFM pilot plant reactor is complete. The feedstock chosen for this study was oak sawdust from Missouri. The sawdust, as-received, contained about 40% moisture and 1% ash by weight, on a wet basis. The sawdust contained a few large chunks of wood which could not be fed into the reactor. Because of this and to eliminate other feeding problems, the sawdust was screened to -1/8" particle size. The particle size distribution of the sawdust is shown in Figure 2. This sawdust was air-dried in Lubbock overnight on the floor of our pilot plant facility. At the end of 16 hours of air drying, the moisture content was reduced to 4%. This was much lower than desired but due to the fact that it was the only feedstock we had it was used in the reactor. The final stages of the wood tests were conducted on sawdust that had been rewetted by soaking in water. This sawdust was used in 4 runs in the reactor and contained between 40%-50% moisture when it was loaded to the feed hopper. This is not the most desirable method to operate and future plans call for obtaining fresh, green sawdust and operating the reactor with this material at about 40% moisture content.

The material balance data for the dry and wet sawdust are shown in Tables 2 and 3, respectively. The material balance closure shows that in all runs the unaccounted for material was less than 11%. It is felt that this discrepancy is primarily due to the fact that the tar is difficult to collect and secondarily because variations in the bulk density of the fresh wood will cause some variation in the feed rate to the reactor. The screw-feeder feeds on a volumetric basis and not a weight basis. It is felt that the variation in feed rate should average out over the period of the run and consequently this is not a major contributing factor to the problems with material balance closure.

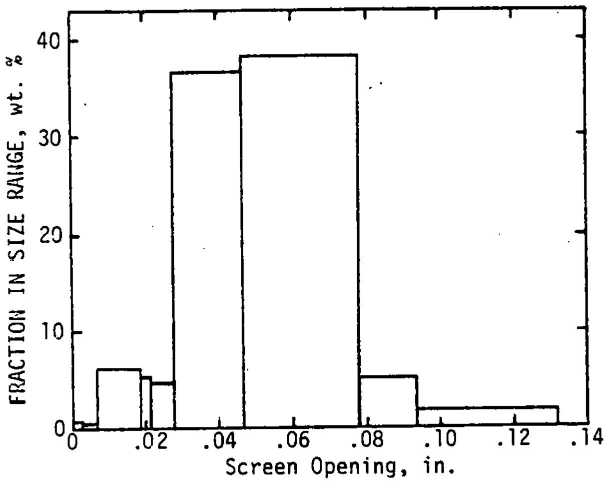

Fig. 2. Particle Size Distribution of Sawdust (-1/8")

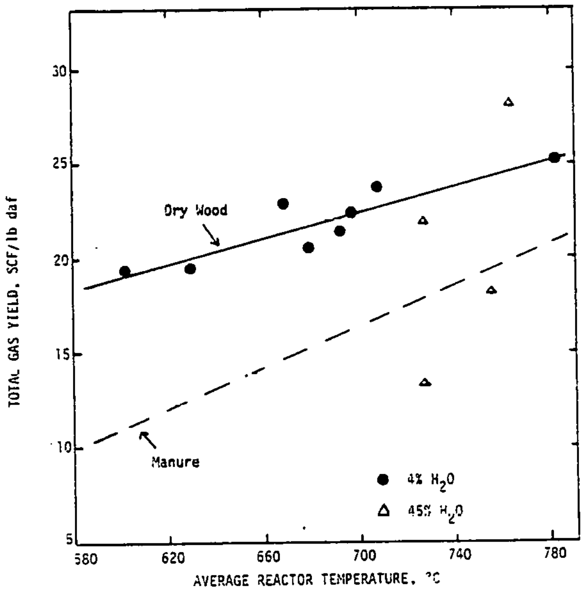

Fig. 3. Total Gas Yield From Oak

Table 2
MASS BALANCE RESULTS FOR WOOD GASIFICATION, 4% H_2O

Run Number	38	39	40	41	43	44	45	49	50
Input (lb/hr)									
1. Wet Sawdust	16	19	19	25	29	23	29	23	18
2. Air	8.03	12.6	11.79	9.35	9.35	7.6	5.3	8	8
3. Steam	7	6	6.1	6.5	8.8	8.0	8.0	11	8
4. Total input	31.03	37.6	36.89	40.85	47.15	38.6	42.3	42	34
Output (lb/hr)									
1. Product Gas	24	15.2	29.57	33.2	32.5	33.5	38.49	31.2	28.23
2. Water & Organics	5.26	18.1	3.48	6.73	12.8	4.62	3.63	7.86	1.51
3. Tar	0.35	1.24	0.4	0.32	0.4	0.57	1.16	0.09	0.25
4. Cyclone fines	0	0	0	1.83	0.2	0.07	0.02	0.30	0.21
5. Char	0.04	0.05	0.47	0.04	0	0.96	0.44	2.99	0.11
6. Total Output	29.67	34.59	33.92	42.12	45.9	39.72	43.74	42.44	30.31
Material Balance Closure % Output ÷ Input x 100	95.62	92.0	91.95	103.1	97.35	97.10	96.60	99.0	89.20
Av. Reactor Temp., °C	669	504	668	605	631	707	680	692	780

The gas yields obtained from wood as a function of temperature are shown in Figure 3. For Comparison, the gas yields obtained when using manure are also shown in Figure 3. It can be seen that a significantly higher gas yield on a dry, ash-free basis is obtained from wood when compared to manure. This is probably due to the difference in composition of the two materials. Cattle manure is relatively low in cellulose because the cow has digested the cellulose in its feed and excretes a material which is high in lignin content and high in protein content. Previous studies by Stamm [3] show that hemicellulose and cellulose gasify at a much higher rate than does lignin. As a result, materials such as wood which are high in cellulose and hemicellulose will produce more gas than will a material that is high in lignin, such as cattle manure.

The composition of the raw gas obtained from wood is shown in Table 4. Also shown in Table 4 is the calculated higher heating value of this gas. It can be seen that this gas is very suitable for use in boiler applications because no derating of the boiler would be required, but it is also suitable for chemical synthesis because of the high hydrogen content and the favorable H_2 to CO ratio. This aspect of the process has not been evaluated a this date. The principal use of this gas, at time, is assumed to be for use in boiler firing, or it could be used in an internal combustion engine.

Table 3
MASS BALANCE RESULTS FOR WOOD GASIFICATION, 45% H_2O

Run Number	54	55	56	57
Input(lb/hr)				
1. Wet Sawdust	14	14	14	19.5
2. Air	10.5	11.12	11.07	11.61
3. Steam	3.1	2.75	0	0
4. Total input	27.6	27.87	25.07	31.11
Output(lb/hr)				
1. Product gas	8.6	10.15	16.58	17.06
2. Water & Organics	17.9	12.99	5.04	10.35
3. Tar	1.3	1.18	0.45	1.25
4. Cyclone fines	0.3	0.21	0.14	0.17
5. Char	0.6	0.30	0.42	1.35
6. Total Output	28.7	24.83	22.63	30.18
Material Balance Closure % Output ÷ Input x 100	104.0	89.1	90.3	97.0
v. Reactor Temp., °C	728	752	760	727

Table 4
GAS COMPOSITION, REPORTED AS MOLE %

Run	38	39	40	41	43	44	45	49	50	54	55	56	57
H_2	14.96	12.59	11.36	15.33	10.95	15.86	12.89	13.79	12.99	1.48	5.72	6.18	10.11
CO_2	32.75	26.35	14.54	25.85	20.44	20.09	22.39	20.37	18.19	22.83	29.29	28.04	23.34
C_2H_4	3.43	1.89	4.18	2.49	3.01	4.23	4.94	3.47	4.30	4.05	3.53	3.52	2.94
C_2H_6	1.14	1.17	0.91	1.66	1.09	0.97	1.00	0.73	0.89	6.42	0.86	4.37	--
C_2H_2	T	0	0.18	1.01	0	0.53	0.18	0.18	0.27	0.09	0.09	0.19	--
N_2	15.93	21.22	24.73	18.09	26.28	14.18	6.76	14.05	17.29	29.64	33.11	38.31	55.51
CH_4	11.79	7.91	12.00	6.37	8.12	11.89	12.79	12.54	13.26	--	9.25	11.88	6.07
CO	19.98	26.62	31.09	27.88	28.19	31.63	38.02	36.12	32.52	41.11	18.03	11.22	1.93
C_3^+	--	2.16	1.00	1.29	2.46	0.62	0.73	0.18	0.27	--	--	--	--
Avg. MW	27.95	28.34	26.10	27.59	28.01	25.77	26.74	26.55	25.99	32.97	30.07	30.57	28.35
HHV raw gas, BTU/SCF	307	311	369	320	337	381	410	365	376	316	243	312	147
HHV CO_2 free	456	423	431	406	424	462	500	451	459	410	343	434	192

eat Balance Results. For each run on the reactor, n energy balance was performed in an attempt to etermine the net heat of reaction for that partiular run. Included in the heat balance data are he heat losses from the reactor due to inadequate nsulation and non-adiabatic reactor operation, as the heat input provided by the heaters. The c heaters on the reactor are designed to e the heat losses but they also pose some problems in accurately determining the energy required to conduct the reaction.

The heat balance results are shown in Tables 5 and 6 for the dry and wet sawdust, respectively. These heat balance results use an estimated heat capacity for the organic liquids (tar) produced and the aqueous waste which is assumed to contain about 5% organic material. The heat capacity of these materials was estimated assuming that the organic portion is about 50% long-chain fatty acids, 25% short-chain fatty acids, and the balance aromatic and polyaromatic hydrocarbons. This composition was assumed based on the analyzed waste from the cattle manure runs performed in Phase I of this contract. The net heat of reaction which was calculated for each run was correlated with the ratio of the air fed to dry, ash-free material fed to the reactor. This is shown in Figure 4 for the dry sawdust and Figure 5 for the wet sawdust. A linear regression was performed for these two cases to extrapolate the net heat of reaction to zero air-to-feed ratio which should give the heat of pyrolysis. These results indicate that the overall pyrolysis reaction with dry sawdust is mildly exothermic with a heat of reaction of about -210 BTU/lb daf material converted. The heat of pyrolysis for the wet sawdust is endothermic with a heat of reaction of about 320 BTU/lb daf material converted. If one examines the overall net heat of pyrolysis between the wet sawdust and dry sawdust, the difference is 530 BTU/lb daf material. This is the energy that is required to vaporize the additional water contained in the sawdust and consequently it is felt that the pyrolysis of dry, ash-free sawdust is exothermic which confirms the results of Coffman [4].

Table 5
HEAT BALANCE RESULTS, 4% H_2O

Run Number	38	39	40	41	43	44	45	49	50
Input, BTU/hr (1)									
1. Sawdust	0	0	0	0	0	0	0	0	0
2. Air	1643	2857	2292	1863	2233	1899	1224	2286	2212
3. Steam	8645	7650	7381	8918	11466	10560	10280	15180	10984
4. Helium	23	26	13	29	34	35	30	44	42
5. Heaters	14740	0	11669	12971	14740	15930	16255	14740	14642
6. Total input	25051	10533	21355	23781	28473	28424	27789	32250	27880
Output, BTU/hr									
1. Losses	10654	7867	10087	9281	9694	10752	10357	10420	11824
2. Steam	9968	7908	8418	9139	12760	11400	11360	15675	11960
3. Product gases	5950	2148	5089	7346	7404	7738	8888	6725	7590
4. Char	13	12	146	11	0	317	140	1001	90
5. Tar	183	461	198	145	189	301	589	52	172
6. Org. liquids	2748	3366	1726	3047	6062	2439	1844	4546	1041
7. Helium	49	31	28	32	55	47	34	49	58
8. Cyclone fines	0	0	0	518	59	23	6	108	47
9. Total output	29565	24811	25692	29519	36223	33017	33218	38656	32782
Net heat of Reaction BTU/lb daf	297	615	240	241	281	210	197	293	287
lb air/lb daf feed	0.53	0.69	0.65	0.39	0.34	0.35	0.19	0.36	0.47

(1) Reference Temperature: 25°C

Table 6
HEAT BALANCE RESULTS, 45% H_2O

Run	54	55	56	57
Input, BTU/HR				
1. Sawdust	0	0	0	0
2. Air	3097	3357	3163	3425
3. Steam	4359	3900	0	0
4. Helium	46	47	37	31
5. Heaters	19654	18098	14372	15968
6. Total Input	27,156	25,402	17,572	19,424
Output, BTU/HR				
1. Losses	11005	11682	11825	11429
2. Steam	4328	3911	0	0
3. Product Gas	1610	2040	3450	3202
4. Tar	709	669	732	681
5. Aqueous Waste	9758	7362	2502	5642
6. Char	204	160	136	460
7. Cyclone Fines	102	74	45	58
8. Helium	45	48	44	46
9. Total Output	27,761	25,946	18,334	21,517
Net Heat of Reaction BTU/lb daf	76	71	100	200
lb air/lb daf	1.377	1.458	1.452	1.536

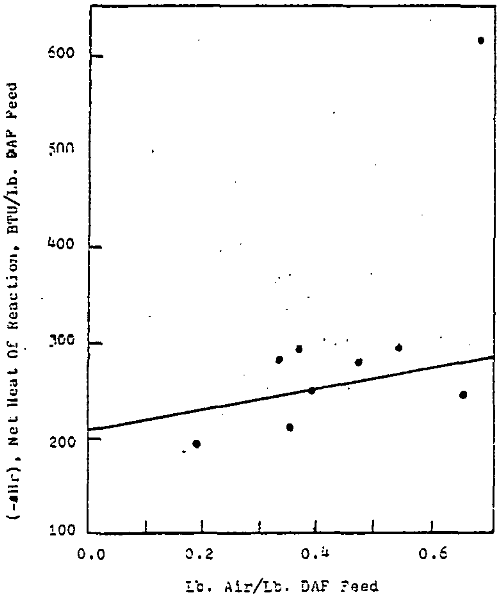

Fig. 4. Estimate of Heat of Reaction For Sawdust (4% Moisture)

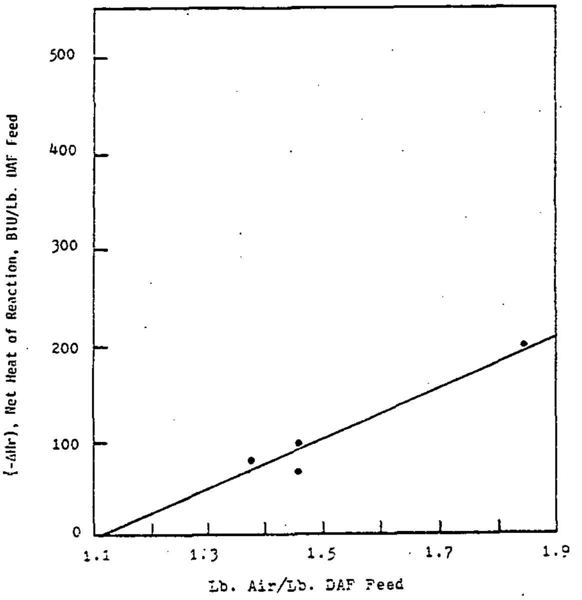

Fig. 5. Estimate of Heat of Reaction for Sawdust (45% Moisture)

ing the heat of pyrolysis results, it was calcu-
ıted that wood containing up to 30% moisture could
fed to the reactor and the reactor would still
erate in the autothermal mode without the neces-
ty of adding external heat. This was performed
suming that the permanent heat losses are 10% of
e total sensible heat out of the reactor and that
% of the remaining sensible heat in the product
ses can be recovered in a usable form. Huffman
] reported that for cattle manure in the same
FM reactor, the moisture content of the feed could
as high as 50% even with the assumed losses.
e discrepancy between wood and cattle manure has
t been satisfactorily explained at the present
me, but is being carefully evaluated.

rn Stover Gasification

ree runs have been attempted using corn stover in
e SGFM pilot plant. In two of these runs feed-
g problems forced termination of the test. In
e third run an acceptable material balance was
tained. For this study, the corn stover was
ound in a hammermill such that it would pass a
4" screen. This does not mean that the particles
e all smaller than 1/4" because the material tends
be very fibrous and some particles were greater
an 1" long but only about 1/16" in diameter. The
rn stover contained 3.5% moisture and approxi-
tely .9% ash.

e successful run was conducted at an average re-
tor temperature of 674°C. The gas yield was
.4 SCF per lb daf which is slightly less than
od but much greater than the manure. No conclu-
ons can be drawn at this time related to corn
peration in the reactor. A new cyclone is
stalled in the reactor because the entrain-
solids using corn stover is much greater
an from either wood or manure and the existing
cyclone was unable to handle the higher solids loading.

Table 7
BASES FOR ECONOMIC EVALUATION OF SGFM PROCESS

FEED: 1,000 ODT/CD Wood Feed
50% Moisture by Weight

STREAM FACTOR: 90%

YIELDS: From Pilot Plant Run 44
Gas: 20 SCF/lb daf
HHV: 375 BTU/SCF
Reactor Temp.: 700°C
Tar: 0.026 lb/lb daf (recycle to reactor)
Char: 0.050 lb/lb daf

ECONOMIC CRITERIA:
100% Equity Financing
12% Discounted Cash Flow Rate of Return (after taxes)
20 year straight line depreciation
48% Federal Income Tax

Gas delivered at 70°F and 30 psia.

Economic Evaluation

The results of the wood gasification tests were used to perform a preliminary design and economic evaluation of the SGFM process. This design was based on 1000 oven-dried tons per calendar day of green wood fed to the plant. Assuming a 90% stream factor and 50% moisture in the wood, the as-received feed rate to the plant was 2,222 tons per stream day. The Bechtel report was used as the basis for the design but no gas purification was included which meant that the product gas contained both carbon dioxide and nitrogen. Run 44 in the pilot plant was used as the basis for the design. In this run, the gas had a higher heating value of 375 BTU/SCF. The product from the plant is gas at 70°F and 30 psia. This is a low pressure gas as opposed to the reaction section of the Bechtel report in which a compressor was included to produce synthesis gas at 520 psia. As a result of deleting this compressor, the capital cost is much less than what Bechtel reported for the reaction section of their design.

The economic design bases used for this plant are shown in Table 7. A summary of the capital costs is shown in Table 8. The total investment for the plant is $22.4 million, which includes both installed equipment and working capital. A summary of the operating costs for this process are shown in Table 9. These are shown for wood delivered to the plant in the form of chips at $10, $15, and $20 per green ton. For these three cases, the required selling price of the gas to achieve a 12% ROR using 100% equity financing was calculated. The results of this are shown in Figure 6. It can be seen that the future of this process is very promising when one considers that natural gas which will be imported from Canada and Mexico will sell for $2.50-$3.00 per million BTU's and that imported LNG is

Table 8
CAPITAL COST SUMMARY

ITEM	MILLIONS OF DOLLARS
Battery Limits	10.3
Off Plots	7.0
Total	17.3
Interest During Construction	1.7
Start Up Cost	1.2
Depreciable Investment	20.2
Land	0.2
Working Capital	2.0
Non-Depreciable Investment	2.2
Total Investment	22.4

Table 9
OPERATING COST SUMMARY

ITEM	ANNUAL COST, $MM		
Wood Chips ($/green ton)	7.3 (10)	11.0 (15)	14.6 (20)
Chemicals	0.3	0.3	0.3
Utilities			
Water($.40/MGAL)	0.4	0.4	0.4
Electricity (35 mil/KWH)	0.6	0.6	0.6
Labor	1.2	1.2	1.2
Overhead	0.6	0.6	0.6
Local Taxes and Insurance	0.6	0.6	0.6
Ash Disposal	0.2	0.2	0.2
Depreciation	1.0	1.0	1.0
Total	12.2	15.9	19.5

selling for greater than $3.00 per million BTU's.

FUTURE PLANS

The remaining three months on this contract will be utilized to obtain data on as many different feedstocks as possible. The primary feedstocks which will be investigated are corn stover, sorghum stalks, cotton gin trash, and sunflower stalks. In addition, if time permits and a source can be found, wheat straw and bagasse will also be tested. The kinetic model which was discussed as one of the tasks, will be completed by the end of the contract. A crude model is presently available but is being improved based on the results from wood gasification. In addition to the pilot plant results, a study of the fluidization characteristics of cattle manure has been completed. This shows that some biomass feedstocks do not behave in a manner such that a fluidized bed can be easily designed for gasification. By the end of the contract, the critical parameters which must be determined for biomass should be identified and will have been measured for cattle manure.

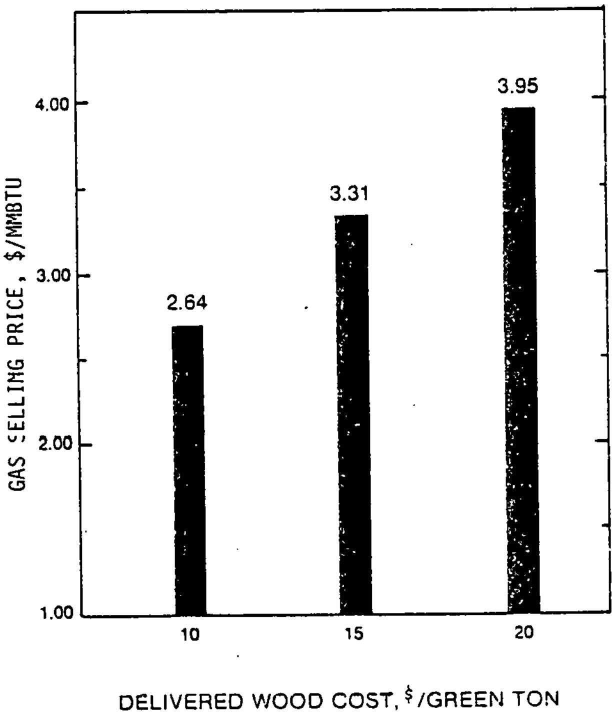

Fig. 6. Product Gas Selling Price

A continuation of this project has been requested from DOE. During the continuation period, or Phase III, of this project, the SGFM reactor will be operated in an enriched-air, or oxygen-blown mode. The primary purpose of this work will be to determine if methanol production, using the SGFM process, is feasible. Based on the air-blown results, it appears that a very good synthesis gas can be produced in the SGFM reactor.

Another major objective of the continuation work will be an attempt to maximize ethylene production in the SGFM reactor. It appears that if the ethylene content of the product gas can be increased to about 15%, it may be economically feasible to separate this ethylene and sell it as a separate product from the plant. There is currently a great deal of interest in industry to find new sources of ethylene, particularly by the major chemical and plastics producers, and biomass may be one option for this in the future.

CONCLUSION

The work to date on the SGFM project has demonstrated that this reactor and process is flexible enough to handle several different types of biomass feedstock in a fairly wet form. The econ evaluation shows that the Synthesis Gas From M process developed at Texas Tech University is a viable option for production of gaseous fuel from biomass feedstocks.

ACKNOWLEDGMENTS

he author wishes to thank the U.S. Department of nergy (Contract No. EY-76-S-04-3779) for supporting this work. We also wish to thank the Texas attle Feeders Association, Pioneer Corporation, nd the Environmental Protection Agency for funding the early phases of this work which permitted s to develop the process to the point where it s now.

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NOTES

STEAM GASIFICATION OF BIOMASS

John A. Coffman
Wright-Malta Corporation
Ballston Spa, NY 12020

Contract No. ET-78-C-02-4736
Started Sep 29, 1976

The project purpose is development of a process and equipment for conversion of as-harvested green wood chips into medium Btu fuel gas. The chemistry that has proven to be most satisfactory is wood ash-catalyzed steam gasification (1150° F, (300 psi), one hour dwell time). The equipment that has been evolved is a slender, jacketed rotary kiln, heated regeneratively and by the wood decomposition exotherm.

Process chemistry has been defined in two bench-scale (10 lbs/hr) equipments. The "minikiln", a 1' x 3' rotating autoclave, is a batch reactor, and provides data as a function of temperature. The "biogasser", a 2" x 10' auger reactor, is continuous, and provides integrated gas composition data. The minikiln has also given useful information on the exotherm and heat transfer.

Calculations have shown that the product gas, if raised to 1500° F and 200 psi, shifts to a H_2/CO ratio of 2/1, with only a small residual methane content. Condensation of this syn gas will yield methanol at an over-all energy efficiency (green wood to methanol) of 70-75%.

The process development unit (6 green tons/day), presently under construction, will be highly instrumented for complete process definition. Its mechanical features include a cylinder/lock valve feed, lock valve discharge, small diameter stuffing-box seals, and provision for steam as well as regenerative gas heating.

With operation of the PDU in the second half of 1980, development of the fuel gas process will have been completed. Syn gas shifting and methanol synthesis will be project tasks in 1981.

NOTES

CATALYTIC GASIFICATION OF BIOMASS

L. K. Mudge
L. J. Sealock, Jr.
R. J. Robertus
D. H. Mitchell
S. L. Weber

Pacific Northwest Laboratory
Richland, Washington 99352
Operated by Battelle Memorial Institute

ABSTRACT

Results of laboratory and process development unit (PDU) studies on gasification of biomass (wood) in the presence of catalysts are presented. Previous studies in this program have shown that alkali carbonates are effective catalysts for enhancing the steam gasification of wood. This paper reports on the use of an alkali carbonate catalyst in combination with commercial catalysts for the conversion of wood directly to a valuable gas product: methane or a synthesis gas.

Laboratory investigations have shown that use of an alkali carbonate catalyst in combination with commercial methanation and cracking catalysts gives a gas mixture with about 50 volume % methane and the remainder carbon dioxide. About 70% of the wood carbon is converted to gas. Operations in the PDU with the same catalysts, but different flow conditions, have given a gas mixture with about 25% methane. Operations are at atmospheric pressure and at temperatures from 450° to 700°.

The PDU and laboratory investigations have shown that catalyst combinations can be used that give a synthesis gas directly from steam gasification of wood with over 90% of the wood carbon converted to gas. The synthesis gas has a 2 to 1 mole ratio of hydrogen to carbon monoxide with the remainder of the gas carbon dioxide and about 1 volume % methane.

INTRODUCTION

Gasification of wood has been practiced for many years in fixed bed (also called moving bed) gasifiers. Since all of these gasifiers were air blown, the product was exclusively a low Btu gas with a heating value of about 150 Btu/ft^3. Wood gasifiers have been used in combination with the steam-iron process to generate indirectly a synthesis gas for fertilizer production [1]. This method for synthesis gas production was a very inefficient use of a wood resource.

The program described in this paper is concerned he direct conversion of wood (biomass) to le gas products. Combined catalyst systems are used for gasification of wood to generate the following gas products: 1) methane, 2) hydrocarbon or methanol synthesis gas, 3) hydrogen, 4) carbon monoxide, and 5) ammonia synthesis gas. Carbon dioxide is also generated with these products and can be removed by conventional techniques.

Laboratory investigations and process development unit (PDU) studies are in progress to obtain the information needed for evaluation of the technical and economic feasibility of direct conversion of wood to valuable gas products. Current results of laboratory studies confirm the technical feasibility of direct conversion of wood to methane and synthesis gases by proper choice of conditions and catalyst combinations. Studies are in progress for generation of other gas products.

Successful completion of these studies will allow use of biomass as a raw material for efficient conversion to convenient fuel forms and ammonia fertilizer. For example, biomass could be converted completely to methanol without generation of residues that may present a disposal problem in the biological conversion to ethanol processes.

Status of the laboratory and PDU studies is presented in this paper. The program has been in progress since December 1977 at a funding level of about $500,000/yr.

RESULTS OF LABORATORY STUDIES

Relative gasification catalyst activity, the effect of wood variety, the effect of gasification catalyst contacting method, and the effect of gasification catalyst concentration were reported at previous conferences [2,3,4]. These results were used to develop methods for generation of gases rich in methane and in hydrogen and carbon monoxide.

Experimental System

The continuous feed system shown in Fig. 1 was used to obtain data on generation of specific

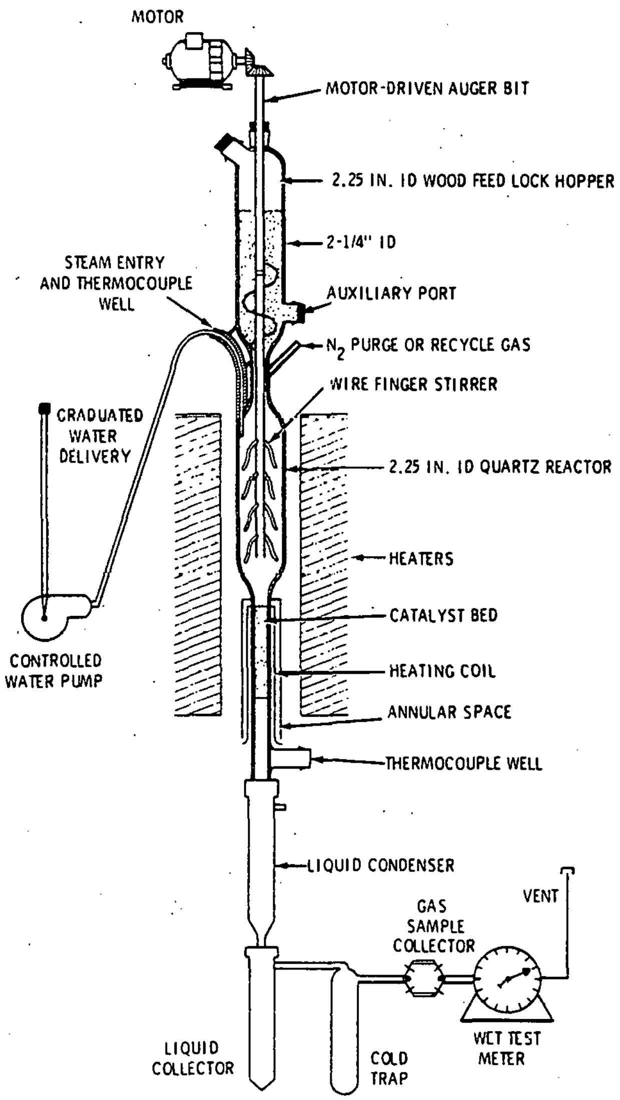

Fig 1. Continuous Wood Feed Reactor

products. The main portion of the continuous wood feed reactor was constructed of quartz. The reactor system was designed for atmospheric pressure operation and was quite flexible with respect control of experimental variables including: 1) wood feed rates, 2) steam flow rates, 3) temperatures up to 850°C, 4) auxiliary gas addition, and 5) product gas recycle.

Experiments were conducted by placing a weighed amount (usually 30-40 g) of impregnated wood into the wood feed hopper, which was purged with nitrogen to remove oxygen. The reactor was maintained at a predetermined temperature and purged with nitrogen before the desired steam flow was established. After steam flow was established, the wood feed system was started to initiate the experiment. Total gas volume generated and gas composition were determined during the experiment.

Wood samples were prepared by solution impregnation of alkali carbonate catalyst on predried wood samples. A concentration of 17 wt % K_2CO_3 (3×10^{-3} mole of K per gram of wood) was used throughout these studies.

Methane Generation Studies

Experiments were conducted to screen potential methanation catalysts with emphasis being placed on development of a consistent method for catalyst activation and on finding catalysts that do not require preactivation. Laboratory results obtained from these studies have produced up to 50% methane by volume in the product gas. Typical results are shown in Table 1 for gasification in the presence of 17 wt % K_2CO_3 primary catalyst.

Conditions must be carefully controlled to effect methanation. Thermocouples located in the catalyst bed indicated that the reactions in the bed gave a temperature gradient as high as 350°C between the inlet (650°C) and the outlet (300°C). This broad temperature gradient gave favorable yields of methane. The combined catalyst studies have shown that nickel, a well known hydrogenation and methanation catalyst, and silica-alumina (Si-Al), a highly acidic cracking catalyst, are an effective combination for methane production under the proper reactor conditions. It is postulated that a sequential or propagating type reaction mechanism is occurring. This mechanism might consist of a stepwise hydrogenation and

TABLE 1. LABORATORY RESULTS ON METHANE GENERATION

Reactor Temperature	550	650	740
Secondary Catalyst	Ni-3266	Ni-1404:Si-Al	Ni-1404:Si-Al
Catalyst Bed Temperature, °C	550	280-650	300-740
Wood Feed Rate, g/min	0.5	0.4	0.4
Steam Rate, g/min	0.1	0.1	0.1
Carbon Conversion to Gas, %	36	47	64
Btu Gas/Btu Wood	0.62	0.70	0.76
Gas Composition			
Volume % H_2	37	2.8	0
CO_2	36	50.2	48
CH_4	21	47.0	52
CO	6	0	0

cracking of the pyrolyzed wood products followed by methanation. Methanation would be the terminal step with no further cracking.

Because of the large temperature gradient in the catalyst bed, the effect of temperature on gas composition was studied with the different secondary catalyst systems. Each secondary catalyst system was tested at temperatures ranging from 250°C to 650°C. Poor methane yields were achieved at both extremes. The optimum methanation temperature for most of the commercial catalysts evaluated was about 350°C. Methane concentrations at this temperature were about 40% by volume.

Experimental results indicate that the addition of cracking catalysts, which were employed successfully when the catalyst bed temperature gradient ranged from 650-300°C, did not improve methane yields when the catalyst bed temperature remained isothermal at 550°C. For isothermal operation at 550°C Harshaw Ni-3210 and Harshaw Ni-3266 produced the highest methane concentration of the catalysts studied; however, the concentration was only 20% by volume.

A summary of the major findings of the laboratory scale methane studies is presented below:

- A combined secondary catalyst system of Harshaw Ni-1404 a/ and Strem b/ Si-Al at a weight ratio of 3:1 respectively produced the maximum methane yields.
- A catalyst bed temperature gradient ranging between 700° and 300°C from inlet to outlet was the optimum system evaluated for methane production.
- Optimum methanation temperature for most Ni catalysts tested was 300-400°C.
- Poor yields of methane were achieved with a catalyst bed temperature below 300° and above 550°C.
- Steam rates greater than 25 wt % of the input wood feed decreased methane concentrations significantly.
- Harshaw Ni-1404, Ni-3266, and Ni-3210 were the only catalysts that demonstrated methanation activity above 500°C.
- Laboratory results showed that for isothermal operation a maximum methane concentration of about 25% by volume is obtainable at a catalyst bed temperature of about 550°C at atmospheric pressure.

Hydrocarbon Synthesis Gas Studies

A number of catalyst systems have been evaluated for the production of hydrocarbon synthesis gas. Most experiments were conducted at temperatures between 750°C and 850°C with recycle of the product gas at a rate of 0.2 l/min. Product gas was generated at 0.5 to 0.7 l/min. Under these conditions, the wood and steam feed rates were optimized to produce a gas having a H_2/CO ratio of 2 to 1. Steam gasification of wood in the presence of a commercial Si-Al cracking catalyst produced a synthesis gas with a hydrogen:carbon monoxide ratio of 2:1 at both 750°C and 850°C. Use of a nickel reforming catalyst reduced methane concentrations below 1 vol %. Typical results are given in Table 2 for gasification in the presence of 17 wt % K_2CO_3 primary catalyst.

Major results of the hydrocarbon synthesis gas studies are:

- A Strem Si-Al cracking catalyst mixed with approximately 10% Ni catalyst proved to be a proper catalyst system for the production of a 2:1 H_2 to CO hydrocarbon synthesis gas.
- A reaction temperature of 850°C was the maximum temperature employed and was more effective than 750°C from both a gas composition and production standpoint.
- The concentration of CO in the final product gas increased more rapidly than the concentration of H_2 with increased temperature.
- The steam required to produce the desired 2:1 H_2 to CO ratio at a reactor temperature of 750°C was approximately 75 wt % of the wood feed.

TABLE 2. LABORATORY RESULTS OF SYNTHESIS GAS STUDIES

Reactor Temperature, °C	750	840	750	840
Secondary Catalyst	Si-Al	Si-Al	90% Si-Al, 10% Ni-1404	90% Si-Al, 10% Ni-1404
Catalyst Bed Temperature, °C	750	840	750	840
Wood Feed Rate, g/min	0.4	0.3	0.3	0.3
Steam Rate, g/min	0.3	0.3	0.4	0.4
Carbon Conversion to Gas, %	0.77	0.90	0.86	0.96
Btu Gas/Btu Wood	0.78	0.80	0.75	0.77
Gas Composition				
Volume % H_2	53	54	60	60
CO_2	12	16	18	19
CH_4	4	4	0.6	0.4
CO	30	27	21	21

---

a/ Harshaw Chemical Company, Cleveland, OH.
b/ Strem Chemicals, Inc., Danvers, MA.

- At 850°C the steam to wood ratio required to maintain the 2:1 H_2 to CO concentration was 1 to 1.
- The amount of CO_2 in the product gas decreased significantly as the temperature increased.

RESULTS OF PDU STUDIES

A schematic of the PDU reactor is shown in Fig. 2. The reactor is designed to operate at atmospheric pressure as a fluidized bed with agitation to mix the secondary catalyst and wood. Internal dimensions of the reactor are 11 in. in diameter by 8 ft high. The reaction zone height is 4 ft. Wood is metered with twin screws from a lock hopper into a high speed feed screw that injects wood into the reactor. The wood enters the reactor through the bottom feed port below the surface of the bed, which consists of either wood char or secondary catalyst mixed with wood char. The wood is gasified in the reactor where conditions can be controlled to produce the desired gas products. Electrical heating elements are used to control the reaction zone at temperatures up to 800°C. Gasifying agents include superheated steam, air, oxygen, carbon dioxide, mixtures of these gases, and recycled product gas.

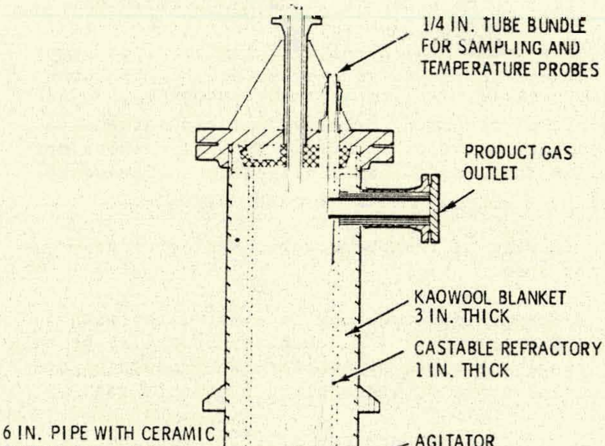

Fig. 2. PDU Biomass Gasification Reactor

The schematic of the PDU process equipment shown in Fig. 3 illustrates feed and product flow streams. The system is designed to generate data required for determination of mass and energy flows. An overall view of the facility is shown in Fig. 4. Initial operations have shown that the system operates reliably for testing conditions needed to generate methane and synthesis gas. The major problem to date has been failure of the internal electrical heaters, which has been circumvented by testing those conditions that give autothermal operation.

Preliminary operations to obtain baseline data on equipment performance and the effect of primary catalyst have been completed. These tests confirmed laboratory results that showed increased yields of gas with alkali carbonate catalyst present. Dry mixing of over 10 wt % catalyst with wood created problems with slagging in the reactor bed. Clinkers of alkali carbonate formed in the annular space between the agitator and the inner wall of the reactor. This problem is alleviated by keeping temperatures below 800°C and by impregnating the catalyst on the wood.

Testing of conditions for generation of methane and synthesis gas has been initiated. Preliminary results for synthesis gas generation are shown in Table 3. These data were obtained

TABLE 3. PDU HYDROCARBON SYNTHESIS GAS RESULTS

Temperature, °C	800	700	700	500
Wt % K_2CO_3	10[a]	10[b]	10[a]	.4
Wt Steam/Wt Dry Wood	1.2	0.48	0.42	2.8
Wt O_2/Wt Dry Wood	0.21	0.14	0.13	0.08
Dry Gas Composition				
Vol % N_2	1	0	1	5
CH_4	2	6	5	10
H_2	38	40	36	18
CO_2	39	36	40	49
CO	18	18	16	14
Carbon Conversion	0.98	0.90	0.92	0.70
Ft^3 Dry Gas/lb Dry Wood	30	18[c]	17[c]	36
Btu Gas/Btu Dry Wood	0.79	0.56[c]	0.45[c]	1.2
Electrical Energy Input, Btu/lb Dry Wood	0	0	430	960

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a) Impregnated on wood

b) Dry mixed with wood

c) Leaks in system may have caused erroneous v

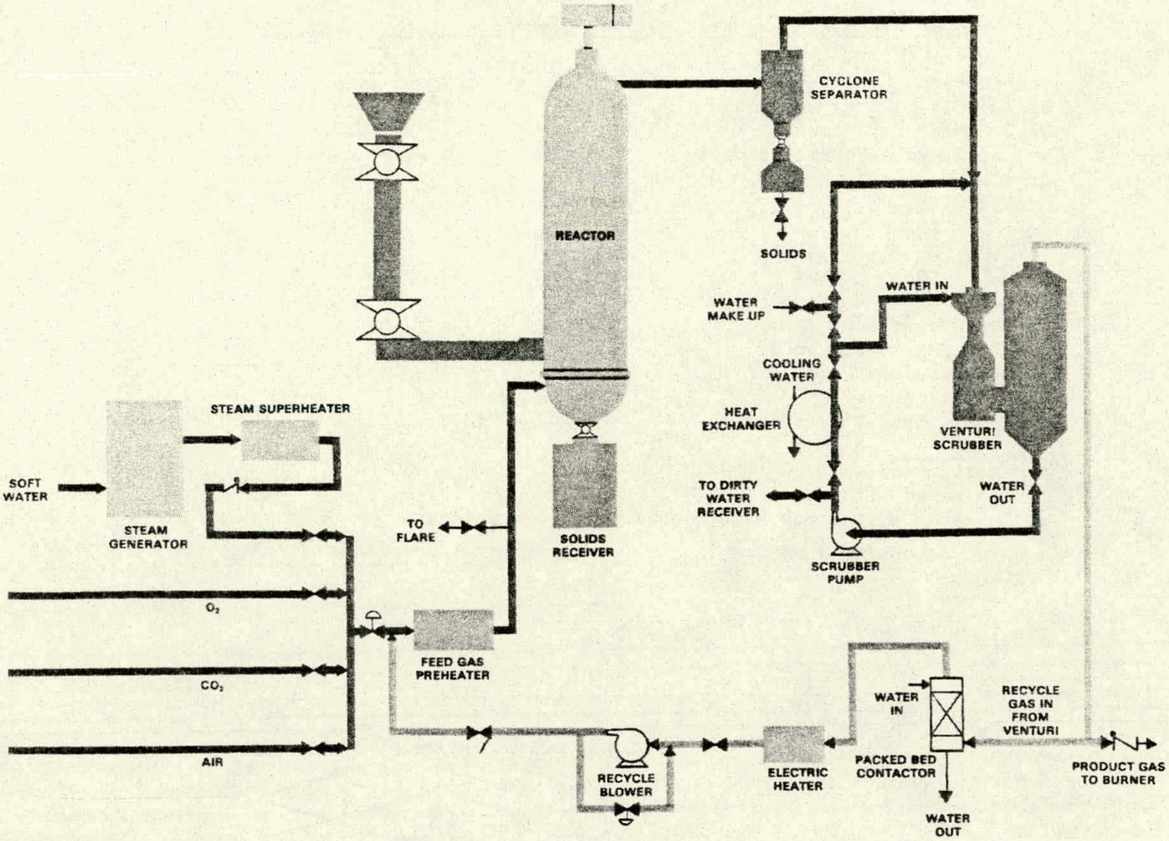

Fig 3. Schematic of PDU

with a silica-alumina cracking catalyst present in the reactor. We detected leaks in the gas outlet lines just prior to making the test at 800°C and suspect that the gas yield data for the other tests are erroneous. The data obtained at 800°C can be used to calculate the potential methanol yield from wood for these conditions. Results are shown in Table 4. The efficiency, defined as energy in methanol divided by total energy input, for generation of methanol from wood for these conditions should be about 0.5.

TABLE 4. POTENTIAL METHANOL YIELD FROM SYNTHESIS GAS PRODUCED FROM WOOD AT 800°C

Methanol Yield, lb/Ton Dry Wood	940
Btu in Methanol/Ton Dry Wood	9.2×10^6
Btu in Methanol/Btu in Dry Wood	0.54

Preliminary results for generation of methane are given in Table 5. The secondary catalysts used for methane generation were a nickel methanation catalyst, Harshaw Ni-1404, and a silica-alumina cracking catalyst in a 3:1 weight ratio respectively.

Comparing the gas compositions at 500°C in Tables 3 and 5 shows that the methanation conditions increase the methane concentration from about 10 vol % to 25 vol %. Methane yield from primary gasification under the conditions at 550°C in Table 5 is 74% of the total methane yield possible by completion of methanation of this gas composition.

We plan further studies on methanation in an attempt to improve conversion and yields. This may be done through operation at elevated pressures, or through use of improved high temperature methanation catalysts. Studies on methanation will resume after new internal heaters are installed.

TABLE 5. RESULTS OBTAINED WITH METHANATION CONDITIONS

Temperature, °C	550	500	500	500	450
Wt % K_2CO_3	10[a]	4[b]	4[b]	10[a]	10[a]
Wt Steam/Wt Dry Wood	0	0.2	0.7	0	0
Volume Recycle/Volume Product	5	6	12	12	7
Dry Gas Composition					
Vol % N_2	1	0.1	1	0.6	2
CH_4	26	21	24	29	34
H_2	26	29	25	19	12
CO_2	39	40	44	43	46
CO	9	9	6	9	5
Carbon Conversion	0.74	0.54	0.50	0.64	0.28
Ft^3 Dry Gas/lb Dry Wood	9[c]	8[c]	5[c]	7[c]	5[c]
Btu Gas/Btu Dry Wood	0.4[c]	0.3[c]	0.2[c]	0.3[c]	0.2[c]
Electrical Energy Input, Btu/lb Dry Wood	1100	1300	1140	1150	1210

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a) Impregnated on wood

b) Dry mixed with wood

c) Leaks in system may have caused erroneous values

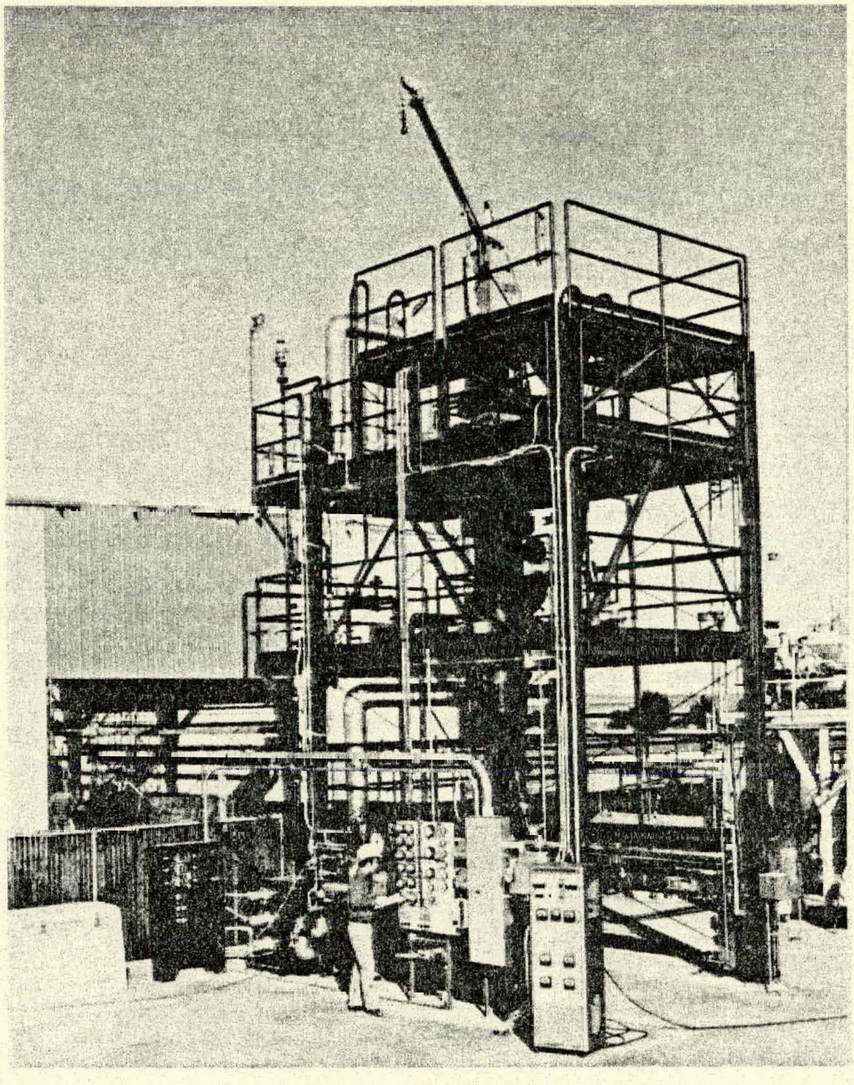

Fig. 4. PDU Facility

ACK EDGEMENTS

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Session V A

RESEARCH AND DEVELOPMENT OF INTENSIVELY CULTURED PLANTATIONS FOR MAXIMUM BIOMASS PRODUCTION

David H. Dawson
Program Manager
North Central Forest Experiment Station
P.O. Box 898
Rhinelander, Wisconsin

NTRODUCTION

aximum Yield Research started in 1971 by the orth Central Forest Experiment Station in reponse to concerns expressed by both industrial oresters and researchers. Industry realized hat new systems for growing more fiber on less and for less money must be explored because of scalating future fiber demands, increasing deands for recreation and nonforest uses of forest and, rising land prices and tax rates, and ineasing harvest and transportation costs.

esearch studies included selecting promising pecies or hybrids, establishing intensively ltured plots, exploring new harvesting methods, valuating pulping qualities of the fiber, and ssessing the economic viability of this system. he program uses a systems approach, and includes oth developmental and basic research, active ooperation with industries, and the collaboration f many scientists from different disciplines. he research team includes silviculturists, hysiologist, ecologists, geneticists, soil cientists, hydrologists, wood technologists, athologists, entomologists, engineers, and conomists.

nding from the Department of Energy since 1977 s enabled researchers to accelerate research d development in methods to establish trees and sure their survival. This is the most critical oblem in the intensive culture system because rger research studies or large-scale industrial antations cannot be established until the echnology to successfully establish hardwood ands has been developed. Although practices r planting softwoods were developed long ago, system for establishing large-scale plantations hardwood forest trees in the upper Midwest has een successful (except for the shelterbelt antings of the northern Plains). The problems periences in large-scale planting operations e often different from those normally considered small plot experiments, and the additional nds from DOE have enabled researchers to explore eans of plantation establishment in plots and to lp industry establish large operational plantgs.

[illegible]y, the intensive culture concept is that ee yield can be greatly increased by using genetically superior planting material and optimizing field environmental conditions. Generally, hardwoods have been used for intensive culture because of their early rapid growth and their ability to sprout from the stump after they are harvested, thus eliminating the need to replant the trees, but some studies involve conifers.

The establishment of intensive culture plantations in any region of the U.S. will include essentially the same kinds of operations, although the precise nature of these operations will vary greatly by species and site. Many of these practices are "agricultural" rather than "silvicultural" in nature.

First, the site must be prepared to eliminate other plants and to make it suitable for the operation of planting and cultural equipment. Seedlings or cuttings are planted with mechanical planting equipment at varied distances from a few centimeters to several meters in rows that are from one to several meters apart. The first growing season is critical to the survival of the small trees, and they must be kept free of weeds by cultivating, planting a cover crop between the rows, or applying an herbicide. Each growing season the trees are provided optimal amounts of nutrients and water by fertilizing and irrigating. Diseases and insect pests must be managed.

After presumably 6-10 years, at a time when the trees reach financial maturity and near-optimal biological growth, the trees will be harvested using specially-designed harvesters. The wood will be chipped and used for energy, chemicals, pulp, reconstituted wood products, feed, or any combination of these products. In the spring following harvest, with most species being considered for energy purposes, one or more new stems will sprout from the cut stumps, and these will become the next rotation of intensive culture trees.

RESEARCH WORK UNITS AND COOPERATORS COMMITTING RESOURCES TO THE R AND D PROGRAM

Currently, research in the R and D Program includes the following work units and cooperators:

Research Work Units

RWU NC-1112, "Intensively Cultured Plantations for Fiber and Energy Production." Project Leader: Dr. Jaroslav Zavitkovski

The involvement of the RWU includes the following research and development studies:

- --Selection and propagation of high yielding woody plants for intensive culture systems.
- --Establishment of promising species in plantations * including problems of site preparation and competition control.
- --Response of biomass and energy yields of selected species and species variants to spacing and harvesting cycles.
- --Response in terms of yields of selected species and species variants to nutrient and moisture levels and to soil types.

RWU NC-1152, "Physiology and Raw Material Evaluation of Intensively Cultured Plantations." Project Leader: Dr. Jud Isebrands

The involvement of this RWU includes the following research studies:

- --Relation of crown architecture of individual trees and canopy density of stands to the quantity and quality of yields.
- --Essential photosynthetic, respiratory, and other related physiological variables influencing the quantity and quality of yields.
- --Raw material quality of genotypes grown under intensive culture and evaluation of material as a furnish for specific end products.

FS-NC-2205, "Canker, Foliar, and Root Diseases of Forests and Christmas Tree Plantations." Principal Scientist: Mike Ostry

- --Diseases of trees grown with treatments to maximize fiber yield: Identification impact on yield and control.

FS-NC-2203, "Insects Affecting Forest Plantation Ecosystems in the NC States." Project Leader: Louis Wilson

- --Potential impact of insects on intensively cultured plantations grown for maximum biomass production and their control.

FS-NC-4252, "Methods for Evaluating Forest Resource Management and Use Alternatives and Organizing Information for Evaluations in Northern Forest Regions." Principal Scientist: David Lothner

- --The energy efficiency and economics of maximum yield silviculture.

FS-NC-3701, "Engineering Systems and Mechanizations for Northern Forest Stands." Principal Scientist: James Mattson

- --Design and modification of equipment for site preparation, planting, seeding, tending.

Major Cooperators

University of Wisconsin, Dept. of Forestry, Madison, WI.

- --Dr. T. T. Kozlowski - Leaf stomatal characteristics in relation to maximizing fiber yields.
- --Dr. Alan Ek - Growth projection and spacing analysis for high yielding silviculture.
- --Dr. Dan Benjamin - Determination of Poplar Borer in deterioration of Poplar.

University of Minnesota, Dept. of Forest Resources, St. Paul, MN.

- --Dr. Ed Sucoff - Characterization of *Populus* water balance in relation to soil moisture, potential evapotranspiration and stand quality.
- --Dr. Dietmar Rose - Economic analyses of intensive timber production systems.

University of Minnesota, Dept. of Forest Products, St. Paul, MN.

- --Dr. Ronald Neuman - Characterization of pulp and paper from woody material produced in intensive agronomic systems.

University of Minnesota, Dept. of Plant Pathology, St. Paul, MN.

- --Dr. F. A. Wood - Impact and control of *Melampsora* leaf rust on tree species used for intensive culture.

Michigan State University, Dept. of Forestry, E. Lansing, MI.

- --Dr. Wayne Myers - Impact of insects attacking trees planted for maximum fiber production.
- --Dr. Donald Dickmann - Patterns of diffusion resistance in *Populus* leaves.

Iowa State University, Dept. of Forestry, Ames, IA.

- --Dr. John Gordon - Effect of genotype on nitrate-reductase activity in hybrid poplar.
- --Dr. Greg Brown - Methods of rapid, early selection of poplar clones for maximum yield potential.

Institute of Paper Chemistry, Appleton, WI.

- --Dr. Dean Einspahr - Determination of the Kraft pulping properties of whole-tree jack pine and larch.

University of Illinois, Dept. of Forestry, Urbana, IL

- --Dr. Jeffrey Dawson - Establishment, early yi[illegible] and soil nitrogen contribution of selected A[illegible] *glutinosa* clones.

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* Partially financed by the Department of Energy.

The d Corporation, Escanaba, MI.

--Mr. Lynn Sandberg - Pilot studies-Mill Effluent as an irrigating medium.

Packaging Corporation of America, Filer City, MI.

--Mr. Mike Morin - Pilot studies-Establishment, culture, harvesting, utilization of intensively cultured plantations.

PROGRESS ON ESTABLISHMENT PROBLEMS

This report is intended only to cover progress in the research problem area "Establishment of promising species in plantations including problems of site preparation and competition control." It is recognized, however, that there is considerable interaction between research problem areas; invariably, progress and results in one affect the others.

The following report by H. A. McNeel, Research Agronomist, summarizes results of the 1978 studies.

Results of Populus Establishment Study, 1978, for Department of Energy

Site preparation, cover crops and weed control are very important factors that must be considered when any short rotation plantations are to be established. These practices will be ranked individually and then combined practice methods will be summarized. Refer to Table 1-A for 25 best combined treatments of the total 245 treatments.

Site Preparation: Moldboard plowing (9-10 inch depth), disking and spike-tooth harrowing was the best individual site preparation of the 7 treatments. Other site preparation treatments are listed in decreasing order in Table 1-C. Rototilling (6 inch depth) ranked second but it is not recommended due to the high cost of maintenance. Disking and harrowing might have been better if the disk had been heavier, to enable greater depth of working the soil, and in proper working condition. Rolling cultivation and no tillage might have had poorer growth and survival if liming and incorporating the lime into the soil by disking had been accomplished the previous fall (1977) rather than just prior to site preparation in the spring (1978). Survival was good on all site preparations but height growth decreased in the order of site preparations listed in Table 1-C from 50.4 to 39.8 cm.

Cover Crops: Legumes - Crown vetch and birdsfoot trefoil were the best cover crops (Table 1-B). White dutch clover and ladino clover was better than the control (no cover crop). During the first half of the growing season hairy vetch and field peas stimulated very good growth of the trees but the second half of the growing season they outgrew the trees and pulled them to the ground. T[illegible]s planted with hairy vetch had no die-back but [illegible]e was very serious die-back of the trees plant[illegible] with field peas. Hairy vetch was the only cover crop that adequately controlled the weeds without a herbicide. Non legume - Creeping red fescue was only slightly better than the control as measured by tree height growth and survival. With wheat, barley and rye survival was good but height growth of the trees was below that of the control.

Weed Control: Herbicide alone or cultivation. Amitrole was better (59.0 cm height growth and 97.6% survival) than Roundup (58.5 cm height growth and 96.7% survival). One caution: You have to be extremely careful when applying the herbicide because it may drift and damage the tree. Tenoran, rolling cultivation with hilling, linuron, rolling cultivation without hilling, sweep cultivation and rolling cultivation plus Roundup was next in order of listing and all were as good or better than the control. Princep and ravage caused severe damage to the trees.

Combination of Cover Crop and Herbicide: Treflan and balan, soil incorporated, in combination with any legume cover crop except field peas were equal to or better than the control. Crown vetch with treflan was the best of all treatments in both studies. Balan was the next best. Birdsfoot trefoil with treflan and balan ranked third and fourth. In no case where a non-legume conbined with diuron or bromoxynil was the treatment as good as the control. Refer to Table 2.

Clones and Planting Material: Clones 5263, 5272 and 5331 were the best for all three planting dates (May, June and July, 1978). Unrooted hardwood (soaked and unsoaked) cuttings had the best height growth and survival in the first planting (May). Linuron and Roundup were the best weed controls. Tefer to Table 3.

In the second planting (june) rooted hardwood cuttings in tubepaks were the best planting material. Linuron was the best weed control. In the third planting (July) all planting materials were good except for the tip cuttings. No weed control and ladino clover had the best height growth and survival. Therefore, when planting early (May) unrooted hardwood cuttings are the best planting material. Rooted hardwood cuttings in tubepaks are the best for June plantings and all except tip cuttings are good for July plantings.

Word of Caution: These results are only from one year's data and two or three years data is necessary for more accurate results. There was also a total of 25.03 inches of precipitation from May 6 to September 30, 1978. Years with less precipitation might have a very different effect of herbicides.

TWENTY-FIVE BEST INDIVIDUAL PRACTICES OF SITE PREPARATION AND WEED CONTROL FROM STUDY "A" (FIRST PLANTING) AND STUDY "B", CLONE 5260 SOAKED HARDWOOD CUTTINGS, 1978 RESEARCH

TABLE 1

SITE PREPARATION, COVER CROP AND HERBICIDE (active ingredient)	AVERAGE TREE HEIGHT (CM)	% SURVIVAL
Moldboard plow, disk, harrow, crown vetch w treflan 0.75#/A	91.7	100
Rototill, crown vetch w treflan 0.75 #/A	86.4	100
Chisel plow, disk, harrow, crown vetch w treflan 0.75 #/A	85.3	100
M plow, disk, harrow, birdsfoot trefoil w treflan 0.75 #/A	00.0	100
M. plow, disk, harrow, birdsfoot trefoil w balan 1.5 #/A	76.0	100
M. plow, disk, harrow, crown vetch w balan 1.5 #/A	75.2	100
Rolling cultivation, harrow, crown vetch w treflan 0.75 #/A	74.2	95.9
Subsoil, disk, harrow, birdsfoot trefoil w treflan 0.75 #/A	70.4	100
C. plow, disk, harrow, birdsfoot trefoil w treflan 0.75 #/A	69.6	95.9
M. Plow, disk, harrow, plus amitrole 2 #/A	69.5	100
Rolling cultivation, harrow, crown vetch w balan 1.5 #/A	68.3	100
C. plow, disk, harrow, crown vetch w balan 1.5 #/A	68.0	95.9
C. plow, disk, harrow, birdsfoot trefoil w balan 1.5 #/A	67.9	91.7
Subsoil, disk, harrow, crown vetch w balan 1.5 #/A	66.1	95.9
Rototill, crown vetch w balan 1.5 #/A	65.7	95.9
Rototill, birdsfoot trefoil w treflan 0.75 #/A	65.6	95.9
C. plow, disk, harrow, white dutch clover w treflan 0.75 #/A	65.5	100
Rototill, birdsfoot trefoil w balan 1.5 #/A	65.1	100
M. plow, disk, harrow, ladino clover w treflan 0.75 #/A	65.1	91.7
M. plow, disk, harrow, white dutch clover w balan 1.5 #/A	64.7	91.7
C. plow, disk, harrow, plus tenoran 5 #/A	64.6	95.9
Subsoil, disk, harrow, crown vetch w treflan 0.75 #/A	64.4	95.9
Subsoil, disk, harrow, white dutch clover w treflan 0.75 #/A	64.0	100
Rototill, white dutch clover w treflan 0.75 #/A	63.8	100
Disk, harrow, crown vetch w treflan 0.75 #/A	63.5	95.9
CONTROL (No herbicide, no weed control, no legume)	45.4	95.9

FOLLOWING APPEARED "X" TIMES IN 25 BEST TREATMENTS of total 245 treatments.

Moldboard plow - 7
Rototill - 5
Chisel plow - 6
Subsoil - 4
Rolling cultivation - 2.
Disk - 1

Crown vetch - 11
Birdsfoot trefoil - 7
Ladino clover - 1
White dutch clover - 4

Treflan - 14
Balan - 9
Amitrole - 1
Tenoran - 1

SITE PREPARATION ONLY (These results from Study "B" only)		
Moldboard plow, disk and spike-tooth harrow (9-10" depth)	50.4	90.4
Rototilling (6" depth)	47.0	89.2
Chisel plow, disk and spike-tooth harrow (12" depth)	45.0	86.8
Subsoil, disk and spike-tooth harrow (21-22" depth)	44.4	87.0
Disk and spike-tooth harrow (6" depth)	42.6	86.8
Rolling cultivation and spike-tooth harrow (½" depth)	42.4	86.4
No tillage	39.8	86.0

Note: The above site preparations are considering all herbicides, cover crops and other weed control methods used in the individual site preparation strips.

RESULTS OF COMBINATION OF STUDY "A" (FIRST PLANTING) AND STUDY "B", CLONE 5260, SOAKED HARDWOOD CUTTINGS, 1978 RESEARCH PLOTS (ALL SITE PREPARATIONS WERE COMBINED)

TABLE 2

WEED CONTROL TREATMENT	AVERAGE TREE HEIGHT (CM)	% SURVIVAL
Crown vetch with treflan 0.75 # a.i./Acre	74.6	98.2
Crown vetch with balan 1.5 # a.i./Acre	66.4	97.6
Birdsfoot trefoil with treflan 0.75 # a.i./Acre	65.5	98.8
Birdsfoot trefoil with balan 1.5 # a.i./Acre	60.8	97.6
Amitrole 2 # a.i./Acre	59.0	97.6
Roundup 2 # a.i./Acre *	58.5	96.7
White dutch clover with treflan 0.75 # a.i./Acre	58.2	95.2
White dutch clover with balan 1.5 # a.i./Acre	56.3	97.0
Tenoran 5 # a.i./Acre	55.1	96.4
Ladino clover with treflan 0.75 # a.i./Acre	53.2	92.3
Ladino clover with balan 1.5 # a.i./Acre	52.7	92.9
Rolling cultivation with hilling	52.4	97.0
Hairy vetch with balan 1.5 # a.i./Acre	52.3	95.2
Crown vetch with no herbicide	51.3	95.3
Birdsfoot trefoil with no herbicide	51.2	97.6
Linuron 2 # a.i./Acre *	50.3	96.7
Rolling cultivation without hilling *	50.1	93.3
Creeping red fescue with no herbicide	47.0	98.8
Ladino clover with no herbicide **	46.9	95.8
Field cultivation (sweeps)	46.3	95.2
Hairy vetch with treflan 0.75 # a.i./Acre	46.1	94.1
White dutch clover with no herbicide	45.8	94.6
Rolling cultivation plus Roundup 2 # a.i./Acre *	45.8	90.0
Control (no herbicide, no cultivation) ***	45.4	95.8
Field Peas with balan 1.5 # a.i./A	45.0	95.3
Creeping red fescue with bromoxynil 1 # a.i./Acre	45.0	97.0
Field peas with treflan 0.75 # a.i./Acre	41.1	91.7
Wheat with no herbicide	40.4	98.8
Barley with no herbicide *	39.6	100
Wheat with bromoxynil 1 # a.i./Acre	38.5	97.6
Field peas with no herbicide	35.4	91.1
Rye with diuron 3 # a.i./Acre	32.9	75.6
Rye with no herbicide	32.6	95.8
Wheat with diuron 3 # a.i./Acre	32.4	60.7
Hairy vetch with no herbicide	31.1	76.2
Rye with bromoxynil 1 # a.i./Acre	25.4	91.1
Creeping red fescue with diuron 3 # a.i./Acre	20.4	58.3
Princep 4 # a.i./Acre *	19.4	61.7
Princep 8 # a.i./Acre plus polymer	6.2	22.6
Ravage 1 # a.i./Acre	0.2	00.6

Treatments in Study "A" *

Treatments in both Study "A" and "B" **

Control (No weed control was attempted) ***

BEST CLONES, PLANTING MATERIAL AND WEED CONTROL, STUDY "A" 1978 DOE RESEARCH PLOTS

TABLE 3

CLONES TESTED	PLANTING MATERIAL	WEED CONTROL
5260	1- Unsoaked hardwood cutting	Princep 4 # a.i./Acre
5263	2- Soaked hardwood cutting	Linuron 2 # a.i./Acre
5272	3- Rooted hardwood cutting in tubepak	Roundup 2 # a.i./Acre
5326	4- Rooted tip cutting in hydroponics	R. Cult and Roundup 2 # a.i./Acre
5331		Rolling cultivation
		Ladino Clover
		Barley

FIRST PLANTING (May 10, 1978)

CLONE	PLANTING MATERIAL	LINURON 2 # a.i./Acre Ave. Ht CM	% Survival	Roundup 2 # a.i./Acre Av.Ht.CM	% Survival	R. Cult. and Roundup Av.Ht. CM	% Survival
5263	2	118.3	100				
5272	2	115.7	100				
5263	2			110.2	95.0		
5263	4			108.1	86.7		
5331	2			106.7	95.0		
5272	2			97.8	91.7		
5331	1					93.4	88.3
5263	1			92.6	90.0		
5331	1			90.8	95.0		
5331	4			90.6	76.7		

SECOND PLANTING (June 7, 1978)

CLONE	PLANTING MATERIAL	LINURON 2 # a.i./Acre Av. Ht. CM	% Survival	Rolling Cultivation Av.Ht.CM	% Survival	Ladino Clover Av.Ht. CM	% Survival
5263	3	112.1	90.0				
5263	1	101.7	100				
5263	2	99.3	96.7				
5272	1	99.2	93.3				
5272	3	96.9	90.0				
5331	3	86.7	88.3				
5331	3					80.7	93.3
5331	3			80.2	88.3		
5331	4			80.0	80.0		

THIRD PLANTING (July 13, 1978)

CLONE	PLANTING MATERIAL	No weed control Av. Tree Ht. CM	% Survival	Ladino Clover Av. Tree Ht. CM	% Survival
5331	3			67.0	90.0
5272	3	66.6	96.7		
5263	2	63.7	95.0		
5272	2	62.3	95.0		
5263	3			61.7	93.3
5272	1			59.9	88.3
5263	3	59.4	90.0		
5263	2			59.0	91.7
5263	1			56.9	86.7
5331	3	55.0	88.3		
5263	1	54.4	95.0		

INDU AL COOPERATORS

Problem:

Industry will ultimately determine the commercial success of intensive culture concepts. Some of the intensive culture techniques developed under research conditions may not be feasible in large industrial woodlands programs. Therefore, it is extremely important to cooperate with industrial foresters throughout the research and development effort.

Research Objectives:

The technical and economic feasibility of intensive culture cannot be determined by small field experiments such as those conducted at Rhinelander. Large plantations must be established to learn whether research results can be applied in industrial forest operations and to obtain information about the cost of operational intensive culture systems.

Research Results:

One of the pulp mills in the Lake States has been exploring the possibility of using mill effluent from a secondary treatment plant as an irrigating medium on trees growing on land adjacent to the mill. The company is interested in obtaining uniform high quality wood and reducing transportation costs, harvesting costs, and taxes on vast acreages of forest land.

Moreover, the company is concerned about developing a means of disposing of 30 million gallons of effluent a day, which is presently treated and then discharged into a river. One possible use for this effluent is irrigation of intensive culture trees. In areas of the United States where water is limited and industrial plants have to find a way to dispose of effluent, irrigation with treated effluent may be a solution to several problems.

A large-scale pilot study is in progress with this corporation to test the most intensive methods for growing fiber. A 40-acre site has been cleared, plowed, and tilled, and *Populus* cuttings will be planted at two spacings. An irrigation system to distribute the effluent will be installed to maintain an optimum moisture regime for the trees. In addition, a 1-acre test plot of willow (*Salix*) and several *Populus* clones has been established to measure the tolerance of these species to high levels of effluent and the effect of high levels of effluent on groundwater quality.

The groundwater at the site has been thoroughly mapped to identify depth, direction of flow, and chemical nature. Throughout the study it will be monitored to detect changes due to irrigation with the effluent.

separate effort more than 1,200 acres of us and *Alnus* were planted in 1978 by the ackaging Corporation of America near Manistee, Michigan, in cooperation with the Maximum Yield Program. Nurseries of superior clonal material were established and hardwood cuttings were planted in 3.3 x 3.3 m spacings. Agricultural techniques and equipment were used to mechanically prepare the site, fertilize and cultivate. Smaller acreages were established during the previous 3 years, with considerable failure due to stand establishment problems and weed competition. Successful cultural practices for many of these problems have gradually been worked out and survival of trees for 1978 was over 90%. It is expected that an additional 1,200 acres will be established in 1979.

NOTES

INTENSIVE PRODUCTION OF TROPICAL GRASSES

Alex G. Alexander
University of Puerto Rico, Rio Piedras, P.R.

Abstract

Tropical grasses are being evaluated as renewable energy sources by the University of Puerto Rico's Center for Energy and Environment Research. The research is progressing in three concurrent phases: Greenhouse, field-plot, and field-scale. Approximately 110 cultivars have been screened to date, with superior candidates emerging in the genera Saccharum, Pennisetum, Sorghum, Arundo, and Erianthus, and from both interspecific and intergeneric crosses. Three categories of tropical grasses have emerged based on the time required to maximize their dry matter yield: Short rotation species (2-3 months), intermediate rotation (4-6 months), and long-rotation (12-18 months). Superior candidates for these categories include Sordan 70A, napier grass, and sugarcane, respectively. A fourth category, minimum-tillage grasses for marginal lands, is being investigated with "wild" Saccharum species (S. spontaneum, S. sinense) and Johnson grass (Sorghum halepense) being the test species. Maximum dry matter yield to date was from sugarcane (27.5 oven-dry tons/acre in 12 months). Harvest machinery evaluations have begun on short-and intermediate-rotation species. Special emphasis is directed to the rotary-scythe and bulk-baler concepts using solar radiation as the principal means of moisture removal in the field.*

INTRODUCTION

This project was initiated June 1, 1979 as a contribution to the ERDA Fuels From Biomass Systems Branch and the biomass energy program of the UPR Center for Energy and Environment Research (CEER-UPR). It is directed toward sugarcane, tropical grasses related to sugarcane, and other tropical grasses, both domestic and wild, having large growth potentials on a year-round basis.

Primary objectives include: (a) Determining the agronomic and economic feasibility of mechanized, year-round production of solar-dried biomass through the intensive management of sugarcane and napier grass as tropical forages, and (b) examination of alternate tropical grasses as potential candidates for intensive biomass production. A secondary objective relates to the selection and breeding of new sugarcane progeny having superior biomass productivity as their main attribute.

Emphasis is directed toward a highly intensive and mechanized production of tropical grasses as solar-dried forages. This is a deviation from both conventional cane and cattle feed production in that total dry matter rather than sugar and food components is the decisive parameter. Management of production inputs-particularly varieties, water, nitrogen, and harvest frequency—will vary appreciably from established procedures. On the other hand, the production expertise and mechanization technologies that have been perfected within the sugar cattle forage industries is being utilized to the maximum extent possible for dry biomass production.

Optimization of production inputs for tropical grasses requires the identification of a few select clones from a large number of candidate grasses, and clarification of the conditions required for their production in an economically realistic operation. This is being done in three concurrent phases, including greenhouse, field-plot, and field-scale investigations (Table 1). A fourth phase, commercial-industrial operations, follows logically but lies beyond the scope of the present project.

CANDIDATE SCREENING

Greenhouse:

Studies are concerned with the screening of candidate clones plus the response of superior candidates to growth input and management variables. All plants are propagated either by sand culture or in 1:1 or 2:1 mixtures of soil and cachaza contained in 5-gallon galvanized cans. Sand culture offers precise control of water and nutrient variables. Soil-cachaza mixtures are very adequate and convenient media for determining relative growth rates between clones, and the annual growth curves, recutting tolerance, and responses to chemical growth regulators by candidates having superior growth potentials. All clones receive constant water and nutrient supplies in adequate amounts that are not rate-limiting for growth. In virtually all experiments the interspecific cane hybrid PR 980 has been used as a reference clone having recognized excellence as a high tonnage variety under PR conditions. Approximately 110 clones from 20 species have been screened to date. About half of these were in non-replicated "observation" trials.

Field-Plot:

By August of 1978, field-plot data has been gathered for sugarcane and other species over a time-course of one year. This included six 2-month harvests, three 4-month harvests, two 6-month harvests, and one 12-month harvest. Biomass yield data have shown the following trends: (a) Napier grass is superior to sugarcane when

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*Head, AES-UPR and CEER-UPR Biomass Energy Program, University of Puerto Rico, [illegible]yaguez Campus, Rio Piedras, P.R. 00928.

harvested at 2-, 4- and 6-month intervals, but sugarcane is superior when harvested once at 12 months; (b) sugarcane responds more readily to narrow row centers than napier grass, but decreasingly so with advancing age; (c) biomass yields increase with decreasing frequency of harvest; (d) optimal biomass production is a disphasic process, requiring an initial rapid tissue expansion which is highly visible but consists mostly of water, followed by a fiber accumulation phase involving little outward change in the plant's appearance; and (e) fertilization (N-P-K) treatments based on conventional sugarcane and cattle forage production requirements are inadequate to sustain maximum biomass yields.

Four categories of candidate tropical grasses have emerged to date having distinct periods of land occupancy and time intervals required for maximum dry matter production (Table 2). The first category, or "short rotation" plants would have only a brief occupation of land otherwise committed to conventional food or fiber crops (Table 3). Sordan 70A, a sorghum x sudan grass hybrid developed by the Northrup-King Company, is an outstanding short rotation candidate (1). Recent data indicate that Sordan 77 is an even better candidate owing to greater DM production and drouth tolerance (2).

A second category includes forage grasses that will establish quickly and withstand recutting for 8 to 18 months. Energy is the predominant agricultural commodity in this instance. Napier grass (Pennisetum purpureum) is the outstanding tropical grass in this category at the moment (Table 3). A third category consists of very durable grasses that will withstand repeated harvests for up to five years. These are long-rotation species that will only rarely give up their site to vegetables or other food crops. Sugarcane requiring about a year to maximize its dry matter, yet having a potential to produce up to five additional ratoon crops, is the superior type crop for this category.

A fourth category, the "low till" or minimum tillage grasses (Table 2), will produce at least moderate yields with the barest minimum or production inputs. This requirement is underscored by the lack of water in large areas of the United States and Puerto Rico, and by the fact that economic considerations will not always permit a maximum expenditure of production resources even when these are otherwise available. The outstanding minimum-tillage candidates to date are the "wild" Saccharum clones US 67-22-2 and SES 231 (Table 3). Johnson grass (Sorghum halepense) has also shown promise in this category (2).

MAXIMUM YIELDS

Optimal yields for the first year's harvests were clearly a function of species category and harvest frequency. The "fastest growing" grass studied to date to Sordan 70A, but only in a comparatively short growth period of 10 to 12 weeks duration. Sordan 70A is easily exceeded by napier grass in a growth span of 4 to 6 months, and both species are exceeded by sugarcane in growing periods in excess of 6 months (Table 4).

As indicated by data in table 4, frequent recutting of sugarcane does not increase yields, and in fact it greatly depresses yields. Once planted, sugarcane should be [illegible] standing for at least 12 months to maximize total dry matter. This has important implications for the mainland US where even the southernmost cane regions do not have a year-round growing season. In this respect the recent emphasis on sweet sorghum by Battelle-Columbus workers (3) is a correct decision for the production of fermentable solids on the US mainland.

Narrow row spacing for sugarcane (50 cm as opposed to a standard 150 cm) failed to increase dry matter in two of three varieties tested. This is consistent with Battelle-Columbus contractor findings in Florida, but is not consistent with their findings in Louisiana where increases in the order of 40-50% were obtained (3,4). Possible explanations for these divergent results are detailed elsewhere (1). Alternately, the collection of sugarcane "trash" (leaf and leaf-sheath tissues which have desiccated and detached from the cane stalk) gave significantly higher yields (4.5-6.0 tons/acre) for each of the sugarcane varieties tested to date. Traditionally, such materials have been burned by extensive field fires shortly before the harvest of conventional sugarcane, that is, of cane grown for sucrose and molasses (5, chap. 13).

The maximum dry matter yield obtained to date (27.5 tons/acre) was produced by the sugarcane variety NCo 310 propagated at a narrow row center over a time-course of 12 months. This value is approximately 3 times that of the commercial sugarcane industry in Puerto Rico. However, this is far from the projected total yield capability for sugarcane. Ultimate maximum yields, in the order of 50 dry tons/acre year, will require "new generations" of sugarcane and the propagation of ratoon (regrowth) plants for several years after a given crop of sugarcane is planted (Table 5).

MECHANIZED HARVESTING

Initial tests of biomass harvest equipment were performed during the spring of 1979. Special attention is given to the harvesting of thin-stemmed grasses such as Sordan 70A, napier grass, and "wild" forms of sugarcane (S. spontaneum, S. sinense). These crops will be handled as solar-dried forages. They will be dried and baled in the field and stored for later use as boiler fuels during Puerto Rico's rainy months (August-December) when sugarcane bagasse is not available as a fuel source. Principal machinery items include a model 8700 Ford Tractor (120 hp), a Mathews Company rotary scythe (model 9-E, with 9-foot mowing swath), and a New Holland model 851 Round Baler.

Preliminary tests were performed with the rotary scythe using wild Johnson grass (Sorghum halepense) as the test material. This implement does not cut or mow grasses as does a conventional sickle-bar mower, but rather breaks off and "conditions" the grass with a series of steel plates rotating at high speed with extremely powerful force. The rotary scythe is a thoroughly rugged machine. Relatively few factors can inhibit its performance short of an inadequate power supply (tractors having less than about 90 hp), or [illegible] encountering of plant materials or sufficient mass [illegible] stop the blades or the tractor engine.

No [illegible]iculty of any kind was encountered in the first tri[illegible]th Johnson grass. This material amounted to roughly 10 to 12 green tons/acre. The rotary scythe was moved to a second field where Johnson grass had grown wild for several years. The implement performed quite adequately, with the exception of "heavy" areas where accumulating dead Johnson grass had formed mats approximately 2 to 3 feet thick. In such areas the mats sometimes tended to push ahead of the implement rather than pass under it in contact with the rotating blades. It should be noted that the rotary scythe is designed to function most effectively on individual plant stems. The stems would preferably be upright but the rotary scythe is also effective on lodged material. There is a strong possibility that it will perform adequately in formal trials on Sordan 70A and napier grass.

All of the materials that were harvested (conditioned) with the rotary scythe were effectively solar-dried within 2 to 3 days. The drying process was assisted with a side-delivery forage rake. The Round Baler performed quite effectively on these materials, producing bulk bales weighing in the order of 1200 to 1500 pounds.

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TABLE 1. RESEARCH PHASES FOR BIOMASS PRODUCTION STUDIES WITH TROPICAL GRASSES

Research Phase	Class of Objectives
Greenhouse	Physiological-Botanical
Field Plot	Botanical-Agronomic
Field Scale	Agronomic-Economic
Commercial-Industrial	Economic

TABLE 2. CATEGORIES OF CANDIDATE TROPICAL GRASSES

Cropping Category	Production Interval [1] (Months)	DM Max. [2] (Months)
Short Rotation	4-6	2-3
Intermediate Rotation	8-18	4-6
Long Rotation	36-60	12-18
Minimum Tillage	Indeterminate	(?)

1/ Replanting frequency; at least one ratoon crop is anticipated.

2/ Time required physiologically to maximize dry matter.

TABLE 3

CANDIDATE TROPICAL GRASSES

SHORT ROTATION	INTERMEDIATE ROTATION	LONG ROTATION	MINIMUM TILLAGE
Sordan 70-A	Napier Hybrid 7350	*Saccharum* Hybrids	*Saccharum spontaneum*
Sordan 77	Napier Hybrid 30086	NCo 310	US 67-22-2
Trudan 5	Common Napier	PR 980	US 72-70
Millex 23	NK Hybrids	B 70-701	SES 231
Bermuda Grass (Pasto Rico)	*Saccharum spontaneum*	AES-UPR Breeding Selections	*S. spontaneum* Hybrid (Wild)
NK Hybrids	US 67-22-2	Napier Grass	*Arundo donax* (Wild)
Roma	US 77-70	Intergeneric Hybrids	*Panicum* Bunch Grasses
	SES 231		Millex 23
	S. spontaneum Hybrid (Wild)		NK Hybrids
	Intergeneric Hybrids		Napier Grass
			Intergeneric Hybrids

☐ Leading Candidate

▣ Confirmed Source

FIGURE 10. Categories of tropical grasses and leading candidate clones under investigation as renewable energy sources in Puerto Rico.

TABLE 4. HARVEST FREQUENCY VS YIELD FOR SUGARCANE AND NAPIER GRASS PROPAGATED OVER A TIME-COURSE OF 12 MONTHS

Harvest Interval (Months)	Total Yield (Tons DM/Acre) For —	
	Sugarcane 1/	Napier Grass 2/
2 (x6)	6.5	12.7
4 (x3)	11.1	22.6
6 (x2)	16.6	25.6
12	25.5	19.3

1/ Mean values for three varieties and two row spacings.

2/ Mean values for one variety and two row spacings.

TABLE 5. SUGARCANE YIELD POTENTIALS; PRESENT AND FUTURE HYBRID

Hybrid Category	Crop	Principal Attributes	Total Biomass (Oven Dry Tons/Acre Year)
Existing Generation	Plant	Sugar-Fiber	27.5 1/
	Ratoon	Sugar-Fiber	33 2/
Second Generation	Plant	Biomass-Fiber	30-35 2/
	Ratoon	Biomass-Fiber	35-40 2/
Third Generation	Plant	Biomass-Fiber	40-45 2/
	Ratoon	Biomass-Fiber	45-50 2/

1/ Yields already attained (2)

2/ Projected yields.

ENERGY AND CHEMICALS FROM WOODY SPECIES IN FLORIDA

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Wayne H. Smith
Center for Environmental and
Natural Resources Programs
University of Florida
Gainesville, FL 32611

Louis F. Conde
School of Forest Resources
and Conservation
University of Florida
Gainesville, FL 32611

ABSTRACT

The project is attempting to assess the potential of six native and exotic woody species in Florida for the sustained production of fuel and chemicals as alternative sources to fossil fuels and petrochemicals. Included in this assessment will be evaluations of the amount of land available for production, effects on production of various cultural options, potential site effects of biomass-farming on water use and nutrient conservation, and energy input/output efficiencies. Test plantings for each species will be installed this year.

Preliminary findings indicate that commercial: private ownership ratios vary from 1:3 in northern Florida to 3:1 in southern Florida. Superior selections of sand pine, slash pine, and Eucalyptus have been made from numerous existing progeny tests. Selections of promising trees of Melaleuca and Casuarina have been made from natural stands in south Florida. Wood and foliage heat values are each relatively uniform across species but bark heat value is quite species dependent. Mean annual increment of aboveground dry biomass of trees in existing stands ranges from 2.7 to about 7 tons/acre/year.

DESCRIPTION OF TASK

Production of biomass and silvichemicals for energy generation and/or petro-chemical substitutes shows great promise in Florida. Florida climate is well suited for producing biomass. In certain areas three agricultural crops can be harvested each year. Some of these crops produce large biomass yields but require such large subsidies of energy that net energy yields are small. In contrast, woody species, both native and exotic, exhibit rapid growth in the sub-tropical environment and require less energy subsidization. The unique site conditions in Florida, especially in southern Florida, make it unlikely that species suitable for biomass farms in other parts of the country would be suitable for Florida. In addition, the desire to keep costs of transporting the harvested biomass to the conversion facility to a minimum emphasizes the need to utilize species that are well suited to local conditions.

[illegible] of the more promising candidates are native [illegible]es of Pinus and the introduced Casuarina, Eucalyptus, and Melaleuca. These species grow rapidly on the infertile sandy soils and in the warm, moist climate of Florida. Other more temperate woody species either do not grow as well in the Florida climate or require the more fertile soils which have higher priority for agricultural production. The species we choose to test have the common characteristic of good performance on soils not prized for agricultural production.

In essence, our task is to assess the potential for sustained production of biomass by these species and of selections within species. We are attempting to make a realistic appraisal of the amount and type of land actually available for biomass production in Florida. Effects on rates of production of various cultural treatments will also be evaluated. In addition, site effects of biomass-farming on water use and nutrient balances will be assessed. Final results will include evaluation of rotation age and energy input/output efficiencies.

OBJECTIVES

The general objective, as stated, is to determine the potential for producing biomass and chemicals from selected woody plants in Florida. In order to determine this potential, several program objectives need to be addressed:

1. Rotation age--time interval for maximum production under an array of treatment-species-land combinations.

2. Management intensity alternatives for species by land type.

3. Economic alternatives for optimizing output with various inputs to be most cost effective.

4. Net energy efficiency using simulation modeling procedures.

Specific objectives that will be assessed within these program objectives are:

1. Determine the amount of land potentially available in each site class for biomass/

chemicals production.

2. Determine the species, selections, or genotypes most suitable for northern and southern Florida.
3. Determine levels of production as a function of site preparation, spacing, fertilizer requirement from commercial or waste sources, rotation age.
4. Determine energy and chemical content of biomass produced.
5. Determine water and nutrient use by the candidate species.
6. Determine energetic and economic input during site establishment.

APPROACH

The program of work requires a broadly interdisciplinary approach. We recognize that full evaluation of several species for biomass and silvichemicals depends on the stand establishment procedures employed and the cultural treatments applied. With coppicing species, harvesting sequence also influences - and is influenced by - the biological response to treatment. For these reasons a general approach to evaluating woody biomass production was chosen. Thus it is essential to involve disciplinary specialists working as an interactive team (Appendix I).

The specialists involved have competencies in land resources inventory, forest genetics, wood science, silviculture, forest nutrition, forest soils and hydrology, and forest botany. A first step in the effort is for the inventory specialist to develop reasonable estimates of land available for each of the species and to receive input from the other specialists in establishing the site productivity classifications and their dispersion. Concomitantly, the forest geneticist will screen existing progeny tests and natural stands for those selections that show much promise for rapid growth in a biomass farm. The wood scientist and forest botanist will commence the evaluation of silvichemical content and energy values of plant parts. Candidate species will then be placed under culture and given treatments by the silviculturist with the assistance of specialists in forest soils and hydrology and forest nutrition.

At the time of establishment of the test plantings, procedures will be employed that will allow assessment of water and nutrient use. During forest growth and plant development, the geneticist will evaluate differential responses by selections and the biomass and silvichemical teams will continually monitor energy and chemical yields. All operational costs will be recorded and energy outputs evaluated continuously. We believe that five years or more will be required for ample data acquisition to reach a complete rotation economic and energetic assessment.

The experimental field design will first dra species that show both biomass and silvichemical potential and are adapted to the range of sites in Florida. Certain existing provenance trials and progeny tests will be first evaluated for candidate selection for biomass farm testing. Selections of Melaleuca and Casuarina will be from natural stands since no young plantations exist. Slash pine, sand pine, and certain species of Eucalyptus will be tested in north Froida where frost is common in winter. In south Florida emphasis will be on Melaleuca, Casuarina and other species of Eucalyptus. Sites will be located on university or private lands that will provide security and ample time for undisturbed experimentation.

The experimental design provides for two basic types of testing: (1) a test of selections, and (2) a test of management practices. The selection testing within a species will all be carried out at one site preparation and fertilizer level. Three types of selection testing will be conducted: many selections at one spacing, few selections at three spacings, and many selections at variable spacing. The combination of the three tests should provide a very firm base for determination of selection x density interactions. Management practices such as site preparation level and fertilizer levels will be tested at one spacing using one selection. Chemical and sewage fertilizers will be tested. A minimum of three replications are proposed for all tests.

Program goals will be realized as data are accumulated and synthesized. The convergence of the data from individual projects will generate answers to program level questions such as:

1. Production potential of each species and the land area potentially available for biomass farms of each species.
2. Management alternatives selected from interacting sets of choices.
3. Rotation age or time required for each species to produce maximum biomass or silvichemicals under an array of treatment options.
4. Output of usable energy for the least energy subsidy or the most cost effective approach to biomass farms.

From this synthesis, feasibility specifications fo biomass farms to produce fuel for energy and silvichemicals for petrochemical substitutes should be possible. For each Florida region the proper species or selection and the most energy effective combinations of establishment and cultural treatments can be chosen. This should derive from biological production, balance of economic and energy inputs, water use and nutrient conservation.

RESULTS

The major results for the program to date are given here for each project within the program. It must be emphasized that results within any given project are often derived from the joint efforts of personnel of several projects.

Land Assessment

Procedural strategy is briefly described as follows: the state has been divided into two areas; 1) a northern section in which slash pine, sand pine, and Eucalyptus would be grown, and 2) a southern section in which Casuarina, Eucalyptus, and Melaleuca would be grown.

A simple random sample, with a sampling intensity of 10%, is being used to assess land potential. The basic sample unit is the township, a unit 6 miles on a side. For each of the 154 townships selected, land use maps and forest soil potential maps are being manually combined. For each township, the acreage in each combination of land use and soil potential can be determined for slash and sand pine. Thirty counties currently have had land use/soil potential maps combined.

Growth requirements of the introduced species will be compiled from extant and ongoing silvicultural work. For each soil series, individual soil properties will be assessed as to poor, fair, or good suitability for Casuarina, Eucalyptus, and Melaleuca. Multiplying the weighting factors for each soil property will result in an overall quantitative soil series rating, which can be used to determine high, medium, or low productivity.

In addition, ownership patterns are being studied in an attempt to determine what proportion of land that is potentially suitable for biomass production would actually be available for production. Of the 29 available county plat books, all have been mapped to a resolution of 10 acres by land ownership category. Preliminary findings show that, on average, northern Florida counties have a commercial: private ownership ratio of 1:3, while central and southern counties have ratios of 1:1 and 3:1 respectively.

Selections

Genetic improvement of any forest tree species is long established and often quite substantial. Current achievable gains in volume production of loblolly pine are as high as 25%, and a 50% gain in the second-generation is expected. Similar volume improvement of slash pine is anticipated. Forestry firms in the south anticipate a 22% gain due to using genetically improved planting stock [3]. An economic evaluation of genetic improvement of loblolly pine suggests a 30% increase in present net value [4]. Substantial differences in oleo-resin yields and growth of three year old [illegible]h pine grown at a 1 1/2 x 3 foot spacing have been detailed [7].

Sand pine. All appropriate existing stands of Choctawhatchee sand pine have been surveyed, and those progenies showing potential for biomass plantings have been identified by comparison of growth performance, wood properties, and energy yield. Seed from 24 superior progenies have been obtained and have been sown in the U.S. Forest Service nursery on the Chipola Experimental Forest.

Slash pine. Several existing high density plantings of slash pine, involving over 300 progenies, have been reviewed. Progenies having superior growth and energy production have been selected, and the seed of 40 selections has been collected and sown in the nursery.

Eucalyptus. To our knowledge, no high density plantings of Eucalyptus exist. Thus we were unable to evaluate performance of existing selections in anything approaching biomass plantation conditions. With the aid of the U.S. Forest Service, 33 of the best sources of E. grandis currently available in terms of coppicing ability and growth have been selected and seed sown in the nursery. Similarly, with the aid of the N.C. State Univ. Cooperative Tree Improvement Program, 10 cold hardy sources of E. viminalis have been selected and their seed sown in the nursery.

Melaleuca and Casuarina. Since little or no selection work had been done on Melaleuca or Casuarina, individual trees of these biomass candidates were selected in natural stands primarily on the basis of growth habit. Wood properties and energy yield of all trees selected are being determined. As a result of the field work, 36 selections of Melaleuca and 26 selections of two species of Casuarina are now being grown in the nursery.

Energy Content. The characterization of the fuel and silvichemical properties of the biomass candidates has contributed to the selection process, particularly for slash and sand pine where a large number of potential sources exists. Availability of these sources allows screening for desirable fuel and biomass properties.

In the case of Eucalyptus, Casuarina, and Melaleuca, basic determination of wood properties had to be made. The wood heat values of all of these species were similar (8200-8700 BTU/lb) and typical of those for domestic hardwoods. Heat values for Melaleuca bark were much higher (10,500) while Eucalyptus bark was much lower (7000). Foliage heat values were higher than the wood heat values for each species, ranging from about 8500 to 9300. The specific gravity of Casuarina wood (.7) was much greater than any of the other species tested, which generally ranged about .5 [1].

Silvichemical work has concentrated on Melaleuca to date since some species are known to be rich in extractives [5,6]. It is generally believed that certain of these extractives affect the heat of combustion values. Heat values and extractive contents have not been reported for trees grown in Florida. Such information is basic in order to determine how the effects of treatments or selections can increase extractives.

Some silvichemical properties of Melaleuca have been determined. The quantity of extractives, using 6 organic solvents, in the sapwood was limited. The heartwood contained approximately 5 percent less heat value after extraction compared to unextracted controls. Large amounts of extractives are found in Melaleuca bark, including polyphenols and fatty substances. Bark heat value was reduced by ten percent after extraction.

Biomass. The primary objective of the past year was the development of preliminary regression equations which would allow the prediction of biomass by component for young trees grown at close spacing. This has been largely accomplished. The allometric regression has proven to be the model of choice, where

$$\ln Y = b_o + b_1 \ln X$$

and ln is natural logarithm, Y is the dry weight in grams of the biomass component, b_o and b_1 are regression coefficients estimated from the data, and X is the independent variable, in the present case, either diameter breast high (dbh) or diameter breast high times tree height. This regression form has worked well in the past [2,8] and has the added advantage of requiring few measurements in the field in order to estimate standing biomass.

In the case of the sand pine plantations measured, a stratified random sample of 9 or 10 trees has been sufficient to develop regressions for bole dry weight with a good fit ($r^2 \geq 95$) using dbh alone as the independent variable (Appendix II).

Slash pine plantations have been measured in the same fashion but regressions have not yet been developed.

We have been unable to develop regressions for young, closely spaced Eucalyptus since we have been unable to locate any such stands that we could destructively sample. Similarly, no plantations exist of young, closely spaced Melaleuca or Casuarina. In the latter cases, natural stands of indeterminate age and variable spacing were sampled to check the feasibility of developing regressions. In both cases good fits ($r^2 \geq .88$) were obtainable using dbh, height, or dbh times height as the independent variable (Appendix II).

Once obtained, the regression equations can be used to estimate stand biomass and, if stand age is known, mean annual increment. In the case of the sand pine, when both spacing and age are known, the mean annual increment of three stands measured ranged from 2.7 to 4.5 dry tons/acre/year (6.1 - 10.2 mt/ha).

Yield figures from the non-uniform natural stands of Melaleuca and Casuarina must be interpreted cautiously. However, biomass data from relatively uniform seedling stands (average height 4.9 feet = 1.5 m) indicate a total above-ground dry weight yield of 6.1 tons/acre/year (13.8 mt/ha) if the stand was two years old or 4.1 tons/acre/year (9.3 mt/ha) if the stand was three years old. Similarly, a stand of larger trees (average height 20 feet = 6.2 m) had a yield of 7.5 tons/acre/year (17.0 mt/ha) if five years old or 6.2 tons/acre/year (14.1 mt/ha) if six years old. A mixed age stand of Casuarina was estimated to be producing at the rate of 4.8 tons/acre/year (10.9 mt/ha). For both these species, yield figures need to be verified in biomass plantings.

Silviculture. Locations for five test plantings have been established throughout the state from the panhandle in the northwest to south peninsular Florida in the virtually frost-free zone. Site preparation activities are underway at all sites.

Water use. Considerable testing has been done to determine the feasibility of various types of lysimetric installations to monitor water use. One of the more practicable types involves surrounding a small plot with an impermeable barrier from the soil level down to the underlying nearly impermeable substrate. For deep soils or soils not underlain by an impermeable layer, tension lysimeters have been determined to be the most feasible option.

Synthesis is just beginning as data on energy inputs associated with site preparation, plantatio establishment, and cultural treatments are startin to accumulate. Data from this program as well as relevant data from related activities will be formulated into a simulation model. We propose to us the model as a basis for predicting growth rates o the species of trees being tested for biomass plantations. Possible effects of alternative site treatment and harvesting schemes will also be explored. Subsequent phases will allow optimizing economic and energy rotations when biomass/silvichemical values are discounted for expenditures of dollars or energy.

FUTURE PLANS

The most important development in the near future will be the actual installation of the test plantings this summer and winter (1979). After installation, biomass production and performance of individual selections will be monitored on at leas an annual basis. In addition, we intend to get an early assessment of the coppice production of Melaleuca by installing test plots in existing natural stands. Thus we should have an indication of the amount of coppice regrowth to expect as wel as any problems that might arise in coppice formation before it is time to harvest the biomass test plots.

Work will continue on the energy content of extractives from the biomass candidates as well as refining the data on wood properties, including drying rates. As the plots are harvested, the fue quality and the amount of the biomass produced wil be measured. During development of the test stands, water use, nutrient content, and nut[illegible] input/output flows will be monitored. We als[illegible] intend to have an increasingly accurate handle on l [illegible] availability as that data base is expanded d [illegible] g the next year.

Continuous development of the simulation model is expected throughout the course of the study. Eventually it is anticipated that final input-output analyses and projections will be made that will allow selection of the best species-management options for given site conditions.

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APPENDIX I

Research Team

Dr. Loukas G. Arvanitis - land resources inventory
Dr. Louis F. Conde - forest botany
Dr. Katherine Carter Ewel - systems ecology
Dr. Richard F. Fisher - silviculture
Dr. Jacob B. Huffman - wood science
Dr. Hans Riekerk - forest soils and hydrology
Dr. Donald L. Rockwood - forest genetics
[illegible] Wayne H. Smith - forest nutrition, forest soils
Dr. Edward T. Sullivan - forest economics
Dr. Shih-chi Wang - wood chemistry

APPENDIX II

Regression equations

Sand Pine 6.5 year old stand; .9m x 1.2m spacing; $2.0 \le \text{dbh} \le 7.1\text{cm}$

$\ln \text{bole dry weight} = 4.8082 + 1.8667 \ln \text{dbh}$
$\ln \text{branch dry weight} = 3.8195 + 2.0520 \ln \text{dbh}$
$\ln \text{foliage dry weight} = 4.4485 + 1.7250 \ln \text{dbh}$

12.5 year old stand; .9m x 1.2m spacing; $2.3 \le \text{dbh} \le 9.1\text{cm}$

$\ln \text{bole dry weight} = 4.6039 + 2.1794 \ln \text{dbh}$
$\ln \text{branch dry weight} = 3.0945 + 2.2380 \ln \text{dbh}$
$\ln \text{foliage dry weight} = 3.0025 + 2.1630 \ln \text{dbh}$

17.5 year old stand; .75m x .9m spacing; $3.8 \le \text{dbh} \le 11.9\text{cm}$

$\ln \text{bole dry weight} = 4.3083 + 2.5245 \ln \text{dbh}$
$\ln \text{branch dry weight} = 2.2115 + 2.4135 \ln \text{dbh}$
$\ln \text{foliage dry weight} = 1.4807 + 2.5993 \ln \text{dbh}$

Melaleuca $0.2 \le \text{dbh} \le 14\text{cm}$

$\ln \text{bole dry weight} = 3.6320 + 1.1874 \ln \text{dbh} \cdot h$
$\ln \text{branch dry weight} = 1.3971 + 1.4838 \ln \text{dbh} \cdot h$
$\ln \text{foliage dry weight} = 2.9067 + 1.8057 \ln \text{dbh}$

Casuarina $0.6 \le \text{dbh} \le 6.9\text{cm}$

$\ln \text{bole dry weight} = 2.5636 + 2.7573 \ln h$
$\ln \text{branch dry weight} = 1.4676 + 1.2203 \ln \text{dbh} \cdot h$
$\ln \text{foliage dry weight} = 2.7646 + 1.5562 \ln h$

NOTES

NON-COMMERCIAL WOODY PLANTS AS POTENTIAL BIOMASS FUEL PRODUCERS: AN ECOLOGICAL RATIONALE FOR THEIR SELECTION

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Harvey L. Ragsdale and James N. Skeen
Emory University
Atlanta, Georgia 30322

ABSTRACT

It appears likely that large-scale biomass energy plantations as presently envisioned (i.e., efforts involving such commercially-important species as Populus, Pinus, Alnus, and Platanus) could eventually prove cost-ineffective owing to the intensity of management required and the corresponding loss of highly productive lands to food crop orientation. A viable alternative to this dilemma is a woody species selection program directed toward successional species, toward exotic and/or rapidly colonizing urban species, and toward species inhabiting extreme or unfavorable habitats. These species have become, over time, well-adapted to a wide variety of marginal soils and habitats. Such species, once selected according to well-defined ecological and physiological criteria and evaluated for growth and biomass fuel species characteristics by comparative field trials at a common site, may then be outplanted to areas of marginal soils not presently under crop production. Such an approach has the obvious advantage of making a broader segment of the landscape usable while at the same time preserving our more productive agricultural lands for food crop production.

We will describe the results of first-year efforts with such an approach in the southeastern U.S. Twenty non-commercial species of particularly high potential and promise have been intensively screened and evaluated. The criteria utilized in selecting these highly promising species (e.g., physiological and environmental requirements, pollutant tolerances, weather/climatic constraints, disease resistance, growth form and propagation characteristics, biomass yields, etc.) will be presented. Data from preliminary germination, propagation, and culturing trials involving the more prominent species will be presented and their implications will be discussed.

INTRODUCTION

Our research is an ecologically based study to identify and screen non-commercial woody plant species for their potential as biomass fuel producers. This work is centered in the Piedmont and Upper Coastal Plain Provinces of the south-[illegible]rn United States, a region with great pro[illegible] for woody biomass fuel production. The project is organized into two distinct phases. Phase 1 addresses the identification and gross screening, through expert opinion and literature review, of the most promising woody species for biomass fuel production on marginal soils of the southeastern United States. Phase 2 is field-screening of those most-promising species in Biomass Fuel Species Gardens at favorable and marginal soil sites.

OBJECTIVES

Our objectives are: 1) to identify non-commercial woody plant species which may be potentially useful biomass fuel producers; 2) to screen these plant species through a goal-oriented literature analysis; and 3) to field-screen, on both favorable and marginal soil sites, the most promising non-commercial woody plants by comparing growth and caloric yield of each promising species to those same measurements for Populus and Platanus, commercially important species with a history of short rotation culture.

APPROACH

Perspective and Rationale

Recent research into intensive plantation culture of woody species and fast rotation growth of woody plants has proceeded, for the most part, with cultivated commercial species such as Populus, Alnus, Pinus, and Platanus [1,2,3,4]. These species have been developed as significant commercial species in an historical context in which the higher costs of energy, resources, and management were not as significant as they are today [5]. The effects of pollutants on commercial species was not a significant factor in the selection of current "commercially" productive species. Further, there was no requirement that these species be productive or nutrient efficient on marginal soils. The approach of using these "commercial" species is, of course, expedient since a sizable literature, on-going research programs, and long term field experience are available for these species [6].

However expedient it may be to consider biomass fuel production from the current array of commercial species, this approach is deficient. It ignores the availability of a large number of naturally occurring non-commercial woody plant species which are adapted to a wide variety of marginal soils and habitats. The identification and screening of potentially productive non-commercial species would negate the deficiency of considering only the currently commercially important species.

Strong justification can be made for the use of an ecologically based selection and screening program. Ecological considerations are not particularly important for selection of maximally productive species under intensive culturing since near optimum conditions are provided under which high production species are identified. However, when the objective is scientifically to identify and to screen woody plant species which have maximum production on marginal soils under non-intensive culture, the use of fundamental ecological knowledge and approaches is indispensable.

The rationale for using an ecological approach is that a wide variety of woody plant species already successfully develop and survive in these marginal soil habitats. Studies of secondary succession have characterized a number of species which are fast-growing, tolerant of wide extremes of environment, resistant to insect pests and environmental pollutants, and highly efficient in absorption and storage of nutrients [7,8,9]. However, a complete list of such potentially useful biomass producing species is not available, and little is known of the growth and productivity characteristics of such species under moderately intensive culture.

Methods

The species screening and selection program was planned as two distinctly separate but interrelated phases. Phase 1 has consisted of species identification and a goal-oriented literature analysis. Potentially useful plant species have been identified from an in-depth review of published studies involving secondary succession in the southeastern U.S., from field trips to naturally wooded marginal lands in the area, and by inquiry of ecologists and foresters who were currently researching natural systems within the Eastern Deciduous Forest. Species identified by these means were then preliminarily screened to produce a list of the most promising species for a goal-oriented literature survey. The results of the literature survey were used to formulate a final list of species whose potential as biomass fuel producers could be assessed by field trials.

Phase 2, the field screening program, will compare the growth and productivity of the selected woody plant species. All selected species will be cultured in a Biomass Fuel Species Garden located in a favorable Piedmont soil habitat. The growth of these selected species on this optimum Piedmont site will be compared and the Biomass Fuel Species Garden will also be used to culture seedlings for field screening on marginal soils. The field trials will proceed by outplanting one-year-old seedlings of selected species to marginal soil habitats in the Piedmont and Upper Coastal Plain Provinces. The plants at marginal sites will be non-intensively cultured for a two-year period. In the Biomass Fuel Garden and at both marginal soil sites, growth of potential biomass fuel species will be compared to two internal controls, Populus deltoides and Platanus occidentalis, commercially important species which have been intensively studied for their biomass fuel potential [10,11, 12,13,14]. The control species will undergo plantings identical to our non-commercial "experimental" species. Expected results of Phase 2 include a comparison of the growth and biomass fuel characteristics of the selected species based on their culture in a Biomass Fuel Species Garden at a favorable Piedmont site and on marginal soils in the Piedmont and Upper Coastal Plain.

KEY RESULTS/ACCOMPLISHMENTS

An initial working list of 126 woody species of possible biomass fuel importance was prepared from our review of secondary succession in the southeast, field trips and our previous experience in marginal site woodlands, and solicited responses from professionals in forestry, ecology, and botany. A working list of 20 non-commercial species (Table 1) of high promise was derived using the collective recommendations of the professional group and our review of the known characteristics (e.g., growth form, rapidity of growth, moisture requirements, etc.) about each of these species. A Species Information Outline (Table 2) was developed to organize pertinent information on environmental and physiological tolerances, ecological aspects, growth form, propagation requirements, bio-chemical characteristics, and biomass fuel aspects of each potential biomass fuel species.

Several appropriate information data bases (e.g., AGRICOLA-National Agricultural Library; BIOSIS PREVIEWS - Biological Abstracts and Bioresearch Index; CAB ABSTRACTS - Commonwealth Agricultural Bureaus) were searched for references on the 20 species of interest. These literature data bases include the major journals in biology, ecology, forestry, soils, and agriculture. Approximately 3000 titles of potential interest were collected from the combined manual (pre-1969) and computerized literature searches. In some instances the SSIE CURRENT RESEARCH (Smithsonian Science Information Exchange) data base was used to determine whether funded research was in progress.

In order to simplify sorting and collation of references, a utility computer program has been developed. Sorting categories correspond to our Species Information Outline so that retrieval [illegible] be effected using any outline categories for [illegible] or more species. Such a scheme facilitates rapid access to available information as well as pointing

TABLE I

WOODY SPECIES OF POSSIBLE
BIOMASS FUEL IMPORTANCE

EXPERIMENTAL SPECIES

Acer negundo - boxelder
Ailanthus altissima - tree of heaven
Albizia julibrissin - mimosa tree
Alnus spp. - alder
Aralia spinosa - Hercules' club
Catalpa spp. - catalpa
Diospyros virginiana - persimmon
Gleditsia triacanthos - honey locust
Liquidambar styraciflua - sweet gum
Paulownia tomentosa - princess tree
Prunus pensylvanica - fire cherry
Quercus laevis - turkey oak
Quercus marilandica - blackjack oak
Rhus copallina - winged sumac
Rhus glabra - smooth sumac
Robinia hispida - bristly locust
Salix nigra - black willow
Sassafras albidum - sassafras
Ulmus parvifolia - chinese elm
Zelkova serrata - zelkova

CONTROLS

Platanus occidentalis - sycamore
Populus deltoides - cottonwood

TABLE 2

SPECIES INFORMATION OUTLINE

- 1.0 Genus-species, Common name, Family
- 2.0 Distribution
  - 2.1 U.S.
  - 2.2 World
- 3.0 Physiological and Environmental Tolerances
  - 3.1 Physiological
    - 3.1.1 Moisture
    - 3.1.2 Nutrient Levels
    - 3.1.3 Soils
    - 3.1.4 pH
    - 3.1.5 Temperature
    - 3.1.6 Pollutants
  - 3.2 Environmental
    - 3.2.1 Snow
    - 3.2.2 Ice
    - 3.2.3 Wind
    - 3.2.4 Fire
    - 3.2.5 Frost
    - 3.2.6 Salt water, spray
    - 3.2.7 Other
- 4.0 Ecological Aspects
  - 4.1 Insect damage
  - 4.2 Fungal damage
  - 4.3 Allelopathic
  - 4.4 Escape from cultivation
- 5.0 Growth and Propagation
  - 5.1 Growth
    - 5.1.1 Seed
    - 5.1.2 Coppice
    - 5.1.3 Suckers
    - 5.1.4 Root System
  - 5.2 Propagation
    - 5.2.1 Sexual
    - 5.2.2 Asexual
- 6.0 Biochemical Characteristics
  - 6.1 Toxins
  - 6.2 Resins
  - 6.3 Other
- 7.0 Biomass Fuel Production Aspects
  - 7.1 Commercial use of species
  - 7.2 Location of commercial stands
  - 7.3 Yield/acre
  - 7.4 Caloric content
  - 7.5 Woody density
  - 7.6 Water retention of logs
  - 7.7 Rotting/Storage characteristics

up a lack of information for specific species an categories.

The collected literature plus standard texts provided the information for completion of the Species Information Outline, Seed Protocols, and Species Summaries. The Species Information Outline (Table 2) provides categorization of research data in a brief summary format sufficient to reveal the presence or absence of information and the major strengths or weaknesses of a species as defined by the outline. The Seed Protocols are informational summaries consisting of seed collection and storage information; pre-germination preparation; germination treatments including controlled environment, greenhouse, and field methods; preventative disease treatment; seedling development; a materials list; and the literature on which the Seed Protocol is based. If sufficient research has been completed and published, then a Seed Protocol is adequate to guide a user from seed collection through development of seedlings. The Species Summary is an integration of the existing written knowledge of these potential woody biomass fuel species organized by the Species Information Outline (Table 2).

Our literature analyses, including current unpublished research projects, and consisting of more than 700 pertinent papers, reveals major research deficiencies in the areas of mineral nutrition efficiency, pollution tolerance/resistance, and caloric yield for the 20 noncommercial woody species under consideration for field trials. It can be assumed that sufficient caloric measurements will be made on commercial species of interest in several other biomass fuel programs. However, there is a paucity of data on nutrient efficiency and pollution tolerance of both commercial and noncommercial woody species. Nutrient efficiencies and pollution tolerance are important and possibly among the overriding characteristics for both species selection and clone development. The lack of research results in these areas means that scientifically and managerially sound decisions on species or clone selection cannot be made. There is so little data available on nutrient efficiencies and pollution tolerance that we strongly recommend every effort be made to initiate such research.

During 1978-79 we initiated limited studies to provide the basis for the culturing of plants in Phase 2 in the second year. Germination procedures were tested using Rhus glabra (smooth sumac). The locally collected seeds were scarified and divided into two groups of 500 seeds each. One group was germinated in sterilized petri plates followed by seedling development in peat pellets. The second group of 500 seeds was planted in sterilized soil:sand (3:1) mixture. Germination was about 60% in both groups but subsequent survival was 30% for the seedlings in peat pellets and only 8% for seedlings developed in soil.

Ailanthus altissima, one of the more promising plant species for marginal habitats, was selected to test propagation of root cuttings. Root cuttings of the lateral and tap roots were cold-stratified for 3 weeks, followed by culture in the greenhouse and growth chamber. The Ailanthus were cultured over a period of 3 months with the greatest survival (60% vs. 9%) in the growth chamber. Under controlled growth conditions there was no difference in tap- or lateral-root cutting survival. Under greenhouse conditions, the tap root survival greatly exceeded that of lateral roots (31% vs. 9%).

Peat and soil growth mixtures were compared using Acer negundo (boxelder). The media comparisons were initiated with 6-cm seedlings which were cultured in the greenhouse and in a heated greenhouse germination bed. Over a two month period, boxelder cultured on soil both in the greenhouse and in the germination bed had superior height growth, 68 and 31 cm respectively, as compared to boxelder cultured on peat, which exhibited about 9 cm height growth.

Intensive plant culturing in our greenhouse resulted in a continuing infection by white flies. Biological controls were substituted for the chemical insecticides which were in use. These controls consisted of "traps" for the emerging flies and use of an "insecticide" blended from garlic, onion, and peppers. The "insecticide" solution was developed by filtering the preceding ingredients, adding detergent, and diluting. The traps were yellow-painted boards with a heavy oil and vaseline coating. Adult flies were trapped on the boards and a noticeable decline in adult white fly population was observed. The "insecticide" was applied routinely and eventually eliminated the white fly population by killing the emerging adults. These controls have been continued with no re-infestation by white flies.

FUTURE PLANS

The reemergence of the fuel crisis, the current problems surrounding the increased use of nuclear power, and the energy shortfalls projected for the next two decades all point to the need to develop alternative fuel and energy sources. Biomass fuels are an attractive solar energy source since they are renewable and at the present time, intensively managed woody biomass fuel farms are favored as a source of continuing woody biomass. As suggested earlier in this paper, the biomass fuel farm concept requires enormous energetic inputs (fuel, fertilizer, herbicides, water) to intensively culture the most popular candidates, such as Populus and Platanus, in a biomass fuel farm context. Given that only agriculturally marginal land may be available in the near future for tree culture, it seems only reasonable to evaluate those woody plant species which are adapted to these existing "marginal" soils. These species are optimally selected having survived both germination and seedling development and maturation to a sexually reproducing stage. We believe that it is ecologically sound to consider as woody biomass fuel species those plants which already are adapted to these marginal habitats. The important questions to be answered regarding these species are to what extent these species respond to different levels of fertilization and moisture.

In a recent article stressing the interrelatedness of eco-physiology, genetic suitability, and nutrient utilization and efficiencies [15], it has been suggested that biomass production may be limited more by stress factors than by inefficiency of the plant's photosynthetic apparatus. Mineral deficiencies and/or toxicities, temperature extremes, moisture limitations and salinity are noted as possible stress factors. Others [16] also have offered evidence which suggests that plants differ markedly in their capacity to extract and absorb nutrients from relatively unavailable sources as well as in the efficiencies with which the nutrients are utilized metabolically. In some instances differences encountered within a species are as great or greater than differences observed between different species. These findings coupled with our proposal and first year results offer a strong argument for basing a species selection program around species already adapted to marginal site requirements rather than emphasizing species with an expectation of success only when grown under optimal soil and moisture regimes. Further, credence is given species selection based on genotype breadth rather than a selection based on specific varieties or hybrids which, while they may exhibit excellent germination and establishment characteristics, may later in their growth and development encounter pronounced deficiencies, toxicities, or pest/disease susceptibilities. By selecting from species already attuned to a wide variety of marginal habitats, and making the propagule selections within each respective species from a variety of seed sources, the problem of pre-adaptation to narrow and specific site requirements can be largely controlled. By additionally comparing, at both optimum and marginal sites, the significant ecological and biomass-yield parameters of marginal site-adapted experimental species against internal controls (Populus and Platanus), both a performance ranking for species by site-type and a comparison of species performance within sites can be effected. Consequently, the expected result of a scientifically and ecologically sound biomass fuel species screening and selection program will have been achieved.

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SPECIES SELECTION AND SILVICULTURAL SYSTEMS FOR PRODUCING FUELS FROM WOODY BIOMASS IN THE SOUTHEASTERN UNITED STATES

Department of Energy Contract Number: ET-78-G-01-3038

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ouglas J. Frederick, Associate Professor and
roject Leader, School of Forest Resources, North
arolina State University, Raleigh, NC

illiam E. Gardner, Liaison Silviculturist, School
f Forest Resources, North Carolina State Univer-
ity, Raleigh, NC

imothy K. Slichter, Graduate Assistant, School
f Forest Resources, North Carolina State Univer-
ity, Raleigh, NC

Robert C. Kellison, Director, Hardwood Research
Cooperative, School of Forest Resources, North
Carolina State University, Raleigh, NC

Michael Williford, Research Associate, School of
Forest Resources, North Carolina State University,
Raleigh, NC

BSTRACT

he Hardwood Research Cooperative in the School of orest Resources at North Carolina State Univerity is screening and evaluating woody species, nd testing spacing and various silvicultural echniques for optimizing energy production. Analsis of 43 existing species-site tests installed y the Cooperative show loblolly pine to out prouce most hardwoods. Several hardwoods can produce igh yields but intensive cultural practices are equired and planting must be done on specific soil ypes.

ine new species-site tests have been established uring 1978-1979. A bottomland and an upland site sing 8 species was planted in North Carolina to valuate 3 spacings, fertilization and cultivation n growth, biomass and energy yields. Both of nese sites are considered marginal for agriculure, the bottomland site because of susceptibility o flooding and the upland site because of nutrient eficiencies and past exploitation and erosion. ney are representative of large acreages in the outheast, and could represent the first sites vailable for energy plantations.

NTRODUCTION

ne use of wood as an energy source for forest-ased industries and residential homes continues o accelerate nationally. However, greatest nterest in the Southeast for use of wood as a ajor energy source is shown by the small and edium sized non-forest based industries. Several rick and textile plants in the region have conerted to wood, and Pullman Woodex, a wood pelletzing company, is building a $1.5 million plant ear Goldston, North Carolina to supply wood fuel or industrial use. Other industries including obacco processors are considering using gasified [illegible]o replace fuel oil for leaf curing. These [illegible] will intensify as the price of petroleum fuels rise. Estimates for North Carolina show that twenty percent of the State's energy needs could be obtained from wood with no detrimental effect on supplies to the wood-using industries. Other heavily forested states in the Southeast have similar potential.

The producer and consumer system is not well developed in the region and wood shortages are being projected particularly along the Atlantic and Gulf Coastal Plain where most of the wood-using industries are concentrated and competition is most intensive. These projected shortages can be alleviated to a degree by increasing utilization in the forest and during processing, increasing yields through intensive management and genetic improvement and by rising prices for the resource [13].

Short rotation energy plantations are likely to become an important long term source of woody biomass in the Southeast. Some industrial members of the North Carolina State University Hardwood Research Cooperative are considering managing existing plantations and establishing new plantations for energy purposes. For example, Olinkraft Corporation in Louisiana has considered establishing Eucalyptus energy plantations to safeguard against precarious natural gas supplies for its West Monroe pulp mill.

These trends make it imperative that we learn more about the inventory, availability, utilization, and management of the resource in plantations and natural stands.

TASK DESCRIPTION, OBJECTIVES AND APPROACH

This paper describes the research project funded by the Department of Energy, National Biomass Program, that the North Carolina State Hardwood Research Cooperative has initiated to investigate the best species and optimum silvicultural systems to use in short rotation energy plantations in the South. The project began June 1, 1978, and consists of two major parts: Division I - proper species and source determination and, Division II - optimum silvicultural systems and quantification of energy yields. The two divisions have been identified to most efficiently utilize currently available research information and the resources of the Hardwood Research Cooperative.

The N. C. State Hardwood Cooperative is supported by 20 industrial forestry organizations and the North Carolina Forest Service. These organizations own or control about 15 million acres (6 million ha) of land from Delaware to Louisiana. A wide variety of research projects, encompassing applied and basic management objectives, are coordinated by the staff at N. C. State. Studies are installed on Cooperator lands using their labor and at their expense. Data collected from the trials by Cooperators are analyzed and interpreted by N. C. State, after which the information is made available to the members so they can take advantage of accomplishments. Information obtained by the Cooperative is not considered priority; it is disseminated in oral and narrative forms as quickly as available.

Benefits from this cooperative arrangement for externally funded projects are substantial. For our Department of Energy project, Cooperators are providing: planting sites, equipment, and labor for site preparation, and some aid in plantation aftercare and stand maintenance. A considerable amount of supporting equipment and facilities are also being made available through N. C. State University. These services are exclusive of the project budget.

Division I - Proper Species and Source Determination

Species selection and screening takes high priority in any attempt to evaluate the energy plantation concept. Such programs must encompass known fast growing commercial tree species, lesser-known "noncommercial" species as well as exotics. Desirable fuel conversion characteristics, growth and genetic improvement potential, and other characteristics must be evaluated before long-term committments are made in the final selection of species. Both hardwoods and conifers have characteristics conducive to energy plantation culture and it is likely both will be used under certain conditions. A substantial amount of selection and screening work has been done with southern tree species by the Cooperative and although this information is very useful and supportive, the original intent was not for evaluating species energy potential.

The first of four objectives in our project was to evaluate nearly 80 species-site plantings that had been established by Cooperative members throughout a 12-state region of the South. These plantings were established to elucidate the relationships between the performance of numerous commercial species and site, and identify species best su to plantation management. These original objectives are still compatable with the current project objectives. Some of the plantings were abandoned because of lack of maintenance, poor survival, or growth. However, 43 installations ranging up to 18 years old and involving 3 pine and 22 hardwood species remain intact (Table 1). General conclusions based on these studies have been summarized [9], and include: (1) loblolly pine (Pinus taeda) will grow faster and produce more biomass at a younger age than most hardwood species on all sites where it can be successfully established, (2) intensive silvicultural measures including: site preparation, fertilization, and cultivation are necessary to successfully grow hardwoods, and (3) sycamore (Platanus occidentalis), sweetgum (Liquidambar styraciflua), water oak (Quercus nigra), willow oak (Q. phellos), and green ash (Fraxinus pennsylvanica) show greatest promise for hardwood plantation establishment and management. Hardwoods are very site sensitive and must be matched to preferred soil types for acceptable production [2]. Sweetgum is the most adaptable species of the commercially important southern hardwoods. Although most of the commercially important species have been planted in the original species-site tests, many were not well represented on all sites. Furthermore, numerous noncommercial species and exotics had not been tested.

Table 1. HARDWOOD SPECIES-SITE STUDIES BY FOREST SITE TYPE

Forest Site Type	Coastal Plain	Piedmont	Mountain	Total
Red River Bottom	11	1	1	13
Black River Bottom	1			1
Branch Bottom	2			2
Wet Flat	6			6
Peat Swamp	1			1
Bottomland	5	4	1	10
Coves and Gulfs			2	2
Upper Slopes and Ridges	1	1	1	3
Prairie		1		1
Upland	4			4
Total	31	7	5	43

Our second objective is to screen and identify all species specifically having potential for inte[illegible]sively managed short rotation biomass producti[illegible] in South. Nine species have been planted in ne ials at 5 locations during 1978-79. Concurrently, seed has been collected for planting boxelder (*Acer negundo*), black locust (*Robinia pseudoacacia*), tree-of-heaven (*Ailanthus altissima*), red maple (*A. rubrum*), *Ligustrum* spp., European black alder (*Alnus glutinosa*), cottonwood (*Populus deltoids*), Virginia pine (*Pinus virginiana*), and *Eucalyptus* spp.

All species and sources selected for species-site tests are planted at several locations using at least 4 replications of 25 tree plots at 4 x 8 foot (1.2 x 2.4 m) spacing. The interior 3 x 3 tree block (9 trees) are evaluated and measured at intervals for survival, height and basal diameter with the border 16 trees serving as a buffer. Minimum site treatment includes: site clearing, cultivation and fertilization to correct deficiencies.

In addition to the above measured parameters, our new installations will be measured for biomass and energy yields, and nutrient content of whole trees and components. Coppice production will also be measured from sample replications harvested at age 5.

Even though most species adaptability testing has been with indigenous species, increased emphasis will be placed on exotics. We have found certain *Eucalyptus* spp. to be freeze hardy and show very good yield potential particularly on upland sites south of latitude 32° N [4].

Eucalyptus spp. are particularly promising for intensive plantation culture in the South because of their fast growth, desirable wood properties, energy potential, broad site adaptability, and potential for genetic improvement [10]. There are over 550 species of eucalypts worldwide which includes some of the fastest growing tree species known. The largest scale energy plantation planting program in the world, over 200,000 acres (80,972 ha) per year, is with *Eucalyptus* in South America [16]. Wood from these plantations is converted to charcoal for use in steel mill blast furnaces.

Since the beginning of our *Eucalyptus* evaluation trials in 1971, thirteen Cooperators have screened 475 lots representing 97 species for survival, growth, cold hardiness, pest resistance and suitable wood properties. We are continuing this screening and, evaluating new plantings for energy potential. During 1979, approximately 40 species will be planted.

Species showing greatest promise for the zone between 32° N and 30° N latitudes where freezing temperatures are frequently encountered include *E. viminalis*, *E. nova-anglica*, *E. macarthurii* and *E. camphora* [5]. South of latitude 30° N, where freezing temperatures are less common, species with good potential include *E. grandis*, *E. robusta*, *E. camaldulensis* and *E. teriticornus* [1].

[illegible]ass yields of the best species and sources equal or surpass those of native hardwoods on comparable soils [4]. Annual height growth rates of 10 feet (3 m) are being observed for the best adapted sources on the better sites.

European black alder is also a promising species for energy, as well as timber production, and soil stabilization on spoil banks. The species' capacity for nitrogen fixation and its adaptability to upland sites also gives it advantages over most indigenous hardwoods south of latitude 38° N [7].

We have evaluated numerous sources of black alder and find it often performs better than indigenous species particularly on fertile soils. On upland sites, heavy fruiting at early ages accompanied by growth suppression often occurs which we feel is a result of using sources from northern locations. Over the next few years, selected seed from over 100 alder sources from south of 46° N latitude in the indigenous range of the species will be planted for evaluation. Seed from these sources is being provided by Dr. Richard Hall, Iowa State University, who has coordinated a range-wide seed collection of the species.

Genetics Research

Our third objective is to employ present and refined genetics breeding programs to locate and/or develop suitable trees to be grown for fuel. Seed source studies of numerous indigenous southern hardwoods from across the Southeast have shown that local sources are best and considerable variation exists among trees from a given region. This variation is enough to allow for good genetic gains from a selection program [6, 14]. Additionally, large variation has been found in sweetgum and sycamore [11, 12]. This phenomenon offers opportunity to make additional genetic gains by selecting the best trees from the best stands.

Since 1960, 572 selections representing 25 species have been made by the Cooperative. Criteria for selection have emphasized growth rate, form and wood quality. Trees have been propagated in seed orchards and clone banks for use in breeding programs. Seed orchards of sweetgum, sycamore, water-willow oak, green ash, Eucalyptus and yellow poplar (*Liriodendron tulipifera*) have been established (Table 2).

Table 2. HARDWOOD SEED ORCHARDS BY SPECIES AND ACREAGE IN THE HARDWOOD RESEARCH PROGRAM

	No. Orchards by Type		Acreage of Orchard by Type	
Species	Clonal	Seedling	Clonal	Seedling
Eucalyptus	1	2	1	2
Green ash	1	-	2	-
Sweetgum	6	1	37	4
Sycamore	8	2	18	6
Water-willow oak	5	1	15	3
Yellow-poplar	1	-	6	-

We continue to encourage activity in tree selection because of the continuing loss of the best phenotypes to harvesting, land clearing and natural causes. With the additional emphasis on using plantation forests for energy and chemicals, some additions to the tree selection criteria are needed. For example, wood alpha cellulose content [3] and caloric or BTU values are useful measures of the chemical and energy potentials of wood and have not been measured in earlier selections. However, traditional growth criteria and the wood properties, specific gravity and moisture content are still reliable measures of wood energy and chemical potential [15].

Southern hardwoods rarely occur in stands of single species. This phenomenon makes the comparison-tree system of phenotype selection (comparing a candidate tree with others in the immediate vicinity) highly inefficient [9]. Consequently, the mother-tree or family-comparison approach has been used. In this approach, open pollinated seeds are collected from desirable phenotypic selections for inclusion in family comparison tests. Based on half sib progeny performance, clonal first generation orchards can be established using the best genotypes. Orchards of seedling or clonal origin can also be established from the best trees in the best families of the progeny tests. We currently recommend second generation orchards for obtaining greatest genetic gains in southern hardwoods.

Considerable variation occurs in tree performance within hardwood plantations due to both site variation and heterogeneity of plant material. To capitalize on that variation, selections are being made from progeny and provenance tests of known genetic identity for inclusion in seed orchards.

We feel large genetic gains can be made by selecting superior phenotypes from the large existing hardwood plantation acreage in the South. What kind of gains can we expect in a selection program with hardwoods? Will they be comparable with southern pine? Gains in volume production from use of genetically improved hardwood seed will exceed the commercial check by twenty-five percent [17]. Significant gains in wood quality including those parameters influencing energy and chemical potential will also be realized. Perhaps the greatest gains will be in adaptability. Variation in tree growth and volume production in nearly all hardwood plantations is highly variable, regardless of apparent soil uniformity. By selecting the best performing individuals for establishment of family comparison tests and clonal seed orchards, yields at 10 years for all species being planted could be doubled by using trees with superior performance [8].

Division II - Optimum Silvicultural Systems and Quantification of Energy Yields

Our fourth and final objective is to test the effects of site preparation, spacing, cultivation, and fertilization on survival, growth, and yield of species identified for potential use in short-rotation energy plantations. Two sites, a bottomland and an upland, have been planted in North Carolina using the most promising species identified in species-site tests and genetic selection evaluations. Prior to planting, both sites were intensively site prepared to agricultural standards to assure plantation survival and growth using tractors and heavy disks. Soils were analyzed for nutrient status and will be fertilized during the first 6 months after establishment. Four replications of three spacings were used on each site: 2.5 x 5 feet (.75 x 1.5 m), 5 x 5 feet (1.5 x 1.5 m), and 8 x 5 feet (2.5 x 1.5 m). After planting, sites are being cultivated as needed using a small tractor which can operate between the 5 foot (1.5 m) row width. Spacing, species, and site are the major treatment effects. The bottomland site on lands of Georgia Pacific Corporation in Bertie County was planted to seven species: sweetgum, sycamore, green ash, water-willow oak, European black alder, cottonwood, and loblolly pine while the upland site on lands of Champion International Corporation in Granville County was hand planted to four species: sweetgum, black locust, European black alder, and loblolly pine.

The experimental planting design for each species allows for four replications of 49 tree blocks to be sampled at ages two, four, eight, and twelve years after planting. Sampling will consist of the interior 25 trees being measured for survival, total height, and basal diameter. Trees will be cut at ground level for determination of green and dry weight of total trees and components. Subsamples will be processed for specific gravity, moisture content, nutrient content, and energy yields. Nutrient depletion under short-rotations may become an important factor in this type of intensive culture making nutrient monitoring and analysis a very important aspect of this study. Extent of coppicing will also be measured at various intervals on all harvested blocks to elucidate biomass yields.

At the termination of the project, all coppice plots will be harvested and sprout yields determined. These will then be correlated with annual heights and basal diameter measurements made during intervening years. At this time, specific gravity, moisture content, nutrient contents, and energy yields will be determined for coppice subsamples.

Wet conditions delayed planting of these sites longer than expected. Site preparation on the upland site was not completed until April, 1979, due to wet conditions so the planting was not done until early May. We had planned to plant the bottomland site during December, 1978, however, the site was flooded earlier than expected and planting did not occur until late May, 1979. Both of these sites are considered marginal for agriculture, the bottomland because of susceptibility to flooding and the upland because of nutrient deficiencies and past exploitation and erosion. They are representative of large acreages in the Southeast, and could represent the first sites available if the energy plantation concept beco[illegible] a reality.

RESU[illegible]

A great deal of data has been evaluated on species-site adaptability in the South and we feel enough is known to make definite recommendations for several species. Loblolly pine performs better than most hardwoods on all but permanent river bottom sites. Improved, mechanized planting procedures and genetically superior seedlings giving high yields may give pine added advantages. However, under short rotations of repeated crops, loblolly has serious limitations because of its lack of sprouting ability. Based on past and ongoing planting trials with indigenous species, sycamore, sweetgum, water oak, willow oak, and green ash show greatest promise for energy plantation management. Each must be matched to specific sites and cultural practices must be intensive for good survival and growth. Several additional species including boxelder, tree-of-heaven, red maple, *Ligustrum* spp., black locust, cottonwood, and Virginia pine may be suitable, however, with the exception of cottonwood, these have not been extensively tested under close spacing, short-rotation culture.

The exotics, *Eucalyptus* spp. and European black alder have great potential in the South for energy plantations because of rapid growth, desirable wood properties, and site adaptability. Our species-site trials have indicated this trend and our comprehensive genetic source screening trials are further substantiating it.

Selection trials for the indigenous species have also demonstrated considerable variation among and within families. There is little doubt that desirable energy characteristics can be selected for and considerable genetic gains attained.

With the completion of the Division I and II plantings, this project is on schedule. First year evaluations and measurements will begin during 1980. Already a considerable amount of experience has been gained on the logistics of energy plantation culture and management.

FUTURE PLANS

This project is planned under D.O.E. funding for 7 years. New species-site plantings will be installed in connection with this project for at least 5 years. Beyond 5 years, we will continue to test species throughout the South under the auspicious of the Hardwood Research Cooperative using the same experimental design. Results from these trials will be made available to D.O.E.

Division II plantings will be made during years 2 and 3 making a total of 6 installations. Plans for the second year are for plantings to be on the Coastal Plain of Georgia. Third year plantings will be on the Gulf Coastal Plain. Beyond the D.O.E. funding period, the Hardwood Research Cooperative will continue data collection and [illegible]s as long as this effort is judged useful.

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Session V B

RefCOM (REFUSE CONVERSION TO METHANE)

- A PROJECT REVIEW

Peter J. Ware
Director, Engineering and Technical Development
Waste Management, Inc.
Oak Brook, Illinois

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SUMMARY

In 1975, the U. S. Department of Energy (formerly ERDA) awarded Waste Management, Inc. a contract to design and construct a "proof-of-concept" experimental facility to demonstrate the biological gasification of urban solid waste and sewage sludge to produce methane-rich gas. Named RefCOM, the facility will process between 50 and 100 tons per day of the organic fraction of shredded urban refuse. Jacobs Constructors, Inc., a subsidiary of Jacobs Engineering Company of Pasadena, California, was the engineering/construction management subcontractor for the project.

Construction of the facility was completed by May 1st, 1978, and was followed by a six-month start-up phase. The plant has been in daily operation since November, 1978, and is being run in accordance with a predetermined experimental program which will be of two-to-four years' duration. Waste Management, Inc. has been awarded an operating contract to manage the program and operate the facility.

The Department of Energy (DOE), a successor agency of ERDA, has other programs underway which support the experimental program at Pompano Beach. These include mixing experiments comparing gas and mechanical agitation, methods for recovering energy from the resultant sludge material, and feedstock preparation by means of high pressure steam explosion. As information is developed in these ancillary projects, the experimental program at Pompano Beach may be modified.

Project Background

Recent forecasts of steadily declining supplies of clean fossil fuels have focused attention on the need for development of renewable energy resources. Bioconversion to fuels; the conversion of biomass (i.e. the organic fraction of urban solid wastes, agricultural residues, and terrestrial and marine energy crops) to synthetic fuels, is such a resource.

Bench-scale studies and analyses by Dr. John T. Pfeffer, of the University of Illinois, Urbana campus, sponsored by the Environmental Protection Agency and the National Science Foundation, under Research Applied to National Needs (RANN) program, indicated that methane gas produced by anaerobic fermentation of the organic fraction of refuse could be competitive with alternate sources of natural gas. Dr. Pfeffer conducted a series of experiments primarily using refuse-derived fuel (RDF) from St. Louis. Those experiments indicated that 4.5 to 5.5 cu.f. of gas could be produced per pound of volatile solids added at mesophilic temperatures (100°F) and with 10-to 30-day sludge retention times. Pfeffer's positive findings led NSF to fund two additional studies: The first, by the MITRE Corporation, was a program assessment leading to a recommendation that the next scale of facility should be a 50- to 100-ton-per-day plant. The second was an economic and sensitivity analysis of the process by the Dynatech Corporation who developed a preliminary design and an economic model for a full-scale (1000 tons-per-day) digestion plant.

Their economic computations indicated that urban waste methanation would prove technically feasible at a cost competitive with available sources of natural gas. Most of the equipment necessary for operational scale-up of the laboratory benchtop experiments upon which these preliminary designs and cost estimates were based appeared to be adaptable from other industrial processes, and the investigation had reached the point where pricing of construction and operation of a practical size scale-up was necessary to determine actual production, purification and distribution prices.

Based on the favorable findings of these two reports, a request for competitive proposals to construct a 50- to 100-ton-per-day proof-of-concept plant was issued in March, 1975 by the Energy Research and Development Agency, who had by then assumed the NSF program. Twenty bids were received. The contract was awarded to Waste Management, Inc. (WMI) of Oak Brook, Illinois in June 1975.

Proof-of-Concept

The generation of pipeline quality gas, if economically viable, would have certain advantages over low-BTU gas or solid fuel products. These advantages include an established market, established transmission and distribution facilities, available combustion technology and an environmentally preferred fuel. As a precursor to commercialization of urban waste methanation, however, a proof-of-concept experiment is needed. This is usually conducted at a "pilot" or "semi-commercial" facility.

The U. S. Department of Energy funded a "proof-of concept" plant of 100 TPD scale because several process and engineering factors have to be studied before large-scale commercial exploitation of the system can be undertaken. The largest studies to date have been in 100 gallon fermenters, where the daily refuse feed varied from five to ten pounds.

The specific goals of the solid waste to methane proof-of-concept project are:

1. To establish information concerning the gas product quantities and values
2. To evaluate process reliability and economics
3. To determine optimum design and operation parameter values for each process stage and method of operation
4. To establish a basis for comparing the process to other means of energy production and/or resource recovery from urban waste
5. To establish the technological and economic basis for commercial utilization of the process

Waste Management, Inc. Involvement

Waste Management, Inc. is primarily a service-oriented company, engaged in the collection, haulage and disposal of waste materials. Clearly a technology which reduces the volume of waste to be disposed of, and at the same time yeilds valuable products would be of interest to WMI.

The incentive for WMI to commercialize the technology will surface if the prices commanded for product gas escalate at a higher rate than gas production costs; it is expected that this will happen, but the exact data is uncertain, and determining gas production costs is the whole object of the project's experimental effort.

Site Location

The project is located at an existing facility of Waste Management, Inc. known as the Solid Waste Reduction Center, located in Pompano Beach, Florida. Pompano Beach is a community of about 40,000 inhabitants, situated 10 miles north of Ft. Lauderdale, in Broward County, Florida.

The Solid Waste Reduction Center, its support services, and adjacent shredded landfill were dedicated in September, 1971. When it went into service, it was the first privately owned and operated facility of its type in the nation. The existing facility consists of a receiving area, including a tipping floor and two receiving pit conveyors, each feeding a vertical-shaft hammermill. One of these units is rated at 15 tons-per-hour, while the other, installed in late 1978, processes 65 tons-per-hour. The combined discharge of these two machines is fed to load-out facilit from where it is transferred to landfill via high capacity, live bottom trailers. Other existing facilities located at the Pompano Beach site include district refuse collection offices, collection equipment maintenance and storage facilities, and a container maintenance shop.

The Microbiological Process

Since the dawn of time, anaerobic (living in the absence of oxygen) and faculative (living either in the presence or absence of oxygen) bacteria have decomposed organic material into simpler compounds. The principal products of this process have been methane and carbon dioxide. This gas appears naturally as swamp gas in marshy areas, and is the normal product of similar action occurring in all sanitary landfills where municipal solid waste and/or sewage sludge is buried.

The process, essentially in three stages, is complex and not totally understood. In the first stage, cellulose and other complex organic compounds are broken down to simple sugars and soluble monomers. In the second stage, a group of bacteria convert the simple sugars to organic acids, such as acetic and propionic group of bacteria to methane and carbon dioxide.

Design and Construction

Since the in-house capability of Waste Management, Inc. at that time lay primarily in the area of waste collection, transportation, and disposal by landfill, Jacobs Constructors, Inc. was retained to execute the detailed mechanical design under WMI supervision, and to manage the construction phase.

Design and construction of the facility was completed in April, 1978 at a final cost of $3.6 million. In addition to normal construction changes and delays associated with any project, complications of this effort included uncertainties in design specifications, changes in subcontract bid packages, and complexities of federal procurement regulations. Problems with engineering management and construction supervision contributed to cost escalation.

Project Schedule Milestones

Award of Proof-of-Concept Contract	6-23-75
Start Construction	2-2-77
Complete Construction	4-31-78
Complete Start-up (Start Experimental Program)	11-1-78
Complete Program (Current 25 Month Schedule)	5-31-80

Project Budgets

A - Design and Construction	
Engineering (Jacobs)	$1,150,000
Equipment	923,000
Construction	1,318,000
Project Management	144,000
Contingency	115,[illegible]
Sub-Total	$3,651,000

– Operations (Proposed Estimate)	
t-up	$ 396,000
Year Balance (1978)	324,000
2nd Year (1979, and to May 1980)	1,476,000
Sub-Total	$2,196,000
Project Total	$5,847,000

CHNOLOGY

ocess Description

iefly, the process consists of the shredding of .e urban waste, composed at this plant predominant- of residential waste collected in the local oward County area, delivered by standard packer ucks to a tipping floor at a rate of over 1000 ns-per-day. From the tipping floor, the waste ocessed through the Heil vertical shredders to nominal 3-in. size and ferrous metal is removed a Dings magnet. This portion of the facility is ned and operated by Waste Management, Inc., with- .t cost to the proof-of-concept RefCOM plant. The .redded, ferrous-free waste is then split, with .e bulk transmitted to the Waste-Management- .gineered sanitary landfill and, depending on the .periment performed, 35 - 100 tons/day are placed a storage building for further use in the RefCOM ocess. The storage building with a maximum pacity of 400 tons of nominal 3-in. size waste rves as a surge system to match the 7 day/week .fCOM operation to the 5½ day/week primary .redding operation. An articulating front-end ader is utilized to transfer the shredded terial from the surge pile onto a 10 tons/hr. pacity pan conveyor for metering to the process- .g line.

om the storage building, the waste material is .d to a classification building for pre-processing .d then to two digesters for biological gasifica- on. In the classification process, the shredded ste passes through a trommel screen to remove organic grit and fines, such as glass, silica nds, ash, etc. The material then undergoes condary shredding to achieve a smaller particle .ze, and air classification, to separate the light ganic fraction from the heavy, generally inorgan- fraction. The light material, which in other source recovery systems is sometimes referred to Refuse Derived Fuel (RDF) is often used as is. this system it is introduced into a pre-mix tank .ere it is blended with sewage sludge, recycled .ltrate water, nutrients and steam for temperature ntrol. This slurry is then metered into two chanically agitated anaerobic digesters in which .e biological (bacterial) process converts proximately half of the organic feed solids into product gas composed of approximately 50 percent thane and 50 percent carbon dioxide. In full- ale commercial operations, this gas can then be graded to pipeline quality, using demonstrated s clean-up or purification techniques. The gas currently being flared to atmosphere for dis- sal, but WMI has plans to install a pathological [illegible]tal) waste incineration facility to service the Broward County area, which will utilize approximately 30% of the maximum gas production as auxiliary fuel.

The anticipated optimum digester process conditions will be 5 days detention and 140°F. The digester residue is dewatered, and the water so separated is recycled back to the digesters as make-up for the feed slurry. Feed solids concentrations on the order of 10 per cent are contemplated. The dewatered residue--occupying only 30 percent of the volume of raw refuse--is relatively stable, and can be deposited in a landfill, or burned to a sterile ash with the process heat recovered for use in the operation of the digesters. Expected fuel consumption for these conditions is of the order of 15% of total gas production, so further energy recovery of this type is highly desirable.

Laboratory-scale studies performed to date by Dr. John Pfeffer of the University of Illinois, who is also technical consultant on this project, indicate that approximately 6,000 cubic feet of mixed methane and carbon dioxide gas are produced per input ton of raw refuse; thus, 3,000 cubic feet of methane (equivalent to pipeline quality gas) per input ton could be produced by this process. Based on an average home gas consumption of 100,000 cubic feet per year, a 1,000 ton-per-day plant could serve the gas needs of over 10,000 homes.

Plant Capacity (See attached flow sheet - Exhibit 1)

The facility is designed to operate within the range of 50-100 tons-per-day.

Most of the project experiments center on the digester performance. Since they are designed to run with a fixed volume, variation of process conditions, such as residence time and solids loading will vary the actual throughput within this range.

Experimental Program

A detailed experimental program has been developed for the facility to evaluate, in phases, several independent variables. The most significant areas to be studied are: (1) methane production per unit of organic solids fed (2) reduction in solids remaining for ultimate disposal (3) mixing characteristics of the urban waste/sewage sludge slurry (4) dewatering characteristics of the reactor slurry (5) process stability (6) energy requirements for operation, and (7) chemical cost for nutrients and pH control.

The various phases of experimentation will include determination of optimum fermentation temperatures (mesophilic and thermophilic), evaluation of feed preparation, evaluation of feed solid concentration and residue recycle, evaluation of pH and nutrient requirements, residue dewatering, and residue disposal.

The schedule of these activities is attached as Exhibit 2.

OPERATIONS

Start-up

The initial phase was necessary to familiarize the operating personnel with the various subsystems, to check out equipment under load, and to initiate the anaerobic digesters. This included familiarization with the controls, pumps, and valves, with the biochemical requirements such as pH, temperature, and nutrients, and finally with the proposed experimental program. During this period, the digesters were filled and an anaerobic fermentation established in a suitable substrate. This allowed final testing of the performance of all equipment and also allowed the operators to practice the variations of operational controls prior to experimentation.

This activity also served to identify many of the shortcomings of the engineering design; although 'state-of-the-art' information was incorporated, the initial operation of the plant provided new information relating to the handling and processing of shredded refuse which has been reflected in extensive plant modifications throughout the system. Such characteristics as bulk density (particularly during conveying), and particle size-distribution (particularly of textile and plastic materials) have resulted in redesign of the conveying, agitation and feed systems. Environmental controls, particularly of dust generated in the secondary shredder, was another problem that was resolved during the start-up phase.

By late September 1978, most of the obvious restrictions had been identified and resolved, and the plant was deemed capable of daily operation. It has been fed with refuse and has generated product gas on a daily basis ever since.

The Experimental Phases

Mesophilic Digestion Phase

Currently, a series of experiments are being conducted at 100°F temperature to evaluate constant conditions of slurry feed concentration and solids loading. The variable to be adjusted is the retention time, and sufficient data points are needed to develop accurate relationships. The experimental matrix is given in Table 1 and indicates a total of four experimental conditions, conducted two at a time in the adjacent digesters.

Although it is expected that optimum performance will occur at thermophilic temperature conditions, it is important, nonetheless, to establish the baseline condition at the mesophilic range. These experiments will serve as a check on the range of operating conditions that were established (pH, alkalinity, acidity, nutrients), illustrate the effect of retention time on gas production and show net benefits as a function of retention time, and provide an opportunity to further shake down the total system analytical procedures. Special attention will be given to the degree and efficiency of mixing.

At the 100°F mesophilic temperature, the data Dr. Pfeffer's experiments indicate that gas production ranges from approximately 4.5 stnadard cu.ft./lb. of volatile solids added for 10 days retention time to approximately 5.5 standard cu.ft./lb. volatile solids added for 30 days retention time. The primary purpose of this experiment is to determine gas production within these retention times at the conditions shown in Table 1. This experiment, essentially, will establish the degree to which the results obtained by Pfeffer on a laboratory scale can be reproduced on a large scale.

Thermophilic Digestion Phase

A series of experiments will be conducted at the thermophilic temperature range of 130°F to 140°F at various retention times. Constant conditions of size, slurry feed concentration, and recycle will be maintained.

Table 1. Mesophilic Experiments (100°F)

	1	1A	2	2A
Slurry feed concentration %	10	10	10	10
Temperature °F	100	100	100	100
Recycle rate %	50	50	50	50
Retention time/days	30	20	15	10
Particle size/inches	1.5	1.5	1.5	1.5

Thermophilic Experiments (135°F)

	3	3A	4	4A	5	6
Particle size/inches	1.5	1.5	1.5	1.5	1.5	1.5
Slurry feed condentration%	10	10	10	10	10	10
Retention time/days	20	15	10	5	5	5
Temperature °F	135	135	135	135	135	135
Recycle %	50	50	50	50	50	50

Other Experiments

Having evaluated the performance of the system at thermophilic levels, under constant conditions, various changes in other process variables will be effected to evaluate the system response.

Experiments 7 and 8 will consider feed slurry concentration at higher levels:

Table 2. Feed Slurry Concentration Experiment

	Experiments	
	7	8
Particle size/inches	1.5	1.5
Retention time/days	5	5
Temperature	140	140
Recycle %	50	50
Feed slurry concentrate %	15	20

Remaining experiments will address the impact of varying recycle rations of the final filtrate liquor into the process:

Table 3. Recycle Ratios

	9	10	11
Size/inches	1.5	1.5	1.5
Slurry concentration	15	15	15
Retention time/days	5	5	5
Temperature	140	140	14[illegible]
Recycle %*	50	60	Max[illegible]

*% of make-up water from recycle liquor.

A concurrent series of experiments will be carried out throughout the experimental phase to explore the effect of particle size on the process. At the optimum conditions established, the particle size will be increased to 70% minus 3 in. Following this experiment and attainment of steady-state conditions, we will undertake an evaluation of unit operations of the air classifier and the trommel screen.

At a later date, depending upon results of the program, a series of additional experiments are contemplated to further refine the process. These are to evaluate pH and nutrient requirements and to study residue dewatering and uses. Increased activity for cellulolytic organisms has been reported at increased pH (up to 7.5). Experimentation will be undertaken to explore the benefits of higher pH levels and to compare them to the costs of maintaining these higher pH levels. Similarly, the addition of nutrients (nitrogen and phosphorus) will be optimized and the nutrient and other effects of sewage sludge evaluated. Residue characteristics and the potential of other dewatering devices will be explored. The potential value of liquid recycle has already been mentioned and experimentation conducted. If the solids content of the filter cake can approach 35%, then the cake can be used as a fuel to power the process. In addition, the value of the residue as a fertilizer and soil conditioner will be evaluated.

Data Generated to Date

As of this date of this paper, only experiments 1 and 1A have been started. The data generated is extensive, but, of course, can serve only as a baseline for current and future experiments, rather than predict optimum conditions.

Present indications are that gas is being produced at a rate of 4500 scf per input ton of raw refuse, and has a typical composition of 55% methane 45% carbon dioxide. The next few months will provide more substantial clues as to the potential of the process, as the thermophilic system is explored.

EXHIBIT 1

REFCOM - REFUSE CONVERSION TO METHANE

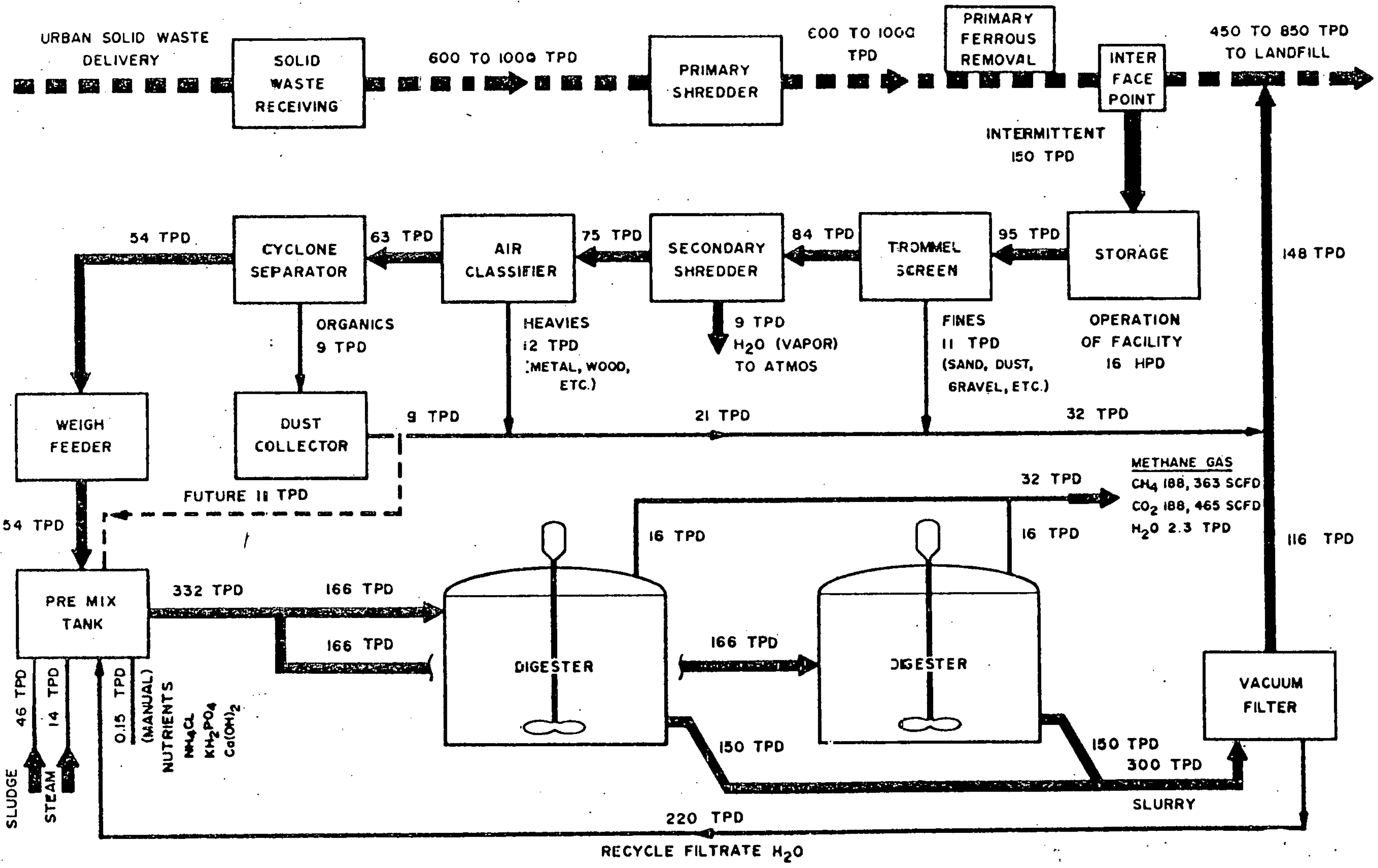
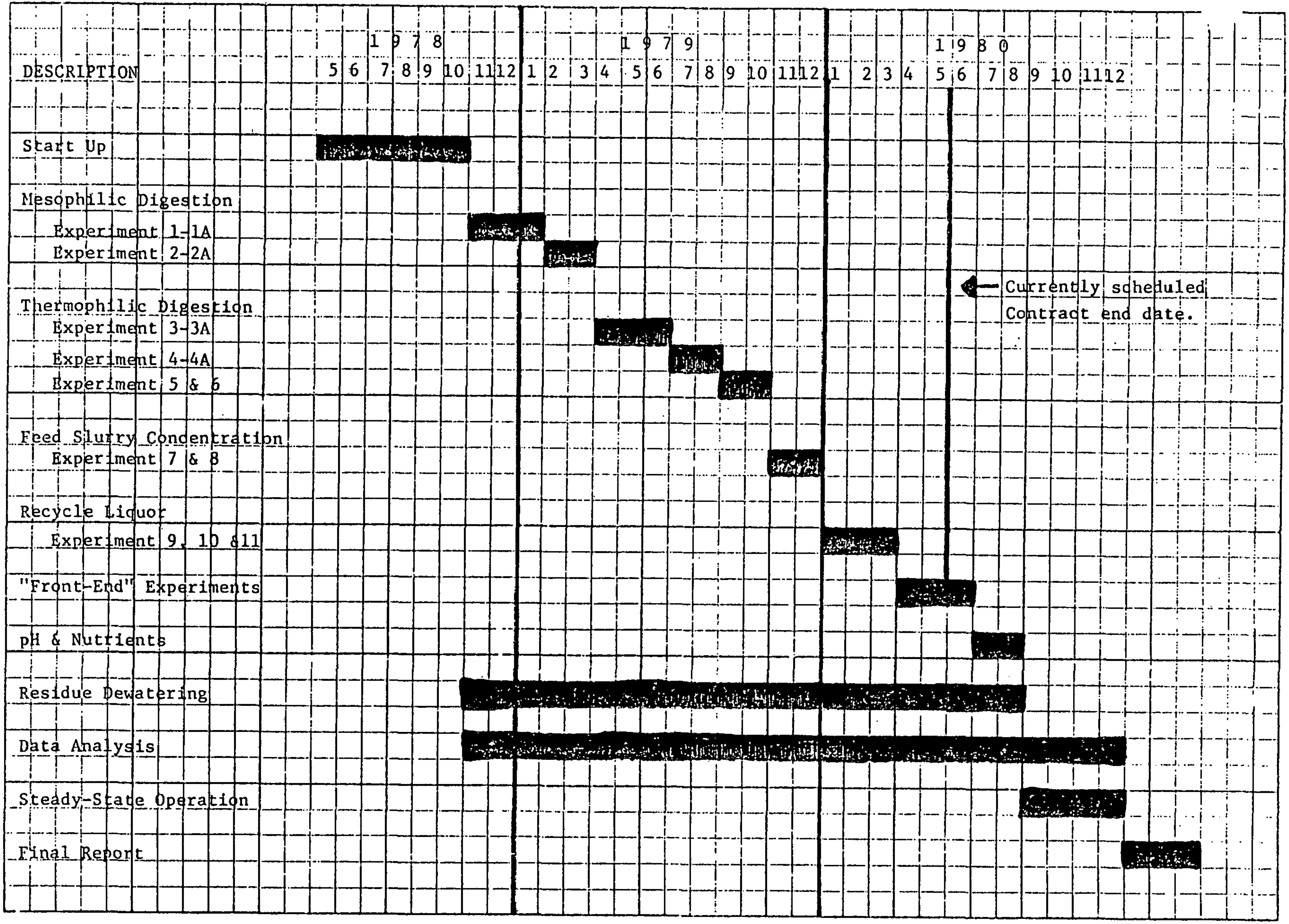


EXHIBIT 2


DESCRIPTION
1978
5 6 7 8 9 10 11 12
1979
1 2 3 4 5 6 7 8 9 10 11 12
1980
1 2 3 4 5 6 7 8 9 10 11 12
Start Up
Mesophilic Digestion
Experiment 1-1A
Experiment 2-2A
Thermophilic Digestion
Experiment 3-3A
Experiment 4-4A
Experiment 5 & 6
Feed Slurry Concentration
Experiment 7 & 8
Recycle Liquor
Experiment 9, 10 &11
"Front-End" Experiments
pH & Nutrients
Residue Dewatering
Data Analysis
Steady-State Operation
Final Report
Currently scheduled Contract end date.

REFERENCES

1. Pfeffer, J. T., "Reclamation of Energy from Organic Refuse", Grant No. EPA-R-80076, Office of Research & Monitoring, U.S. Environmental Protection Agency, National Environmental Research Center, Cincinnati, Ohio (April, 1973)

2. Pfeffer, J. T., and Liebman, J. C., "Biological Conversion of Organic Refuse to Methane", Progress Reports (July, 1973-June, 1974), NSF Grant GL-39191 (July, 1974), January, September, 1975.

3. Kispert, R. G., et al., "Fuel Gas Production from Solid Waste," Final Report, NSF Contract C-827 (January 31, 1975).

4. Bisselle, C., et al., "Urban Trash Methanation Background for a Proof-of-Concept-Experiment", NSF-RA-N-75-002 (February, 1975).

5. University of Miami, Proposal, "Characterization and Environmental Investigation of ERDA's Anaerobic Digestion Facility at Pompano Beach, FL."

6. Walter, D.K., Rines, C., "Status - Refuse Conversion to Methane (RefCOM)", (December, 1978).

DESIGN AND EVALUATION OF A METHANE GAS SYSTEM FOR A HOG FARM

Wayne Turnacliff, P.E.
Micheal Custer
Mary A. Veatch
Bio-Gas of Colorado, Inc.
5611 Kendall Court
Arvada, Colorado 80002

BSTRACT

he objective of the study was to etermine the economic feasibility of onstructing and operating anaerobic igestion systems on small hog farms. 200-gallon (760 liter) pilot digester as operated on hog waste, but the olatile acids never stabilized and the un was terminated. The cause of the igestion failure appeared to be an ntibiotic which is fed to the hogs as control for swine dysentery. However, farm system was sized to provide the quivalent of $300 per month of propane. he total required investment was $112,800 esulting in a gas production cost of 16.50 per million Btu ($15.64 per illion kJ). To approach economic easibility, a small farm anaerobic igestion system would have to be wner-built.

NTRODUCTION

he objective of this study was to stimate the amount of methane that ould be produced by anaerobic digesion from the hog wastes of the ee Pinello, Half a Hill Hog Farm in t. Morgan, Colorado, for the purpose of valuating the economic feasibility of onstructing and operating anaerobic igestion systems for the many small hog arms in Colorado. Gas output, capital equirements and operating costs were eveloped to evaluate economic easibility. Gas output was determined y a two-month laboratory pilot plant un in the Bio-Gas 200-gallon (760 iter) digester. Capital costs were etermined by obtaining quotes and stimates on a design and drawings eveloped by Bio-Gas of Colorado. The perating costs were also developed from he design and estimates of operating arameters.

ABORATORY STUDY

he purpose of the laboratory study was [illegible] etermine the actual gas yield that [illegible] d be obtained from the hog wastes. Excellent gas production results were obtained by a group from the University of Missouri[1] digesting hog wastes and we had hoped to duplicate their success. They had obtained a yield of 8.8 cubic feet of gas per pound of organic matter added (0.55 cubic meters per Kilogram) or 5.2 cubic feet of methane per pound of volatile solids added (0.32 m^3/Kg). The highest yeild obtained in the Bio-Gas study was 4.2 ft^3 methane per lb. VS added (0.26 m^3/Kg), and this figure was used in the system design.

Fresh hog manure was collected at the farm in November 1978. The waste was analyzed for moisture and volatile content and a feed design was calculated. The system was operated at a retention time of 25 days, a loading rate of 0.15 lb. volatile solids/cu. ft. (2.4 g VS/l) and a solids concentration of 7.5 percent. The digester had been started on seed material from a previous study on feedlot manure. By November 16, the digester was on a full daily feeding of hog manure. The concentrations of calcium, magnesium, potassium and sodium were all at acceptable levels.

During November, the digester appeared to be stabilizing, gas production was high and the alkalinity to volatile acids ratio was also high. However, the volatile acids began a substantial rise in December, see Figure 1, and by January exceeded the system's alkalinity. From November to December, the volatile acids had increased by almost twenty times. The volatile acids were exhibiting a biological upset considion which resulted in a stuck or sour digester.

The major cause of the system's failure appears to be an antibiotic feed ingredient used to control swine dysentery. Lincomycin hydrochloride monohydrate is included in the corn-soybean ration at 40 grams per ton of feed. At this dosage, the concentration in the digestion feed slurry at 7.5 percent total solids was estimated at approximately four parts per million lincomycin. Samples of the manure and digester effluent were sent to the manufacturers of the drug, Upjohn Company, to determine the actual concentrations. The manure at 21.7 percent total solids had a concentration of 5.2 ppm lincomycin and the digester effluent at 3.1 percent total solids had a concentration of 4.5 ppm. Since the manure was diluted with tap water to approximately 7.5 percent before being fed to the digester, it is surprising that the digester concentration was not a third or less of the manure concentration. However, this does show that the antibiotic is present in the waste.

Researchers at the University of Missouri[2] have experienced drastic reductions in biological activity in two separate occasions when wastes from hogs fed lincomycin were introduced to a digester. Both times the digesters went sour and could not be revived. The similarity between our system failure and the failures in Missouri strongly suggest that the pilot plant upset was due to the presence of the antibiotic lincomycin.

Lee Pinello has indicated that the lincomycin is very effective on his farm and his veterinarian does not recommend trying to switch to another drug. There also is no guarantee that another drug fed for swine dysentery control would not inhibit digester operation. As long as lincomycin is fed to the hogs, it appears futile to attempt anaerobic digestion for the production of methane gas.

DIGESTER EFFLUENT

An important consideration is the use of the digester effluent and its value. Historically, animal manures have been used as fertilizers or soil conditioners. This is the case at the Lee Pinello farm, where the manure is taken from the lagoons and spread on neighbors' fields. Even though Pinello delivers and spreads the waste, there is no charge as both parties benefit: Pinello gains by getting his lagoons drained and the neighbors gain from the fertilizing and soil amendment advantages of the hog waste. However, due to the nitrogen, phosphorus and potassium (NPK), the waste does have a value that can be determined. This calculation was done for the digester effluent and results in a daily value of $9.24 and an annual value of $3,370. The concentration of the NPK in the digester effluent was 4500, 1100 and 2300 ppm, respectively, and the nutrients were valued at twenty, ten and eighteen cents respectively. The amount of digester effluent available each day was 1,565 gallons (5920 liters). Over eighty percent of the nitrogen was in the form of ammonia which will tend to volatilize off when the digester residue is in the lagoons. The method of application to the fields will also result in some loss of nitrogen. Consequently, it was estimated that thirty percent of the nitrogen, sixty percent of the phosphorus and ninety percent of the potassium would effectively reach the fields. The fertilizer value was calculated for only the amount of NPK which would reach the fields.

The digester effluent can also be used as a feed product. Generally, the effluent is dewatered and the residue mixed with the normal feed ration. Refeeding of manures to cattle has been employed for quite a while, altho it does not appear to be as popular with hog farms. "Hog Farm Management," a trade magazine, had three articles on refeed in 1978. It appears that the reclaimed waste is fed only to the sows and not to the feeder pigs which go to market. Feed bill savings were reported[3] as $75,000 to $100,000 a year from using what was called "reclaimed antibody feed" (RAF) which was dewatered manure. Obviously, there is great potential in the refeeding of the digester residue but not enough is known at this time to estimate a value for the residue.

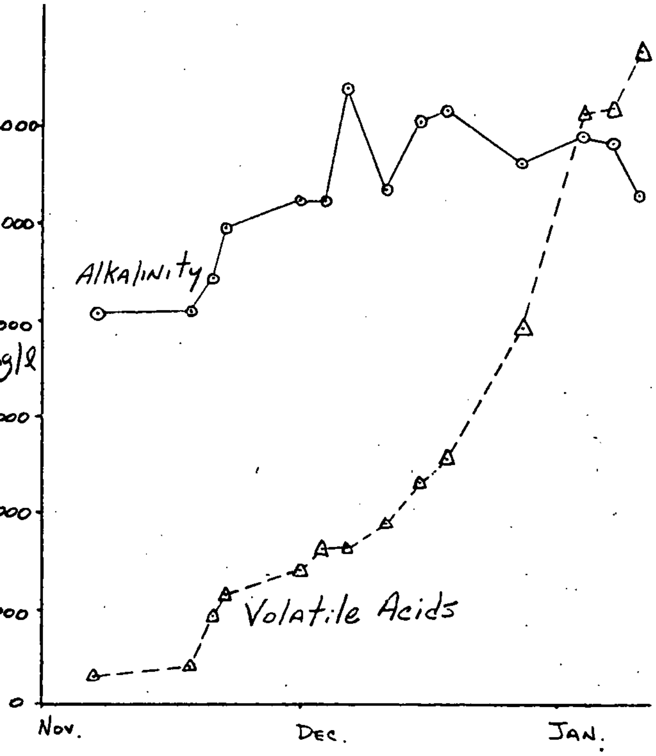
CURRENT ENERGY USAGE

The farm's energy use is a combination of propane and electricity. The propane is supplied by VanGas and the electricity by Morgan County Rural Electric. The propane is used for heating the home and farrowing building. Although there are two separate tanks for the home and farm, separate records are not kept. The actual propane charges for 1977 are shown in Table I where heating degree days were used to smooth and estimate the gallons of propane required each month at a price of 38¢/gallon ($4.06 per million kJ). The total propane bill for home and farm for 1977 was $2,452.96. The eleven hundred square foot home cannot removed from the propane usage estimate and it should be realized that the home is not insulated. Electrical use also is a combined total for the home and farm. Electrical consumption was 43,638 kilowatt hours (157,100 MJ) for 1977 at a cost of $1,475.72. This represents a cost of 3.4 cents per kilowatt hour (0.9 cents/MJ).

TABLE I: ENERGY USE AND REQUIREMENTS
Energy Use and Requirements

1977 Month	Denver Degree Days	Actual Propane Charges	Estimated Gallons of Propane	Estimated Energy Consumption Btu (10^6)	Available Energy Btu X 10^6	Additional Required or (Excess) Btu X 10^6
Jan	1105	794.81	1245	111	61	50
Feb	749	-	845	75	61	14
Mar	771	262.24	690	61	61	0
Apr	414	339.94	895	80	61	19
May	137	129.78	340	30	61	(31)
Jun	-	125.00	330	29	61	(32)
Jul	2	100.12	265	24	61	(37)
Aug	14	-	270	24	61	(37)
Sep	38	207.64	275	24	61	(37)
Oct	358	111.29	295	26	61	(35)
Nov	737	201.89	530	47	61	(14)
Dec	920	180.25	475	42	61	(19)
Total	5245	2,452.96	6455	573	732	83 MMBtu add'l required
						(242) MMBtu

FIGURE 1:
ALKALINITY & VOLATILE ACIDS



SYSTEM DESIGN

Design Parameters

System component sizing is determined by the amount of gas to be produced. In this case, the system was designed to produce the energy equivalent of \$300 worth of propane per month. Based on an average 30-day month and considering a 70% boiler efficiency, the required rate of bio-gas production is 3.67 ft^3/min (1.73×10^{-3} m^3/s). This bio-gas production rate plus a yield of 4.2 ft^3 methane/lb. VS added (0.26 m^3/Kg) and a loading rate of 0.14 lb. VS/ft^3 (2.24 Kg/m^3) set the size of the digester at 40,000 gallons (151,400 l). The use of a conservative loading rate is desirable because it allows some flexibility in digester operation. Selection of all other components relating to digester operation was based on digester size and gas production.

Digester-heating component sizes were set by the requirement of maintaining a constant digester temperature of 95°F (35°C) on a -15°F (-26°C) ambient day and a total heat loop loss of 160 Btu/hr -° F (99 kJ/hr -°C). A figure of 139 Btu/hr -°F was calculated by Ecotope[4] for tank heat loss only, for a similar system. Under these conditions, the digester temperature would drop 1°F (from 95°F to 94°F) in 18.93 hours. A 75,000 Btu/hr (79,000 kJ/hr) output heat exchanger would raise the temperature of the digester 1°F (from

94°F to 95°F) in 5.80 hours. This results in a total cycle time (1°F loss and 1°F gain) of 24.73 hours. It is a simple matter to control temperature to 0.5°F and reduce the cycle time to approximately 12 hours. Controlling the digester temperature this closely is very important in maintaining gas production and digester health. Therefore, a 75,000 Btu/hr. output heat exchanger was selected for digester heat maintenance.

The hot water boiler was sized to supply heat to maintain the digester's operating temperature and to raise the temperature of the incoming feed material to digester temperature. As previously discussed, the heat exchanger for digester heat maintenance was sized for 75,000 Btu/hr output. Assuming a 75% efficient heat transfer in the heat exchanger, a boiler output of 100,000 Btu/hr. (105,000 kJ/hr) is required for digester heat maintenance.

The digester requires a total of 1610 gallons (6090 l) of feed per day. This feeding is to be done over an 8-hour period resulting in an average flow to the digester of 200 gallons per hour (757 l/hr). Assuming the feed material is at ground water temperature (55°F), then an additional 66,640 Btu/hr is required to bring it to digester operating temperature (95°F). This brings the total boiler heating requirement to 166,640 Btu/hr. A boiler was selected to provide 175,000 Btu/hr (184,600 kJ/hr) output resulting in about a 10,000 Btu/hr margin above the total heat demand.

Due to the excessive gas consumption of the digester heating system on extremely cold days, it was decided to design the system to supply all of the farm and home heating requirements on days of 40°F (4°C) ambient or warmer days. Propane back-up would be required on colder days. At an ambient temperature of 40°F, the gas demand for digester heat maintenance and feed heating is 1.35 ft^3/min of bio-gas. Since this figure is such a large percentage of the total gas production, an influent/effluent heat exchange is considered essential to reduce heating gas consumption. A fifty percent reduction of the feed-heating requirement would result in heating gas demand of 0.91 ft^3/min of bio-gas. This is approximately twenty-five percent of the total gas production.

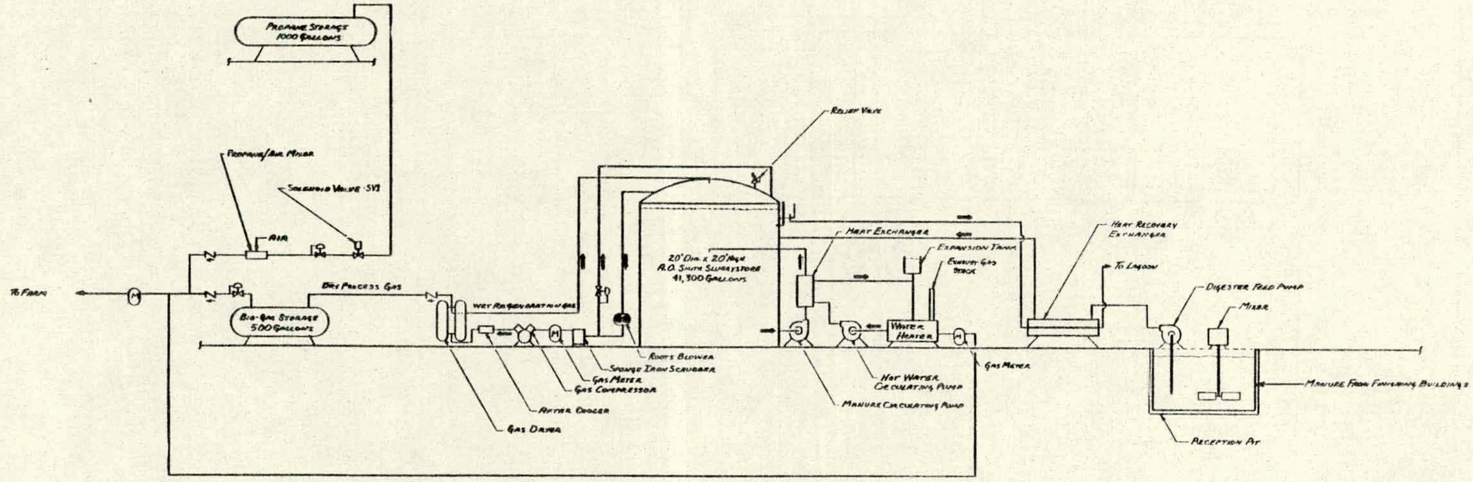
Digester mixing is required to maintain gas production at the specified rate. Three methods of mixing were investigated. They were gas recirculation, mechanical, and pump agitation. A gas recirculation mixing system was sized, quoted and immediately discarded due to the high price. A 30 Hp paddle-type mixer was considered next. Price of the mixer and roof modifications required to support it made the cost of this method prohibitive also.

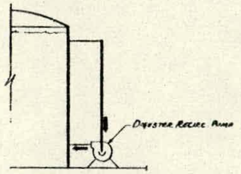
The most economical method investigated was pump agitation. This is accomplished by mounting four pumps equally spaced around the digester vessel and recirculating the digester contents with sufficient velocity to put the solids in suspension. In this case, the pumps, rated at 600 gallons per minute each, draw material from the top of the digester and discharge it tangentially at the bottom to achieve a swirling effect. This method of mixing is used on high digestion rate sewage plant digesters from 35 feet to 100 feet in diameter.

Description of Plot Plan and Schematic

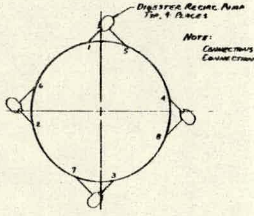
The system location is shown on the plot plan, drawing No. 10042. A location approximately half-way between the finishing buildings and the lagoons was selected to minimize the length of piping from the digester to the manure reception pit and to the lagoons. A reception pit between the two finishing buildings would have to be installed to collect manure for feeding the digester. A weir arrangement between the manure pit and the reception pit would allow the solids to enter the reception pit. Liquids collected in the manure pits would exit to the lagoons through the existing stand pipes. The digester effluent pipe would connect with existing manure pipe from all other farm buildings upstream of the lagoons. The open areas of the finishing buildings must be covered to prevent freezing of the manure pits. The reception pit would be provided with a cover to prevent freezing and provide a walkway.

All electrical apparatus on the gas-handling equipment must be explosion proof whereas the digester-heating equipment electrical apparatus need not be, unless it is housed in an enclosure with the gas-handling equipment. Therefore, two separate enclosures were provided. These enclosures would be only large enough to cover the equipment and would be removed for maintenance. The equipment enclosures protect the equipment from the weather and elimina the necessity of freeze protection during the winter.





Digester Recirc. Pump Installation
Typ. 4 Places - 90° Apart



Digester Recirc. Pump Suction And Discharge Piping Orientation

Note:
Connections 1, 2, 3, & 4 Are Pump Suctions
Connections 5, 6, 7, & 8 Are Pump Discharges

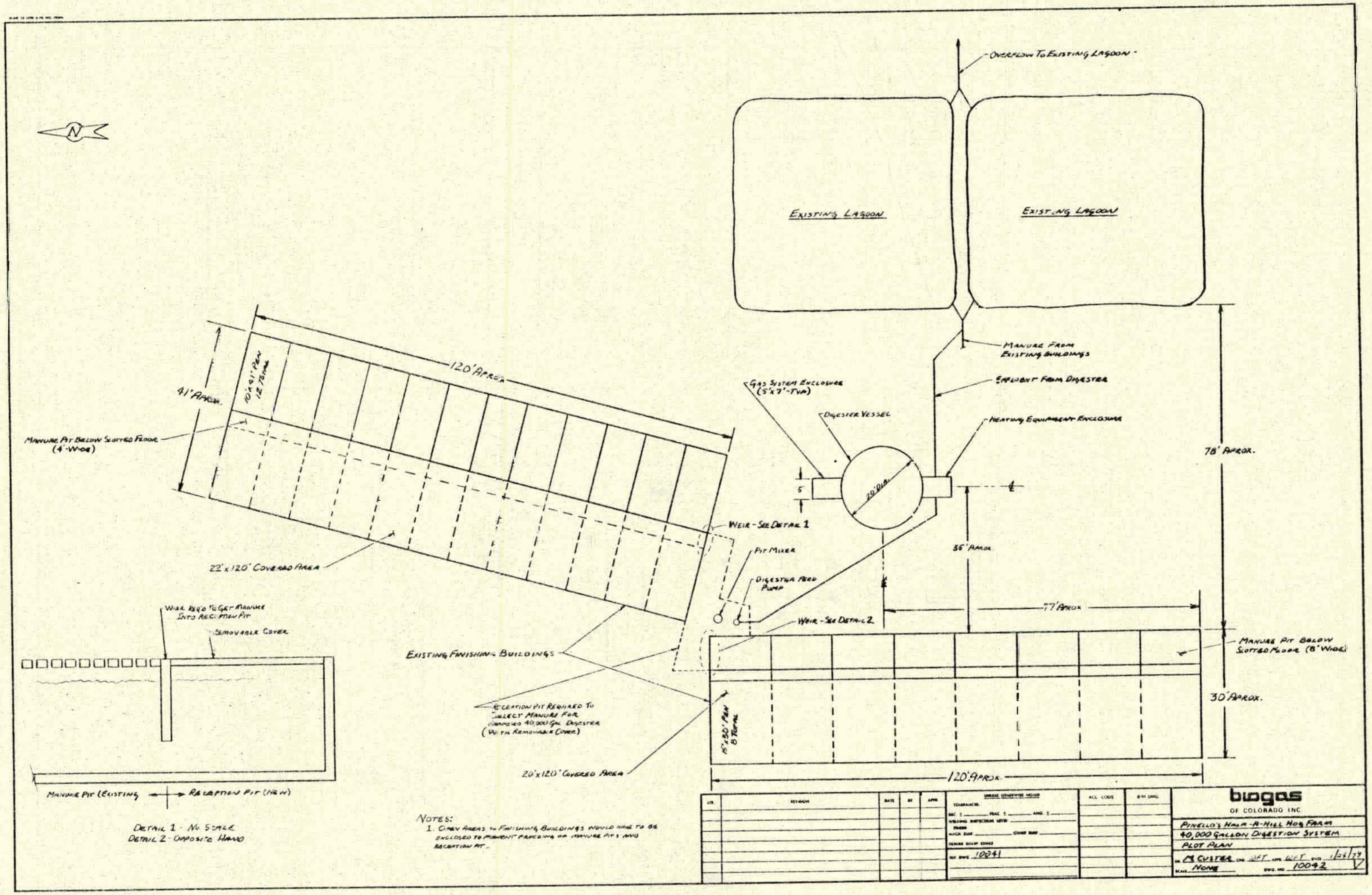
Note:

1. Equipment Enclosures Not Shown For Clarity.
2. Propane Stand-by is Activated by Two-Setpoint Pressure Switch on Bio-Gas Storage Tank Which Controls Solenoid Valve SV3.
3. Blower, Compressor, & After Cooler Are Are Activated by a Pressure Switch on the Digester near [illegible] - on at ¼ Psig, off at 0 Psig.
4. Manure and Hot Water Circulating Pump are Both Operated by a Temp Controller on the Digester
5. Digester Feed Pump and Pit Mixer Are Both Operated by Timers.

LTR	REVISION	DATE	BY	APPR.	UNLESS OTHERWISE NOTED	ACC CODE	SIM DWG
					TOLERANCES: DEC. ± ____ FRAC. ± ____ ANG. ± ____ WELDING INSPECTION LEVEL ____ FINISH MACH SURF ____ OTHER SURF ____ BREAK SHARP EDGES FOR DWG 10042		

biogas
OF COLORADO INC

Anglio's Half-A-Mile Hog Farm
Proposed 40,000 Gallon Digestion System
System Schematic
DR M. Custer CKD WFT APP WFT DATE 1-29-79
SCALE None DWG NO 10041 V



Detail 1 - No Scale
Detail 2 - Opposite Hand

Notes:
1. Open Areas to Finishing Buildings would have to be enclosed to prevent freezing of manure pits and reception pit.

biogas
OF COLORADO INC.

Pinello's Hala-A-Hill Hog Farm
40,000 Gallon Digestion System
Plot Plan

LTR.	REVISION	DATE	BY	APPR.

UNLESS OTHERWISE NOTED
TOLERANCES:
DEC ± ______ FRAC ± ______ ANG ± ______
WELDING INSPECTION LEVEL ______
FINISH
MACH SURF ______ OTHER SURF ______
REMOVE SHARP EDGES
REF DWG 10041

ACC. CODE
B/M DWG

DR. M Custer CHK. WFT APPR. WFT DATE 1/26/79
SCALE None DWG. NO. 10042 V

Th stem schematic, Drawing No. 10041, sh the arrangement of equipment in he system. Manure in the reception it is agitated by a paddle mixer. anure is pumped from the reception it to the digester by a positive isplacement pump. The effluent from he digester flows by gravity through concentric pipe influent/effluent eat exchanger where it transfers heat o the digester feed material. From his heat exchanger, the digester ffluent flows by gravity to the lagoons.

he digester heating circuit is a closed oop system with heat being supplied y an 84 gallon, 175,760 Btu/hr output ater heater. This water heater supplies ot water to a centrifugal pump which orces the water through the shell side f a shell-and-tube heat exchanger and ack to the hot water heater. Digester ontents are drawn off the bottom of he digester by a positive displacement ump, forced through the tube side of he heat exchanger, and returned to the enter of the digester near the top.

as is collected in the heat space bove the liquid level at a maximum ressure of one-quarter psig. This aximum pressure is set by the standard ank roof design. A higher pressure ould require a new roof design which ould significantly increase the cost. relief valve on the tank roof is ecessary to prevent the internal ressure from exceeding one-quarter sig. Gas is piped from the digester ead space to a blower which boosts he pressure to a maximum pressure of ight psig. Operating pressure is aintained by a pressure regulating alve which bleeds gas back to the igester head space. The gas is ressurized to force it through a ponge iron scrubber for hydrogen ulfide removal and a meter. The pressure egulating valve can be adjusted to chieve this pressure.

he gas compressor boosts the gas ressure to one hundred psig. The gas hen goes to an after cooler to be ooled down to approximately 100°F, then o a dryer to be dried to a -40°F ressure dew point, and finally to a 00-gallon propane tank for storage. The egeneration gas line off the dryer s piped back to the digester head pace so that no gas is wasted. Excess ater in this gas stream will condense ut in the head space as it cools since he gas there is saturated with water t 95°F.

pressure is reduced to line pressure gas pressure regulator as the gas leaves the 500-gallon storage tank. This gas goes to the farm and to the hot water heater through gas meters. The gas is stored in a 500-gallon propane tank to provide enough surge capacity to get through peak gas use periods without going on propane back-up.

If propane back-up is required because of peak gas use demand or extremely cold weather or both, then propane is mixed with air in a propane/air mixer to reduce the energy value to that of bio-gas (approximately 500 Btu/ft^3) and put into the piping system feeding the farm and house and the hot water heater.

ENERGY PRODUCTION AND CONSUMPTION

The system will provide 3.17 million Btu per day (3.34 million kJ/day) and will consume an average of thirty-seven percent of the energy produced for system heat demands. There will be just over two million Btu available for use or 61 million Btu per month (64 million kJ/mo). Table I shows how the available energy impacts on the energy requirements: for three months additional fuel, propane, is required; for one month production matches use; and for eight months an excess of bio-gas is produced. The net result is that an additional 83 million Btu (88 million kJ) will have to be purchased and 242 million Btu (255 million kJ) are produced in excess of farm demands.

System electrical consumption was estimated by determining a duty cycle for the pumps, blowers, mixers, compressors, and after-cooler, and then calculating the energy usage from the horsepower hours. Daily electrical consumption was 123 Kw hr (443MJ) an at 3.4 cents per Kw hr the annual electrical consumption would be $1520.

System Components

Major items are listed and described in Table II, Materials and Cost List. It has been our experience that this equipment works well and reliably in handling and digestion of manure. We have never used an A.O. Smith slurrystore tank as a digester. However, these tanks are designed and have been in use for several years specifically for manure storage. The Ecotope Group[4] used an A.O. Smith slurrystore tank almost identical to the one specified in Table II as a pilot digester in Washington state. The side walls and top of this vessel would be covered with four inches of urethane foam insulation to minimize heat loss; the cost of this insulation is included in the $7,500 figure for heat tracing and insulation. The remainder of this cost is for providing freeze protection for the four digester recirculation pumps and piping for digester mixing.

The cost of the equipment in Table II results in a total materials cost of $64,900. About seventy percent of this cost is from suppliers' quotes.

Table II:
MATERIALS AND COST LIST

1 41,300 gallon A.O. Smith Slurrystore, with top + $2,000 for pump connections: $21,446.

1 Hot Water Heater: A.O. Smith model BT251, 84 gallon, 175,760 Btu/hr out: $876.

1 Heat Exchanger: $2,275.

1 Pump, Goulds model 3642: $185.

4 Recirculation Pumps, Goulds model 3196 MT: $8,248.

1 Blower, Roots model 315Xa gas pump: $1,800.

1 Sponge Iron Scrubber: $750.

1 Compressor, Corken model DA190K4FBAB, 5 Hp, 5 CFM at 100 psig: $3,850.

1 Air Cooled After Cooler, Wilkerson model 1/12 Hp: $550.

1 Gas Dryer, Wilkerson model D01AE, -40° dew point out: $800.

1 Pit Mixer, Philadelphia model MT-02, right angle drive, 2 Hp, 30" diam: $1,975.

1 Digester Feed Pump, Moyno model 1L6 CDQ, reverse rotation, 1 Hp: $1,055.

1 Propane/Air Mixer: $3,000

1 Pump (to pump manure through heat exchanger), Moyno 1L6, 30 gpm at 20 psi: $1,100.

3 Gas Meters, Singer-American meter, model AL-250: $300.

1 Gas Pressure Regulator (for Bio-Gas to farm) Fisher Type S301, 100 psig to 5" H_2O: $70.

1 Gas Pressure Regulator, Fisher Type S301: $70.

Misc. Hand Valves: $1,000.

Misc. Pipe and Fittings (including heat exchanger): $1,500.

Misc. Material for equipment enclosure: $1,500.

Misc. Heat Tracing and Insulation: $7,500.

Manure Pit (concrete and forms): $1,000.

Misc. Control Equipment: $2,000.

4 Valves, Rovang 4" M20A: $950.

4 Valves, Roveng 6" M20A: $1,100.

Total Material Cost: $64,900.

Table III:
CONSTRUCTION COST ESTIMATE

Materials	64,900	
Contingency (10%)	6,500	
Sales Tax (3.5%)	2,300	
Subtotal		73,700
Installation		
Labor, 92 days @ $10/hr	7,400	
Fringe (10%)	700	
Subsistance, Travel	4,000	
Subtotal		12,100
TOTAL Construction Cost		85,800

Table IV:
CAPITAL COSTS

Construction Cost	85,800
Contingency @ 5%	4,300
Engineering, 300 hrs @ $15/hr	4,500
Construction Management @ 9%	7,700
Administrative Costs	4,500
Total Facility Cost	106,800
Start-Up	3,000
Building Modifications	3,000
Total Required Investment	112,800

Costs

From the materials cost of $64,900, a construction cost estimate was developed. This is presented in Table III. A ten percent contingency and 3.5 percent state sales tax were added to the materials cost. The installation labor was estimated at 92 man da[illegible] and the labor, fringe, subsistance an[illegible] travel results in an additional $12,100.

T[illegible] brings the total construction cost t[illegible]35,800.

From the total construction cost, a total required investment was developed and this is presented in Table IV. A similar cost development in a Bio-Gas of Colorado final report to the Four Corners Regional Commission[5] was used as a guide. Estimates were made for a contingency, engineering, construction management and administrative costs, which, when combined with the construction cost, gave a total facility cost of $106,800. Biological start-up and mechanical shakedown were estimated at $3,000. The building modifications are specific to this site and are the extension of the roof and the enclosing of the finishing buildings to keep the pits from freezing. These modifications were estimated at $3,000. The bottom line total required investment becomes $112,800.

The annual operation and maintenance costs were estimated to be $3,045 and include electrical charges, $1,520, mechanical equipment replacement, $750, insurance, $400, and auxiliary fuel costs, $375. The mechanical equipment replacement cost was estimated at one percent of the total of materials cost plus contingency and sales tax. It was assumed that the facility would increase insurance costs and a $400 per year estimate was included. The yearly auxiliary fuel requirement had been estimated at 83 million Btu in Table I. Using a propane cost of 40 cents per gallon, or $4.50 per million Btu ($4.27 per million kJ), results in an annual cost of $375.

Principal and interest payments were calculated at nine percent interest for twenty years. This results in an annual payment of $12,400 and, combining this with the annual operation and maintenance costs of $3,045, yields a total annual production cost of $15,445. The gas production cost of $12,075 is obtained by subtracting the fertilizer value from the total annual production cost. Dividing this cost by the annual energy production of 732 million Btu results in a gas production cost of $16.50 per million Btu. Since the propane cost is included in the annual production cost, its energy value can be added to that produced, which lowers the production cost to $14.82 per million Btu. Both of these unit costs are extremely high when compared to propane at $4.50 per million Btu.

If the gas were considered to be worth [illegible]50 per million Btu, then the annual revenue would be $3,290. Including the fertilizer value gives a total revenue of $6,660. With this analysis, the system would have an annual loss of $8,785.

Low Technology Approach

Since the gas cost for the system designed came out so high, $16.50 per million Btu, a look at a low technology approach is warranted. The major portion of the total annual production costs was principal and interest on the total required investment and, if the total required investment could be lowered significantly, then the production costs could be lowered.

The low technology approach considered is usually some form of a pit where the wastes may gravity flow with little or no mixing. The gas is not cleaned nor compressed. Generally, dependability and gas production are sacrificed by this approach.

Without going through another design, a few estimates can be made and a total required investment developed for a low technology approach. If the gas production remained the same, which is doubtful, and the gas was valued at a propane worth of $4.50 per million Btu, then the gas revenue would be $3,290. Along with the ferilizer value, the total revenue would be $6,660. If annual operation and maintenance costs could be reduced to $1,000, mainly by drastically reducing the power requirements, then a principal and interest payment of $5,660 could be justified. At nine percent interest for twenty years, this could justify a total required investment of $51,600.

For this total required investment to be a realistic building figure, all start-up, administrative, construction management and installation costs will have to behandled by the farm labor and not charged to the project. An engineering package would have to be available, possibly from the state or federal government, which would supply the system design. In short, this would have to be a "do it yourself" project with basically free or low cost outside assistance for the system design.

CONCLUSIONS

In summary, it appears to be futile to try and operate an anaerobic digestion system where the antibiotic lincomycin is utilized. Also, from the economic analysis performed, the gas cost is prohibitvely high if a company were to design and construct a system for a client. It appears that, in order for a system to approach economic feasibility, it will have to be owner built. Initial design and proof of concept will have to be provided at a low cost to the farmer and may have to be accomplished by state or federal government. Additional economic incentives are required if small farm anaerobic digestion systems are to be developed.

ACKNOWLEDGEMENTS

The authors would like to thank Lee Pinello and his family for their time, efforts and enthusiasm. Their participation definitely added to both the content and enjoyment of this project. We also thank Dr. C.J. Farho, with the Upjohn Company, for performing the tests for the antibiotic lincomycin. Without his help, we would have not been able to prove the existence and show the concentration of the lincomycin.

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HEAT TREATMENT OF BIOMASS FOR INCREASING BIODEGRADABILITY

Perry L. McCarty, Lily Young, W. Owen, D. Stuckey, and P. J. Colberg
Department of Civil Engineering
Stanford University, Stanford, CA 94305

ABSTRACT

Autohydrolysis of a white fir chip slurry from 150° to 225°C without oxygen and at an autogenic pH of 2.5 resulted in the partial hydrolysis of cellulosic components and an increase in anaerobic biodegradability to methane. Excessive temperature or time resulted in the formation of toxic materials and/or degradation of biodegradable materials into refractory residues. Staged autohydrolysis pretreatment for maximizing conversion of forest product residues to biodegradable materials offers a potential for conversion of 40 percent of the combustion energy into methane. Studies on biodegradation of products from thermochemical treatment of lignin at pH 13 demonstrated that the lower molecular weight fractions and single-ring aromatic compounds are bioconvertible to methane. Studies on thermochemical pretreatment of nitrogenous materials indicate that while mild treatment increases biodegradability, the biodegradation of amino acids, DNA and RNA are hindred through formation of refractory and toxic products. Conditions which minimize these adverse effects are being sought.

INTRODUCTION

The objective of this study is to evaluate thermochemical pretreatment as a method for increasing the anaerobic biodegradability of organic materials so that they can be more readily fermented to methane gas, a potential source of fuel. The study has five specific phases: 1) biological conversion of lignocellulose to methane, 2) biodegradation of aromatic compounds, 3) biodegradation of lignin and lignin fractions, 4) pretreatment of nitrogenous organics for increasing biodegradability, and 5) biochemical methane potential and toxicity testing. This report covers results for the first four phases.

BIOLOGICAL CONVERSION OF LIGNOCELLULOSE TO METHANE

Lignocellulosics, the cell-wall materials of vascular land plants, are the most abundant renewable source of chemical energy on earth, yet their potential remains highly under-utilized, primarily because of their complex structure. Anaerobic digestion has the ability of capturing up to 90 percent of the heating value (heat of combustion) of lignocellulosics in the form of methane gas. How[illegible]r, in practice, only a small fraction is conver[illegible] to methane due to the inability of anaerobic microorganisms to attack the intricate matrix. Lignocellulose biodegradability in general (by aerobic and anaerobic organisms) is limited by its complex structure which physically restricts the accessibility of hydrolytic enzymes.

Aqueous heat treatment without chemical addition (autohydrolysis) was investigated as a pretreatment means for physically separating lignocellulose constituents from the matrix in order to improve bioconversion efficiency to methane in anaerobic digesters. Oxidative and nonoxidative heat treatments, ranging from 150 to 250°C, were evaluated for relative effect on representative complex lignocellulosics and individual components of lignocellulose (cellulose, hemicellulose, and lignin).

Laboratory-scale studies were used for assessing pretreatment potential. Heat treatment was conducted in a two-liter, bomb-type autoclave. The biochemical methane potential (BMP) of samples was monitored with batch, anaerobic serum-bottle techniques which employed long-term incubation periods (thirty days) to permit acclimation to toxic or uncommon products [1]. Similar techniques were used for the anaerobic toxicity assay (ATA) to measure adverse effects of a compound, or sample, on the rate of total gas production from an easily utilized, methanogenic substrate.

Preliminary studies were conducted to compare oxidative and nonoxidative autohydrolysis. Autohydrolysis occurs as a result of a low autogenic pH of about 2.5 due to formation of organic acids from concentrated slurries at elevated temperatures. Two lignocellulose sources selected for study were peat and non-inked newsprint. Peat is enriched in ligneous substances and contains few carbohydrates. Non-inked newsprint was selected as a model complex lignocellulosic. A four-percent total solids (TS) slurry was employed for heat treatment of peat, and 1.5-percent TS slurry for newsprint. Heat treatments were for one-hour duration.

Without heat treatment and because of the high lignin content, peat was not biodegradable anaerobically and newsprint was relatively nonbiodegradable (bioconversion efficiency = 3 percent). Autohydrolysis, oxidative and nonoxidative, improved the bioconversion efficiency of these materials by solubilizing (i.e. hydrolyzing) or releasing organics from the matrix. A comparison of oxidative and nonoxidative hydrolysis of newsprint, illustrated in Fig. 1, shows the recoverable percent of the initial energy content, represented by chemical oxygen demand (COD), as a function of pretreatment temperature. For oxidative treatment, the fraction oxidized was added to that which was bioconvertible to methane, since the oxidized portion is liberated as heat and is, therefore, potentially recoverable. The percent of recoverable energy was proportionally greater for oxidative than nonoxidative treatment, but most of this energy was represented by heat liberated through combustion of products. The biodegradable fraction, however, was actually lower for oxidative pretreatment which indicates that biodegradable products were selectively combusted under oxidative conditions. Solubilization of organics was determined to be the primary mechanism for increasing bioconversion efficiency. Biodegradability of the residual particulate fraction was not improved by treatment.

The effect of oxygen on solubilization was evaluated as a function of temperature and reaction time in a buffered system. The aim was to test the potential of oxygen for increasing the recovery of soluble organics at a normal autogenic pH since oxygen catalysis had been reported by Schaleger and Brink [2]. Representative constituents of lignocellulose were studied (cellulose, hemicellulose, and lignin) as well as a complex lignocellulosic, white fir. The results showed that the presence of oxygen decreased the recovery of solubilized organics for all conditions evaluated, including circumstances where oxygen had been reported as catalytic for hydrolysis of cellulose [2].

Further autohydrolysis studies were conducted with white fir at 6.5 percent TS to assess the potential for improving the bioconversion efficiency and also to test possible formation of toxic products as suggested by others [3]. The bioconversion efficiency of white fir was significantly improved by single-stage, nonoxidative treatment providing methane yields more than six times that of the untreated control. However, condensation re tions were found to reduce yields of biodegradable organics, and in order to minimize these effects and maximize overall bioconversion efficiency, staged treatment was deemed necessary.

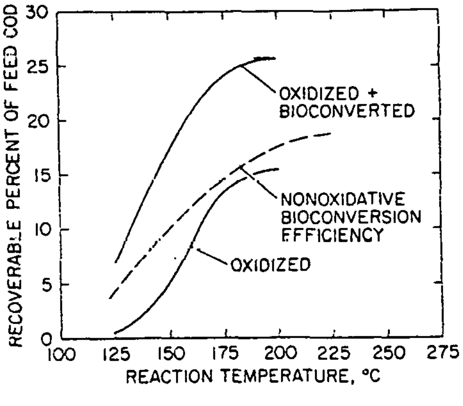

Fig. 1. Fraction of newsprint COD oxidized and/or determined bioconvertibility to methane at various heat treatment conditions with one-hour contact time. Oxidative reactions used air as feed and off-gas flow rate of 0.42 m^3(STP)/hr/kg feed COD.

The results of three-stage heat treatment of white fir are summarized in Table 1. Mild heat treatment was employed initially for solubilization of the easily hydrolyzed polysaccharides while minimizing condensation reactions. Then, more severe conditions followed in order to attack resistant polysaccharides, such as cellulose. The particulate yield of the final reaction product (41 percent) closely approximated the theoretical 40 percent chemical oxygen demand (COD) represented by the lignin fraction of the original solids, and the soluble yield (51 percent) represented approximately 85 percent of the theoretical carbohydrate COD. (Note: lignin would not be expected to be solubilized under these conditions.) Thus, a complex lignocellulosic was transformed into two easily separable fractions: a soluble fraction that originated predominantly from carbohydrates, and a particulate residue comprised primarily of lignin condensation products. Data from BMP assays further showed that about 40 percent of the initial heating value of white fir was recoverable as methane through three-stage treatment with fermentation of the soluble products. In such a scheme, the particulate residue could be wet oxidized to recover energy for heating the pretreatment reactor.

However, toxic products were present in all heat-treatment products and are attributed to soluble dehydration products (furan compounds) formed fro heat treatment of polysaccharides. Representativ results from toxicity assays are illustrated in Fig. 2. ATA data are shown here for two pure compounds (furfural and hydroxymethylfurfural) which have been reported as major dehydration products from heat treatment of polysaccharides, and the soluble product from nonoxidative autohydrolysis of white fir. Other reaction products behaved similarly. The lowest concentrations of all samples tested did not affect results relative to controls; therefore, control gas productions are omitted from Fig. 2 for clarity. Significant inhibition occurred at the intermediate sample concentrations (2.5-3 g/l COD), and severe inhibitio was present at higher concentrations. Anaerobic cultures were able to acclimate to these compound provided the soluble COD concentration was kept less than 3 g/l, as would be expected in typical fermentation systems. Therefore, toxicity is not anticipated to represent a major problem if adequately accounted for.

In summary, these studies showed that a relativel nonbiodegradable forest product residue can be separated into two fractions by autohydrolysis: 1) a soluble fraction that is highly biodegradabl and 2) a particulate residue that is not biodegradable. Furthermore, about 40 percent of the initial heat content is potentially recoverable b bioconversion to methane. The remaining par[illegible]late fraction could be wet oxidized to recove[illegible] associated energy for heating the pretreatment re actor in this system.

TABLE 1

Summary of 3-Stage Nonoxidative Heat Treatment of White Fir

Stage	Reaction Conditions		Feed[a] COD, g/l	Product COD, g/1			Percent of Original COD[b]		
	Temp. °C	Time, hr		Total	Soluble	Particulate	Loss	Particulate	Soluble
1	200	0.0	82.6	82.6	14.3	68.3	0.0	82.7	17.3
2	225	2.0	63.4	61.2	18.4	42.8	3.1	55.9	24.0
3	225	2.0	49.8	45.1	8.51	36.6	5.3	41.0	9.6
Reaction Sum							8.4	41.0	50.9

[a] Actual COD concentration for each reaction. First stage employed 1-liter liquid volume; for stages two and three, a portion of recovered particulates from preceding stage was resuspended in 0.75-liter liquid volume and assayed for COD concentration.

[b] Respective fraction referenced to the original feed COD of stage one, and based on percent particulate recovery of preceding stage as referenced to feed.

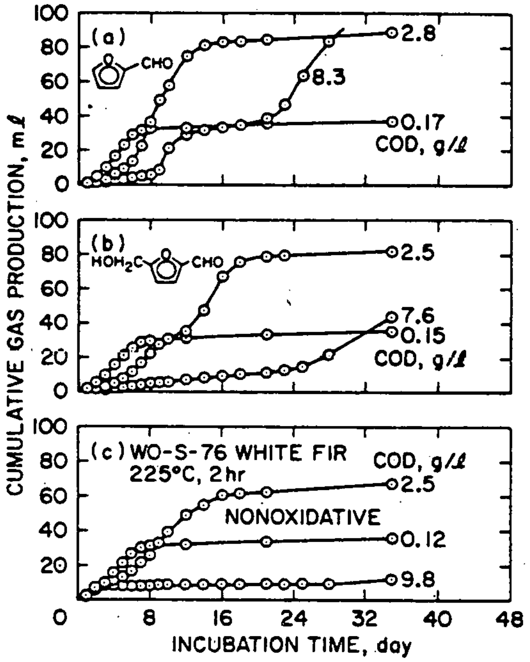

Fig. 2. Anaerobic toxicity results for a) furfural, b) hydroxymethylfurfural, and c) the soluble product from nonoxidative heat treatment of white fir at 225°C, 2 hr. Incubation temperature: 35°C, and fed control 35-day cumulative gas production: 32.6 ml.

One conceptual application for autohydrolysis-bioconversion technology for energy recovery from woody residues is presented in Fig. 3. In effect, the scheme employs: 3-stage, nonoxidative autohydrolysis for solubilization of polysaccharides, anaerobic fermentation for bioconversion of the soluble product to methane, and wet oxidation for energy recovery as heat from nonbiodegradable organics. The general theme is to minimize water use and heat transfer in order to limit associated energy losses. Therefore, all liquid-solid separations would be conducted at processing temperatures, and water would be reused as possible. Details of this system are presented elsewhere [4].

A cursory energy and mass analysis for processing 1.0 kg of woody residue with this scheme are as follows:

INPUTS:

Feed Residue	1.0 kg	20,000 kJ
Makeup Water	2.0 kg at 20°C	
Compressed Air	4.1 kg at 34 atm	
Nutrients	3.0 g Nitrogen 0.7 g Phosphorus	
Energy for mixing, centrifugation, mass transfer, etc. (conservative estimate)		2,500 kJ

OUTPUTS:

Ash Residue	0.05 kg solids 0.5 kg water	1,500 kJ
Biogas	0.11 kg CH_4 0.20 kg CO_2	6,100 kJ
Off-Gas (Wet Oxidation)	1.4 kg CO_2 3.1 kg N_2 1.6 kg steam at 25 atm	4,700 kJ
Heat as low-pressure steam		8,900 kJ

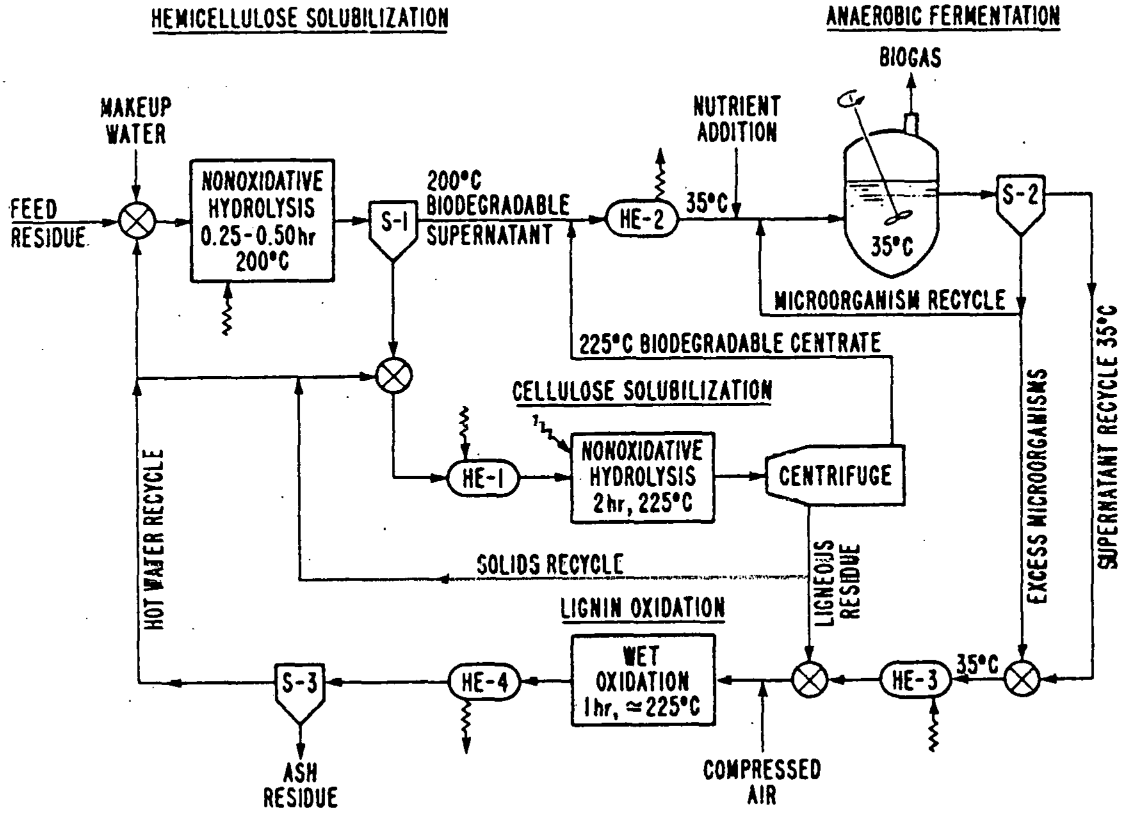

Fig. 3. Schematic diagram of possible application of autohydrolysis, bioconversion and wet oxidation for recovery of energy from wood residues. Notation, HE - Heat Exchanger, S - Settling, ⇝ Heat Transfer.

While this analysis contains many simplifications, it is clearly possible that the overall system could produce a net positive energy, and, in this regard, it represents a significant improvement over other pretreatment and bioconversion schemes, since they typically have net energy losses. A conservative estimate of system potential would be that the net system recoverable energy is the 6,000 kJ contained in the biogas, i.e., a 30-percent conversion efficiency of the feed residue to an easily transported, clean-burning source of fuel. The significant advantage of autohydrolysis is that chemical requirements are low.

BIODEGRADATION OF AROMATIC COMPOUNDS

While lignin does not appear to yield biodegradable products under autohydrolytic conditions, it is solubilized and to some degree fractionated by thermochemical treatment at pH 13. Our previous studies have shown that a wide variety of aromatic compounds released under these conditions are completely biodegradable to methane and carbon dioxide. Further studies on aromatic decomposition to methane has continued to focus on ferulic acid as a model compound. For determination of important intermediates formed during ferulic acid conversion, methane inhibition experiments were carried out with 2-bromoethane sulfonic acid (BESA), an analogue of Coenzyme M (CoM) and a specific inhibitor of methane formation. This procedure resulted in successful inhibition of gas formation to 20 percent of that in an uninhibited culture and a clear buildup of acetate to levels four times higher than normal. In addition, detectable levels of four other volatile fatty acids were also observed (Fig. 4).

Identification of additional intermediates was conducted on esterified ether extracts by gas chromatography/mass spectrometry. A number of ring compounds were detected as follows: 3-phenylpropionate, cinnamic acid, phenylacetate, benzoate, and cyclohexanecarboxylate. These compounds were not detectable in uninhibited cultures or in background controls, and their concentrations were not determined. In addition, the methyl esters of pimelic and adipic acids were also detected in samples, both with and without BESA inhibition. The levels were two to three times higher in inhibited samples and corresponded to approximately 10 mg/l.

Based on the intermediates observed, Fig. 5 illustrates a tentative conceptual model describing methanogenic degradation of ferulic acid. All the intermediates depicted on the top portion have been observed in our cultures. In addition, the detection of benzoate, propionate, and acetate in our cultures clearly supports the proposed overlap a[illegible] merging of the ferulic acid degradation scheme with

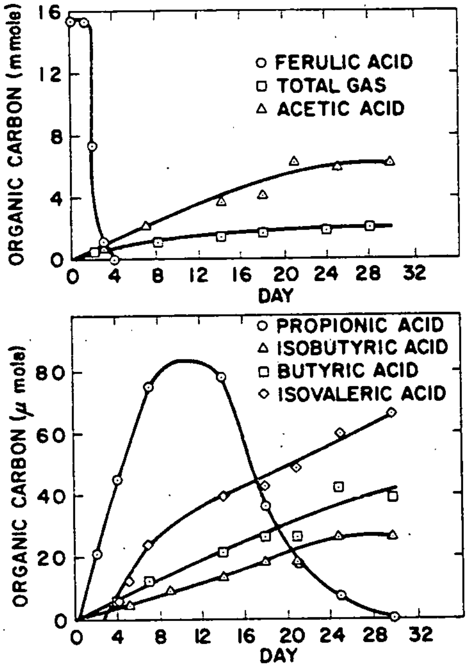

Fig. 4. Buildup of volatile acids in BESA-inhibited culture. A) Acetate buildup. B) Temporary buildup of propionic acid and appearance of isobutyric, butyric, and isovaleric acids.

that of benzoate. These data suggest that other aromatic derivatives can also be converted in a similar manner and that such anaerobic degradation mechanisms for methane formation may be common for aromatic compounds.

BIODEGRADATION OF LIGNIN AND LIGNIN FRACTIONS

While the previous section indicated that aromatic subunits of lignin are readily fermentable to methane and carbon dioxide, the potential for the methanogenic degradation of the more complex molecules following high pH thermochemical treatment of lignin is unknown.

In initial studies, alkaline heat-treated peat lignin was fractionated by gel filtration chromatography [6,7]. Elution of the peat lignin was by descending chromatography in a dioxane-water solvent system (1:1 by volume) in which the peat was soluble. Lignin elution patterns were determined by recording the optical density (OD) of each sample at 280 nm. Fractions of similar molecular weight were pooled and dried under vacuum by rotary evaporation at 50-55°C to be used as substrate for biodegradation studies.

Fractionation on Sephadex LH-20 gel (Pharmacia Fine Chemicals, Uppsala, Sweden), resulted in three major fractions: two low molecular weight peaks (MW 200 and MW 600), and one large, high molecular weight

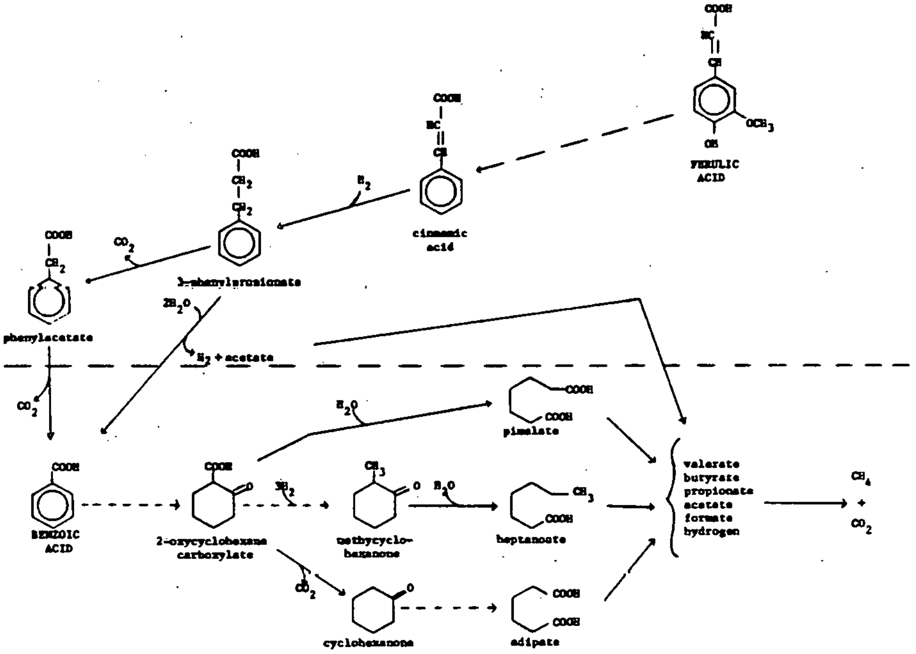

Fig. 5. Proposed conceptual model for the conversion of ferulic acid to methane incorporating previously reported benzoate pathways. Top half of figure shows proposed ferulic acid degradation steps. Bottom half of figure shows previously reported benzoate pathways as depicted by Evans [5].

fraction (MW 800-1400) (Fig. 6). Sephadex LH-60 was used to further elute Fraction 1, and although this revealed the presence of a broad range of higher molecular weight compounds, the majority of materials was still eluted in one major peak.

BMP assays were performed on the three fractions. The preliminary data suggest that Fractions 1 and 2 are more readily converted to methane (30% and 38% conversion efficiency, respectively) as compared to Fraction 3 (7% conversion efficiency). It appears that the observed increase in biodegradation following alkaline heat treatment of lignin results from formation of lower molecular weight compounds.

BIODEGRADATION OF NITROGENOUS ORGANICS

Nitrogen-containing organic materials are prevalent in animal manure, municipal wastewater, sludges, residues from slaughter houses and from the growing, harvesting, and processing of vegetables and fruit. Organic residues from aquaculture such as algae and aquatic plants are also high in nitrogen. Nitrogenous materials are also minor components of most lignocellulosic materials.

Proteins and their building blocks, the amino acids, are major forms of nitrogenous organics. Other major forms are the nucleic acids DNA and RNA and their purine and pyrimidine bases. In addition, nitrogenous organics are generally components of larger biological structures such as cell walls and cell membranes. Analyses of organics remaining after methane fermentation indicate that many nitrogenous organics are not readily biodegradable. Some nitrogenous materials are no doubt protected from biodegradation by other organics just as lignin protects cellulose from attack in lignocellulosics. In addition, some nitrogen-containing molecules are probably sufficiently complex to defy microbial attack. This phase of the overall study is being conducted to determine the degree to which thermochemical pretreatment can increase or otherwise change the biodegradability of such materials, and to assess whether or not the resulting nitrogenous organic products are toxic.

Both complex and pure nitrogenous organics were evaluated. The complex organics were bacterial cells (waste activated sludge, or WAS) from a municipal activated-sludge plant. Some evidence suggests that under thermophilic conditions organics are more completely degraded than under mesophilic conditions. Experiments were carried out on thermochemically pretreated WAS to evaluate this difference. BMP assays were conducted using seed from separate mesophilic (35°C) and thermophilic (55°C) digesters fed WAS.

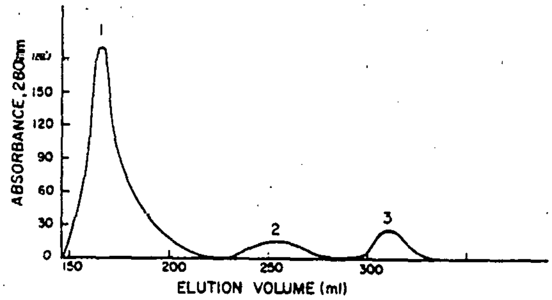

Fig. 6. Elution pattern of alkaline heat-treated peat on Sephadex LH-20 in 1:1 dioxane-water.

Data obtained with 81 day BMPs (Fig. 7) revealed the previously determined trends with temperature of treatment and NaOH addition, with maximum biodegradability occurring with pretreatment at temperatures of 175°C. The data also indicated mesophilic conditions led to higher degradabilities. These data should be interpreted with caution since it is possible that the thermophilic seed used may not have been a "balanced" population.

In another experiment, a different flow scheme was tried to improve biodegradability. The effluents from both mesophilic and thermophilic digesters receiving WAS were heat treated at 175°C for one hour without chemicals, and BMPs were carried out over a 44-day period to determine the increase in degradability of the digested materials. This two-stage flow scheme allows easily degradable organics to be consumed before subjecting the remaining refractory compounds to heat treatment. The data obtained indicated an overall degradability of the two-stage process of 75 percent, an increase from 62 percent without heat treatment.

Thermal treatment of WAS leads to an increase in degradability, followed by a precipitous decline with temperature after an optimum. In order to understand this phenomenon, the biodegradabilities after thermal treatment of the following common nitrogenous organic components were studied: proteins, amino acids, RNA, and DNA. Toxicity by the ATA procedure was also evaluated.

Two proteins, collagen and albumin, were evaluated for toxic effects, while only the former was used in biodegradability tests. The concentrations

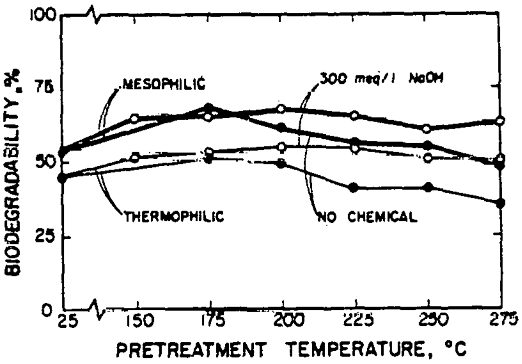

Fig. 7. Effect of heat treatment with and wit[illegible] chemical addition on biodegradability [illegible] WAS under mesophilic and thermophilic conditions.

se e approximately the same as for WAS (40 g D. However, since collagen is a fibrous protein and very gelatinous, a solution of only 20 g/l s used. The albumin used was 34 g/l.

e amino acid solution treated contained 2 g/l of ch of the twenty common amino acids. The DNA ed was extracted from salmon sperm and contained g/l. The RNA was extracted from yeast and conined 57 g/l.

e percent bioconvertible to methane as determined mesophilic BMP analysis for each material is mmarized in Fig. 8. The controls (i.e., no preeatment) indicate that all the basic components ere highly degradable as expected. The high vales for collagen are probably indicative of low easured COD values for the feed solution due to ncomplete oxidation of the highly bound protein.

hermochemical pretreatment in all cases but one owered the BMP. It is hypothesized that this was ue to a thermally catalyzed complexation and polyerization reactions where the simple components ere converted to more complex materials, which ere less susceptible to biodegradation. The exception to this is collagen where there appears to be o significant decrease with pretreatment.

he results of an anaerobic toxicity assay (ATA) on he heat-treated basic components are listed in able 2. Given are the maximum rate ratio (MRR) as function of dilution. MRR values less than 1.0 ndicate inhibition. Under mesophilic conditions

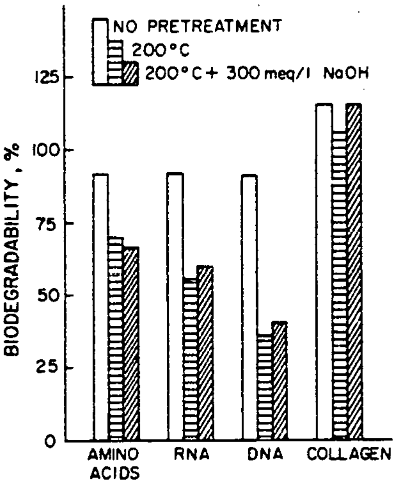

Fig. 8. Comparison of biodegradability for various nitrogenous chemicals without and with heat treatment for one hour.

TABLE 2

ATAs of Heat-Treated Nitrogenous Components[a]

Compound	Heat Treatment (One Hour)	Mesophilic - MRR (3 Days)			Thermophilic - MRR (3 Days)		
		1/30	1/5	1/2	1/30	1/5	1/2
Amino acids	Control	-	-	1.7	-	-	2.28
	200°C	1.2	1.0	0.6 (4)	1.38	2.31	1.99
	200°C NaOH[b]	0.85 (3)	0.6 (5)	0.6 (3)	1.26	1.04	0.72 (3)
DNA	Control	-	-	2.0	-	-	1.96
	200°C	1.1	0.42 (4)	0.2 (7)	1.13	0.31 (11)	0.31 (11)
	200°C NaOH[b]	0.95 (3)	0.55 (4)	0.15	1.06	0.40 (13)	0.34
RNA	Control	-	-	3.7	-	-	3.41
	200°C	1.0 (3)	0.49 (6)	0.29 (6)	1.33	1.43	0.64
	200°C NaOH[b]	1.0 (3)	0.69 (4)	0.37 (6)	1.20	0.75 (7)	0.59 (7)
Collagen	Control	-	-	1.98	-	-	4.23
	200°C	1.18	1.5	1.49	1.22	1.86	2.73
	200°C NaOH[b]	1.19	1.02	0.44 (4)	1.25	2.12	1.43
Albumin	Control	-	-	1.76			
	200°C	1.23	1.17	0.73 (21)			
	200°C NaOH[b]	1.17	1.19	0.90 (6)			

[a] Figures in parentheses indicate time in days to acclimate.

[b] 300 meq/l.

only the amino acid mixture was toxic at a thirtieth dilution, while at a fifth dilution most compounds were toxic, except for the proteins. At the highest concentration, the proteins were still considerably less toxic than the other components.

These results indicate that somewhat mild (175°C) thermochemical treatment can increase the biodegradability of complex nitrogenous materials such as bacterial cells. However, higher temperatures, with and without chemical addition result in chemical degradation of proteins, amino acids, and nucleic acids forming less biodegradable and also inhibitory products. This suggests that if thermochemical treatment is to be used for increasing biodegradability, then a two-stage process is desirable. Here, the biomass would be treated to obtain methane from the readily biodegradable materials. The residue would then be treated thermochemically to convert remaining persistent organics into biodegradable materials. Preliminary studies with such a scheme indicated an increase in biodegradability of bacterial cells from 62 percent with single-stage biodegradation to 75 percent with the two stages.

FUTURE STUDIES

The autohydrolysis process for increasing biodegradability of forest residues has good potential because it requires no chemical additions. The organic acids produced are biodegradable and by proper design of a methane fermenter, neutralizing of these acids can be avoided. Future studies will be concerned with optimizing the staged treatment process with respect to number of stages, time of each, and temperature. Evaluation of chemical products produced at each stage is also planned in order to understand the transformation which takes place. This will help not only in the optimization of the process but may lead to design of a process to produce valuable by-products other than methane.

Lignin is not transformed into biodegradable products by autohydrolysis. Products of alkaline thermochemical treatment will continue to be explored and analyzed in an effort to seek treatment conditions which will yield a maximum of fermentable products. Similarly, products from pretreatment of nitrogenous materials will continue to be explored in an effort to find the best conditions for convering refractory materials into biodegradable products while reducing the extent of conversion of biodegradable materials into either refractory or toxic materials.

ACKNOWLEDGEMENT

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ANAEROBIC FERMENTATION OF BEEF CATTLE AND CROP RESIDUES

Agreement No: DE-A101-79ET 20003
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Andrew G. Hashimoto
Yud-Ren Chen
Roman L. Hruska U.S. Meat Animal Research Center
Agricultural Research
Science and Education Administration
U.S. Department of Agriculture
Clay Center, Nebraska 68933

3STRACT

nis report summarizes the research being conducted t the Roman L. Hruska U.S. Meat Animal Research enter to assess the technical and economic feasiility of recovering methane and high protein iomass from the anaerobic fermentation of beef attle and crop residue. To accomplish this objecive, a pilot-scale fermentor was operated and an conomic assessment was performed based on the ilot-scale results.

ne highest methane production rate (4.65 L CH_4/L ermentor·d) was produced at a retention time of 5 ays, loading rate of 16.8 kg VS/m^3·d, hourly eeding, and fermentation temperature of 55°C.

n economic assessment of the thermophilic, anaeroic fermentation process for beef feedlots was perormed with various biogas handling and use otions. Option A consisted of CO_2, H_2S and H_2O emoval, compression, storage and use in an inteal combustion engine/generator to produce electriity with waste engine heat being used to heat the ermentor. Option B resembled A, except CO_2 was ot removed from the biogas. Option C resembled otion B, except the biogas was not compressed or tored. Option D resembled B, except the biogas as used as fuel for a boiler, rather than used in n engine/generator. Option E resembled D, except ne biogas was not compressed or stored.

ne fermentation plant size at which the electrical roduction costs equals the current rates of $12.50 er gigajoule (GJ) for electricity and $2.50/GJ for atural gas would be 24, 54, 29, 91 and 36 megarams of total solids per day (Mg TS/d) for Options , B, C, D and E, respectively. However, when a efeed credit of $60/Mg dried effluent is applied this analysis, the plant size at which energy roduction costs equals current energy prices are , 6, 4, 5 and 3.5 Mg/d for Options A, B, C, D and , respectively.

ITRODUCTION

ais report summarizes the research being conducted the Roman L. Hruska U.S. Meat Animal Research enter to assess the technical and economic feasi-lity of recovering methane and high protein [illegible]s from the thermophilic fermentation of beef [illegible]e and crop residues. Specific objectives are to:

1. Develop design criteria for optimum production of biomass and/or methane from anaerobic fermentation of livestock and crop residues,
2. Develop efficient methods to recover high protein biomass from the fermented residue,
3. Evaluate the nutritional value of the biomass as a livestock feed,
4. Determine the capital and operational costs; and energy, manpower and safety requirements for methane fermentation systems associated with livestock operations.

This project is jointly funded by the U.S. Department of Agriculture, Science and Education Administration, Agricultural Research ($150,000/yr) and the U.S. Department of Energy, Division of Distributed Solar Technology, Fuels from Biomass Branch ($60,000/yr). The specific objectives of interest to DOE are Objectives 1 and 4 listed above.

APPROACH

The approach used to accomplish Objectives 1 and 4 above, was to operate a pilot-scale fermentor and make an economic assessment of the system based upon the pilot-scale results. Detailed description of the pilot-scale system and procedures used in operating the system have been reported previously [1].

Pilot-Scale Fermentor Operation

A pilot-scale (5.7 m^3), thermophilic (55°C) fermentor has been in operation for over 2 years at the Roman L. Hruska U.S. Meat Animal Research Center converting beef cattle manure into methane (CH_4) and protein [1]. Table 1 summarizes the operating parameters of the pilot fermentor under steady-state conditions. Steady-state was assumed after three hydraulic retention times (Θ); therefore, the results in Table 1 are averages ± 1 standard deviation of the data during the fourth or longer retention times. The results in Table 1 are not completely comparable for the various retention

TABLE 1. SUMMARY OF OPERATING PARAMETERS FOR PILOT-SCALE FERMENTOR[a]

PARAMETER	RETENTION TIME, DAYS						
	20[b]	12	6	4	7	5	5 (once/hr)
TOTAL SOLIDS							
INF., g/l	83.1±6.0	70.1±5.4	74.4±7.3	67.7±4.7	92.4±2.7	92.8±8.9	94.7±5.9
EFF., g/l	50.4±0.7	36.6±2.7	43.0±3.8	43.8±1.7	47.0±0.8	46.4±1.9	51.6±2.0
CHANGE, %	-39.4	-47.8	-42.2	-35.3	-49.1	-52.0	-45.5
VOLATILE SOLIDS							
INF., g/l	68.5±6.1	61.8±5.3	68.7±8.7	59.5±4.5	82.6±2.1	84.9±8.4	83.8±5.1
EFF., g/l	38.2±1.0	29.2±2.8	37.0±4.6	35.8±1.5	37.1±0.5	39.8±1.7	41.5±1.7
CHANGE, %	-44.2	-52.8	-46.1	-39.8	-55.1	-53.1	-50.5
FIXED SOLIDS							
INF., g/l	14.6	8.3	5.7	8.2	9.8	7.9	10.9
EFF., g/l	12.2	7.4	6.0	7.9	9.9	6.6	10.1
CHANGE, %	-16.4	-10.8	+5.3	-3.3	+1.0	-16.5	-7.3
COD							
INF., g/l	77.1±11.3	74.9±13.2	73.8±3.1	73.0±3.6	93.1±11	93.7±11	96.1±13.1
EFF., g/l	54.8±3.8	40.2±7.1	47.2±3.5	47.8±2.4	55.0±3.6	52.9±5.2	56.9±9.2
CHANGE, %	-28.9	-46.3	-36.0	-34.5	-40.9	-43.5	-40.8
TOTAL NITROGEN							
INF., g/l	3.76±0.33	4.32±0.37	3.69±0.41	3.81±0.10	4.25±0.32	3.62±0.31	4.25±0.20
EFF., g/l	4.20±0.12	3.93±0.38	3.82±0.03	4.14±0.23	4.19±0.03	3.88±0.29	4.27±0.18
CHANGE, %	+11.7	-9.0	+3.5	+8.7	-1.4	+7.2	+0.5
AMMONIA-N							
INF., g/l	0.76±0.08	1.13±0.12	1.02±0.28	1.50±0.26	0.93±0.07	0.94±0.13	1.12±0.16
EFF., g/l	1.97±0.01	1.89±0.05	1.82±0.07	1.90±0.07	1.61±0.16	1.44±0.03	1.85±0.14
VOLATILE ACIDS							
INF., g/l	7.87±0.50	6.95±0.79	6.75±0.74	4.56±0.85	7.85±0.66	6.89±0.34	7.70±1.14
EFF., g/l	1.95±0.10	1.15±0.23	1.82±0.21	2.55±0.19	1.27±0.07	1.64±0.12	2.39±0.33
ALKALINITY							
INF., g/l	4.79±0.84	4.06±1.18	3.26±0.54	5.43±0.45	3.73±0.48	2.95±0.59	4.37±0.25
EFF., g/l	11.91±0.44	8.59±0.42	8.53±0.55	9.23±0.11	8.26±0.51	6.12±0.32	8.63±0.63
pH							
INF.	5.2	5.2	4.8	7.65	4.85±0.21	4.61±0.31	5.65±0.30
EFF.	8.2	7.9	7.9	7.93	7.87±0.04	7.70±0.05	7.71±0.12
METHANE, %	58	55	53	50	52	52	55
METHANE PRODUCTION							
L/L·day	0.75	1.59	2.73	3.14	3.57	4.23	4.65
L/g VS added	0.22	0.31	0.23	0.21	0.30	0.25	0.28
L/g VS utilized	0.50	0.58	0.50	0.53	0.55	0.47	0.55
L/g COD utilized	0.68	0.55	0.59	0.43	0.66	0.52	0.59

[a] Data presented as mean ± 1 standard deviation

[b] No mechanical mixing

times because of several necessary modifications in hardware and procedures. The slurry volume in the fermentor was maintained at 5.4 m^3 for θ of 20, 12 and 6 days and at 5.1 m^3 for θ of 4, 7 and 5 days.

Mechanical mixing was not used during θ of 20 days, which probably accounted for the lower volatile solids reduction at this retention time versus those at 12 and 6 days. During θ of 12 days, a mixer with two 0.267-m-diameter (D) propellers spaced 2.5 D apart on a single shaft was installed. This arrangement did not provide enough mixing to prevent scum formation. Before θ was reduced to 6 days, the fermentor was emptied and cleaned, and two 0.381-m-diameter propellers spaced 2.5 D ap[illegible] were installed. The fermentor was then refill[illegible] with the original contents, except for the scum.

ost the data in Table 1 are for a daily itl -feed cycle, where a designated volume depending on the retention time) was withdrawn rom the fermentor then replaced by a similar olume of fresh manure slurry. The exception is he last column where the fermentor was operated at 5-day retention time and fed hourly.

able 1 shows that the fixed solids and total itrogen did not decrease during fermentation, and he total gas and CH_4 production increased as θ ecreased. The CH_4 production was about 0.50 L H_4/g VS utilized, as reported by Morris [2]. owever, the CH_4 production relative to the COD tilized was about 70% higher than the theoretical ield of 0.35 L CH_4/g COD utilized. A possible eason for this discrepancy is that the COD of the iomass produced during the fermentation is eflected as effluent COD.

able 1 compares the results of operating the fermentor at 5 days retention time and being fed once aily and once hourly. The effluent concentrations f the various parameters were slightly higher when he fermentor was fed hourly. This was expected ince the effluent sample was taken about 22 hours fter feeding for the once daily feeding scheme, hereas the sample was taken less than one hour fter feeding for the hourly withdraw-feed scheme. hus, the daily-fed system approximated a semiontinuous system, whereas the hourly-fed system pproximated a continuously fed system.

he volumetric CH_4 production rate was about 10% igher during the hourly feeding (4.65 vs. 4.23 L H_4/L fermentor·d). This steady-state production ate of 4.65 L CH_4/L fermentor·d is comparable to he highest previously reported rate for aboratory-scale systems fermenting of animal residue (4.5 L CH_4/L·fermentor·d [3]) and is at least our times higher than any reported rate for pilot r full-scale systems fermenting animal residue. ossible reasons for the lower CH_4 production rates f other systems may be the inhibition caused by ntibiotics and growth stimulants used in commercial livestock enterprises, and the age and degree f contamination of the manures used in other ystems.

eef Feedlot Economic Assessment

t is generally assumed that beef cattle manure ill be the most economically feasible substrate or systems fermenting agricultural residues ecause large amounts of manure are generated in arge confinement feedlots. Based upon results rom the pilot-scale operation of the thermophilic, naerobic fermentation system, an economic ssessment was performed for beef feedlots.

ystem Description and Design: In the general flow cheme of the proposed fermentation system, manure rom the feedlot is being scraped into a mixing ank where it is diluted with water to the desired onsistency. The mixing tank is equipped with a ixer and mechanism to remove sand and grit from he tank. The slurry is then pumped to the [illegible]ntor, which is equipped with a mechanical [illegible] and heat exchangers for temperature control. Effluent from the fermentor is directly incorporated with other ration ingredients and fed to livestock. Thermophilic fermentation was selected because of the higher CH_4 production rates and presumed advantage of lower pathogen transmissibility. We selected a temperature of 55°C because of our experience with the pilot-scale system operated at that temperature.

Several biogas handling and use options were used in this assessment, depending on the ultimate use of the gas. Option A consisted of CO_2, H_2S and moisture removal by the water stripping, iron sponge and glycol absorption processes. The cleaned gas, predominantly CH_4, was then compressed to 860 kPa (125 psi) and stored in pressurized tanks with a capacity to store 1 days' CH_4 production. The CH_4 was then used as fuel to run an internal combustion engine/generator to produce electricity and the waste heat from the engine was used to heat the fermentor. Option B resembled Option A, except the CO_2 was not removed from the biogas. Option C resembled Option B, except the biogas was used directly after H_2S and moisture removal and was not compressed or stored. Option D resembled Option B, except the biogas was used as fuel for a boiler rather than used in an engine/generator. Option E resembled Option C, except the biogas was used as fuel for a boiler.

The following assumptions based upon our pilot-plant experience were used in sizing the fermentors: loading rate of 16 kg VS/m^3·d, VS:TS ratio of 0.85:1, 5-day retention time, and CH_4 production rate of 3.7 L CH_4/L·d. Note that the assumed methane production rate is 80% less than the maximum methane production rate achieved by the pilot-scale fermentor (4.65 L CH_4/L·d, Table 1). The methane concentration was assumed to be 50% of the biogas. The fermentor volume was calculated by multiplying the plant size (Mg TS/d) by the VS:TS ratio and dividing by the loading rate. The total energy production was calculated by multiplying the CH_4 production rate by the fermentor volume and by the energy content of CH_4 (37.3 MJ/m^3 or 1000 BTU/ft^3).

The fermentors were designed assuming that the tank height to diameter ratio was 1:1 up to a tank height of 10 m (total volume of 785 m^3). Tanks larger than 785 m^3 were designed with a maximum tank height of 10 m and sufficient diameter to accomodate the volume. The maximum tank diameter was assumed to be 80 m, resulting in a maximum tank volume of 5027 m^3. Systems requiring volumes greater than 5027 m^3 were designed with multiple tanks.

Table 2 summarizes the fermentor volume, energy requirements, and net energy production for the various options and fermentation plant sizes. Procedures used to calculate these energy requirements have been presented previously [4]. The net fermentor heating requirement was calculated assuming that 50% of the fermentor effluent heat was recovered and used to heat the influent slurry.

Table 2 shows the calculated electrical power requirements in descending order of size: gas

TABLE 2. ENERGY PRODUCTION AND REQUIREMENTS FOR VARIOUS FERMENTATION PLANT SIZES AND ENERGY USE OPTIONS.

	PLANT SIZE, Mg/d			
PARAMETER	3.5	35	175	350
Volume, m^3	186	1,860	9,300	18,600
Gross energy production, GJ/y	9,400	94,000	470,000	940,000
Net heating requirement[a], GJ/y	1,070	9,870	48,600	97,100
Electrical power requirement[b], kW				
Fermentor mixer[b]	2.20	7.62	37.3	74.7
Influent mixer[b]	2.75	4.04	11.4	24.3
Influent pump[c]	3.44	31.8	150	293
Effluent pump[c]	1.83	16.0	72.8	140
Gas Compressor[b]	6.35	63.5	317	637
Scrubber pump[b]	3.85	38.5	194	385
Electrical generator efficiency, %				
Option A	25.2	34.8	41.5	44.4
Option B & C	13.8	19.2	22.8	24.4
Net electrical production[d], GJ/y				
Option A	1,821	28,500	175,000	377,000
Option B	876	15,000	92,900	201,000
Option C	1,076	17,000	103,000	221,000
Net methane production[e], GJ/y				
Option D & E	7,972	80,800	470,000	810,000

[a] Assumes 50% recovery of effluent heat
[b] Assumes 24 hr/d operation
[c] Assumes 10 hr/d operation
[d] Gross electrical production minus electrical requirement
[e] Gross energy production minus net heating requirement

compressor, scrubber pump, influent pump, effluent pump, fermentor mixer and influent mixer. For plant sizes of 35 Mg TS/d and larger, the gas compressor, scrubber pump and influent pump account for about 40, 25 and 20%, respectively, of the total electrical energy requirement. The energy required to mechanically mix the fermentor is only about 5% of the total requirement.

Since the efficiency of the engine/generator to convert CH_4 into useful work increases as the engine size increases, the efficiency was estimated by the following equation based on the data of Perry and Chilton [5]:

Generator efficiency (%) = -13.05 + 4.18 ln (E) (1)

where E is the gross energy consumed by the engine (GJ/y) (i.e., the energy produced in the fermentor). Equation 1 is applicable only to Option A, since the efficiencies are for engines fueled with 100% CH_4. The engine/generator efficiencies for Options B and C were assumed to be 55% of that calculated from Equation 1, since the useful work output per energy input of a high compression engine receiving gas containing 50% CH_4 is 55% of an engine receiving 100% methane [6].

The net electrical production, shown in Table 2, was calculated by subtracting the electrical energy requirements from the electrical energy produced by the engine/generator. Option A produced the most net electrical energy since the increased engine/generator efficiency was sufficient to offset the increased energy required to compress and scrub the biogas. Option C produced more net energy than Option B, since the same amount of gross electrical energy was produced and Option B required more energy for compressing and storing the biogas.

The waste heat from the internal combustion engine was assumed to be sufficient to satisfy the heating requirements of the fermentor. This assumption is justified since the net heating requirement is [illegible] about 10% of the total energy production, and [illegible]ween 20 to 30% of the gross energy consumed by the ngine can be recovered from the jacket cooling ysi [illegible] Considering that an additional 26 to 30% f [illegible] heat can be recovered from the engine khaust gases, Options A, B and C can be a source f low temperature (75 to 85°C) process water, as ell as electricity.

he net CH_4 production shown in Table 2 was calculated by subtracting the net heating requirement rom the gross energy production. There was no ifference in net CH_4 production between Options D nd E.

apital Costs: Table 3 shows the installed equip-ent costs for major components of a 35 Mg/d fer-entation system. The total installed equipment ost would vary depending upon the options elected. For example, the most capital intensive ystem (Option A) cost $680,000. The least capital ntensive systems (Options C & E) each cost 380,000.

otal capital costs were estimated using engi-eering and inspection fees, contingency, escala-ion and start-up costs of 20, 10, 12 and 10% of nstalled capital costs, respectively. To estimate he total capital costs for different size plants, he scale-up factor of fermentor volume to the 0.7 ower, as shown in Figure 1, was used. Since the ermentor volume of the 35 Mg/d plant is 1860 m^3 Table 3), the total capital costs (TCC) for the arious options can be estimated by:

$$TCC_A = 5320\ (V)^{0.7} \text{ for Option A} \qquad (2)$$

$$TCC_{B,D} = 4140\ (V)^{0.7} \text{ for Options B and D} \qquad (3)$$

$$TCC_{C,E} = 2970\ (V)^{0.7} \text{ for Options C and E} \qquad (4)$$

here:

TCC = total capital costs, dollars
V = fermentor volume, m^3

Table 4 shows the capital costs for the various options and fermentation sizes of 3.5, 35, 175 and 350 Mg TS/day.

Annual Costs: Annual costs, consisting of labor, fixed and utility costs were estimated for the various plant sizes (Table 4). Salaries for the plant operators were assumed to range from $20,000/yr for the 3.5 Mg/d plant up to $80,000/year for the 350 Mg/d plant. Fringe benefits, operational supplies and miscellaneous expenses were estimated to be 15, 10 and 20% of salaries, respectively. Labor costs were assumed to be the same for all options at the same plant size.

Fixed costs were calculated assuming an interest rate of 11% on the total capital costs, and a 20-year straight-line depreciation of the total capital costs. Taxes, insurance, and repair and maintenance were estimated to be 3, 1.5 and 3% of the installed equipment costs, respectively.

Utility costs were calculated based upon the energy requirements in excess of that produced (i.e., electricity and/or heat). Utility rates were assumed to be $12.5/GJ (4.5¢/kWh) for electricity, and $0.10/$m^3$ for make-up water. The only utility cost charged to Options A, B and C was for make-up water since the engine/generator produced more electricity and waste heat than that needed for the fermentation system. The utility costs shown for Options D and E reflect the make-up water and electricity costs.

Table 4 shows the total annual costs for the various options. Options A and D, the higher cost systems, have almost identical total annual costs and constitute the higher cost systems, whereas Options C and E, the lower cost systems, have similar total annual costs. Option B has total annual costs between the two extremes.

TABLE 3. INSTALLED EQUIPMENT COSTS FOR MAJOR COMPONENTS OF A 35 Mg/d THERMOPHILIC, ANAEROBIC FERMENTATION SYSTEM[a]

	COST (in $1000) FOR EACH OPTION				
COMPONENT	A	B	C	D	E
Premixing and degritting	50	50	50	50	50
Pumps	30	30	30	30	30
Fermentor w/mixer	250	250	250	250	250
Heat exchanger	15	15	15	15	15
Piping	10	10	10	10	10
H_2S and H_2O removal	---	50	50	50	50
CO_2, H_2S and H_2O removal	200	---	---	---	---
Gas compression and storage	100	100	---	100	---
Boiler	---	---	---	25	25
Engine/generator w/heat exchangers	25	25	25	---	---
TOTAL INSTALLED EQUIPMENT COST	680	530	380	530	380

[a] Temperature = 55°C, retention time = 5 days, loading rate = 16 kg/m^3·d, volume = 1860 m^3

TABLE 4. COSTS FOR PRODUCING METHANE AND ELECTRICITY FOR VARIOUS FERMENTATION PLANT SIZES

PARAMETER	PLANT SIZE, Mg/d			
	3.5	35	175	350
Capital costs, $1000				
Option A	206	1,034	3,927	7,850
Options B & D	161	805	3,056	6,113
Options C & E	115	577	2,192	4,385
Labor costs, $1000/y	33.9	65.6	104	127
Fixed costs, $1000/y				
Option A	43.3	217	825	1,650
Options B & D	33.7	169	642	1,284
Options C & E	24.2	121	460	921
Utility costs, $1000/y				
Options A, B & C	0.51	5.1	25	51
Option D	5.83	42.5	206	411
Option E	3.33	17.5	81	161
Total annual costs, $1000/y				
Option A	77.7	288	954	1,827
Option B	68.1	240	771	1,461
Option C	58.6	192	590	1,099
Option D	73.4	277	952	1,822
Option E	61.4	204	646	1,209
Electrical production costs, $/GJ				
Option A	42.7	10.1	5.46	4.85
Option A w/refeed credit	20.3	-4.19	-6.19	-5.96
Option B	77.8	16.0	8.30	7.28
Option B w/refeed credit	31.2	-11.2	-13.6	-13.0
Option C	54.4	11.3	5.73	4.97
Option C w/refeed credit	16.6	-12.7	-14.1	-13.5
Methane production costs, $/GJ				
Option D	9.21	3.43	2.35	2.25
Option D w/refeed credit	4.10	-1.61	-2.68	-2.78
Option E	7.70	2.53	1.59	1.49
Option E w/ refeed credit	2.59	-2.51	-3.43	-3.53

Energy Production Costs: The electrical and CH_4 production costs are also shown in Table 4. These costs were calculated by dividing the total annual costs for each option by the annual net energy (electricity or methane) produced by each option (shown in Table 2). Production costs with refeed credit were calculated by subtracting the refeed credit from the total annual costs and dividing the difference by the annual net energy production. The refeed credit was calculated by assuming that the effluent TS concentration was 5%, the refeed value was $60/Mg TS, and all of the effluent solids could be recovered and used.

The effects of fermentation plant size on the electrical energy production costs for Options A, B and C are shown in Figure 2. Even though Option A is the most capital intensive system, the increased efficiency of electricity generation by scrubbing CO_2 from the biogas makes Option A the most cost effective system. The slight increase in cost at about 95 Mg/d was caused by limiting the size of each fermentor to 5027 m^3. The fermentation plant size at which the electrical production costs equal the current electricity rate of $12.50/GJ would be 24, 54 and 29 Mg TS/d for Options A, B and C, respectively.

The effects of fermentation plant size on the CH_4 production costs for Options D and E are shown in Figure 3. The plant size at which CH_4 production costs equal the current natural gas rate of $2.50/GJ would be 91 and 36 Mg/d for Options D and E, respectively.

Based on these results, the best energy production

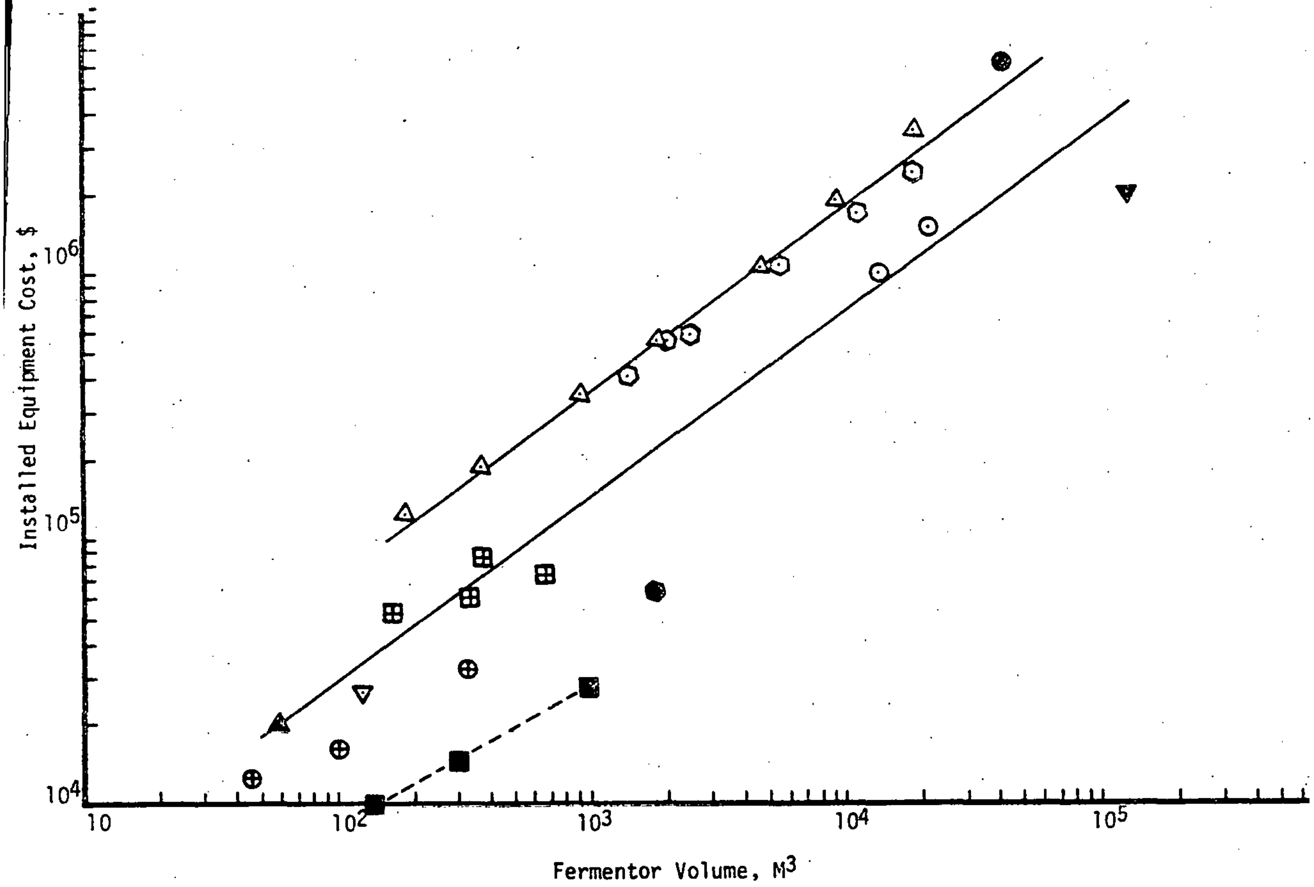

Figure 1. Effect of Fermentor Volume on Installed Equipment Cost (⊙, [10]; ●, [11]; ⊞, [12]; ▽, [13]; ▼, [14]; △, [15]; ⊕, [16]; ■, Plug-flow, [16]; ▲, [17]; ⬢, [18]; ⊙, [19]).

options, in descending order, are Options A, C, E, B and D. However, when the refeed credit is applied to this analysis, the plant size at which energy production costs equal current energy prices are 5, 6, 4, 5 and 3.5 Mg/d for Options A, B, C, D and E, respectively. The reason for the small difference between the various options with the refeed credit is due to energy production costs decreasing at a faster rate for those options that have lower total annual costs.

Implications of this Assessment: This assessment has illustrated the significant impact that the refeed credit has on the overall economics of the fermentation systems. We assumed a feed value of \$60/Mg TS of the fermentor effluent, which is about one-half to one-third the value estimated by Smith and Wheeler [6] for fresh beef cattle manure. Unpublished data from our Center shows that the fermentor effluent contains about twice as much amino acids, on a dry matter basis, as fresh beef [illegible]ure. Thus, the feed value attributed to the effluent seems to be conservative. However, enough feeding trials using the fermentor effluent have not been completed to fully assess its economic value. Research in this area is currently being conducted at the Roman L. Hruska U.S. Meat Animal Research Center and at other locations.

Lipper et al. [8] reported that the utility requirements for energy intensive, commercial beef feedlots in Kansas were 2.2 GJ/head·y for natural gas and 0.32 GJ/head·y for electricity. Assuming that in a confinement beef feedlot each animal produces 3.5 kg/d of recoverable TS, then a 1,000-head feedlot would need a 3.5 Mg/d plant to convert the manure to CH_4. Thus, a fermentation plant for a 1,000-head feedlot would produce 1.8, 0.9 and 1.1 GJ/head·y of electricity for Options A, B and C, respectively; and a 100,000-head feedlot plant would produce 3.8, 2.0 and 2.2 GJ/head·y of electricity, respectively. This would be between 3 to 10 times the electrical requirements of these feedlots. Likewise, Options D and E would produce

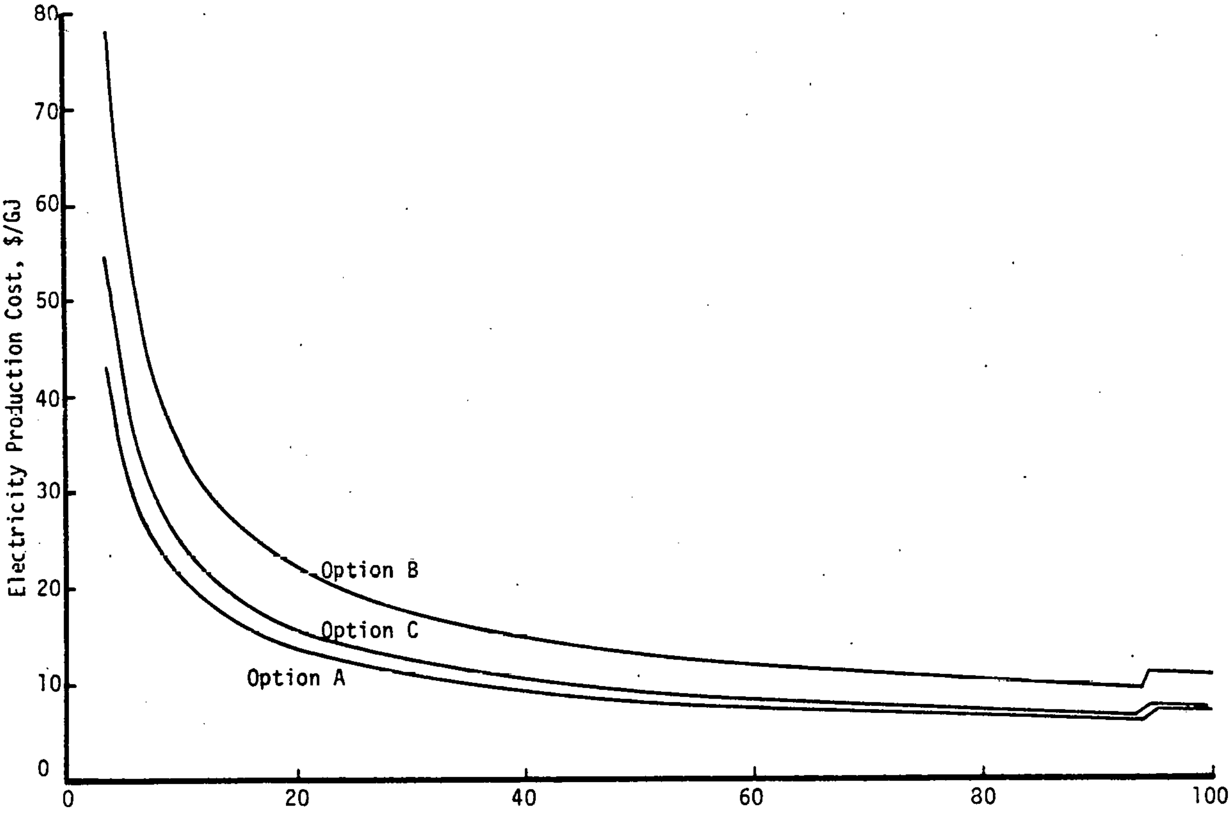

Figure 2. Effect of Plant Size on Electricity Production Cost.

about 8 GJ/head·y of substitute natural gas, which is about 4 times the feedlot requirement.

This assessment of the economic feasibility of anaerobic fermentation of beef cattle manure has shown that CH_4 can be economically generated for moderate confinement feedlot sizes (1,000 to 1,800 head) when the fermentor effluent is used as a protein supplement in feeds. Also, more energy is produced, either in the form of heat or electricity, than can be used by the feedlot enterprise. Strategies to utilize this excess energy must be developed. Conscientious effort to adjust energy demand to production must be exercised at the enterprise level, and there must be the opportunity to sell surplus energy. An analysis of the economic impact of CH_4 generation on a 65-cow dairy farm in the Northeast United States [9] showed that allowing surplus energy to be sold to electrical utilities would improve the national economy. The utility would incur revenue losses; however, this revenue loss results in a commensurate decrease in utility capacity and fuel costs. This conflict between efficiency at the national and utility level would have to be resolved through regulation at the interface between the farmer and the utility.

FUTURE PLANS

We plan to continue operation of the pilot-scale fermentor to develop improved design and scale-up criteria for fermentors and to evaluate methods to recover and utilize the protein fraction of the fermentor effluent. Specific plans are to evaluate different fermentation temperatures, the effect of broad-spectrum antibiotics on fermentation, and combining crop residues with manure as fermentation substrate. Promising polyeletrolytes identified in laboratory studies will be evaluated for its protein capture efficiency and possible toxicity when fed to livestock.

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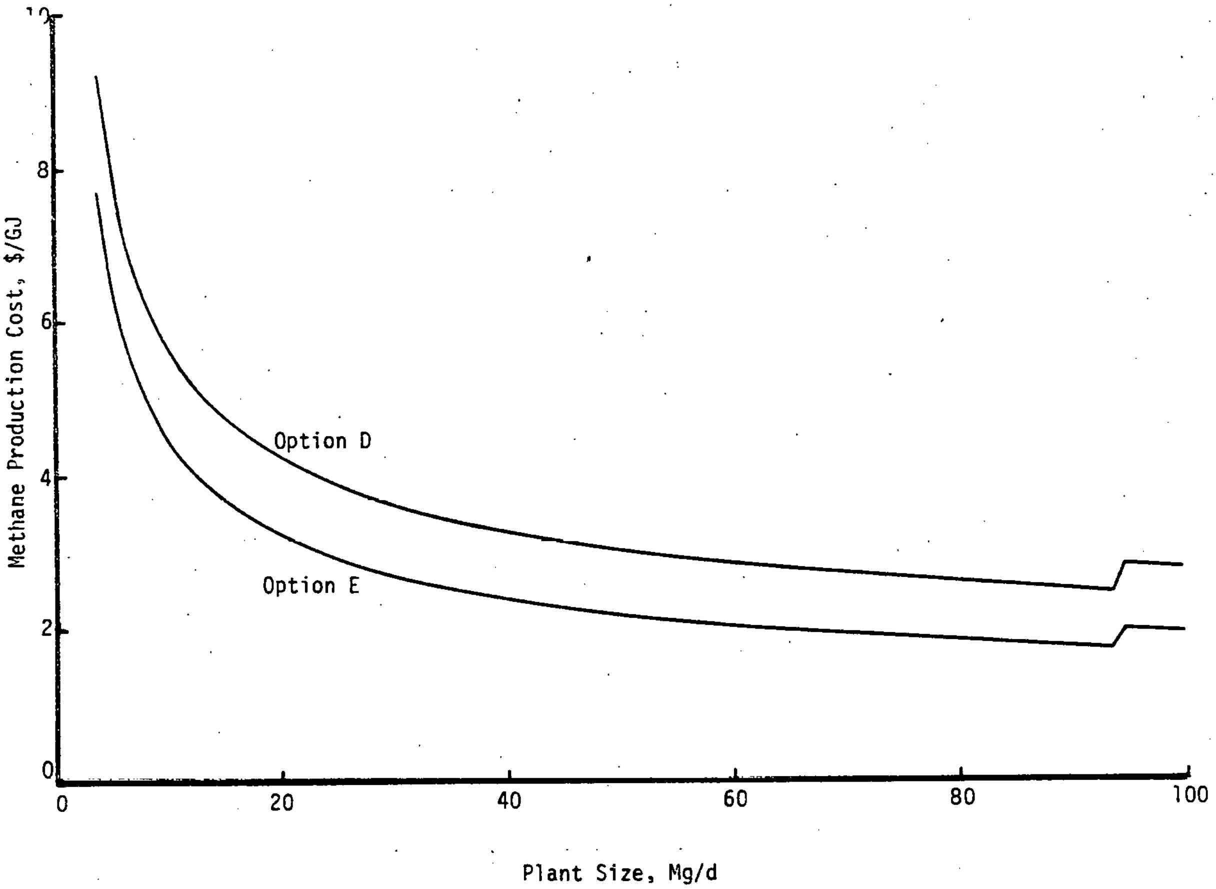

Figure 3. Effect of Plant Size on Methane Production Cost

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THE OPERATION OF A 50,000 GALLON ANAEROBIC DIGESTER AT THE MONROE STATE DAIRY FARM

Elizabeth R. Coppinger
Ecotope Group
2332 East Madison
Seattle, WA 98112

ABSTRACT

A full scale anaerobic digester on the State Dairy Farm has been operated and monitored for the past 22 months under funding from the U.S. Department of Energy, Fuels from Biomass Systems Branch. During the period of its operation, a number of variables have been changed and the impact of those changes on digester performance are described.

The percent of the slurry loaded has been raised from 4% to 12%. In-tank mixing is by gas recirculation and has been decreased progressively from 24 hours per day to none at all. The only internal mixing at present is by convection currents caused as a byproduct of the digester heating system.

Heating requirements have been reduced by tank insulation and reductions in make-up water added. A prototype tube-in-shell influent/effluent heat exchanger has been installed and tested. Based on the operational experience gained from the Monroe digester, design recommendations are presented.

INTRODUCTION

The widespread use of anaerobic digestion on farms requires the demonstration of reliable systems that can be integrated into present farming operations and deliver energy and other benefits at a cost competitive with other energy sources. The economics of systems will improve as experience with plant operation is used to refine designs and reduce capital and operator/maintenance costs of future systems.

A full scale anaerobic digester on a 200-head dairy farm near Seattle has been operated and monitored since July 1977 under funding from the U.S. Department of Energy Fuels from Biomass Systems Branch, contract #EG-77-C-06-1016. Work has focued on improving gas production, reducing gas and electrical consumption, minimizing operator time, reducing maintenance time, and eliminating unnecessary equipment. This paper describes the field operation and experiences gained at the Monroe digester and how that information has impact on the economic feasibility of digesters for on-farm use.

System Description

The Monroe system represents a transfer of state-of-the-art municipal sewage treatment plant technology to an agricultural situation. An emphasis was placed on use of easily obtainable off-the-shelf components under the assumption that using equipment already proven and accepted in the agricultural sector would speed the wide-spread use of digestion technology. (See Fig. 1)

The reactors are two 50,000 gallon A.O. Smith Slurrystore$_{tm}$ tanks fitted with Harvestore$_{tm}$ silo roofs. Certain modifications were made to the tanks for use as experimental digesters. Two thief holes were installed on the digester roof for sampling contents from the tank interior and eight side-mounted sampling ports were installed at three levels around the perimeter of the tank to provide a variety of sampling locations. The most significant tank modification was insulation of all exposed tank surfaces. The interior roof of the tank was sprayed with 3½" of polyurethane foam (R-21). Exterior walls were covered with 4" of Dow Styrofoam SM$_{tm}$ and surfaced with corrugated galvanized iron roofing sheets.

The digester system was designed to integrate with the farm's manure handling system. Manure is scraped once a day from a covered loafing shed and contains some sawdust used for bedding. The manure is received at 14%-16% total solids (TS) and diluted in a mixing tank before being loaded into the digester. This slurry is mixed using a centrifugal chopper pump. Either the chopper pump or a variable speed progressive cavity pump is used to load the slurry into the digester. As manure is loaded into the bottom of the digester, the liquid level rises in the tank and digested manure is displaced, passing through an overflow pipe at the top of the digester. The effluent flows into a storage lagoon and is eventually applied to the fields with spray guns to fertilize crops grown as cattle food.

The system operates at 95°F. Hot water from a boiler or from the coolant system of an internal combustion engine is pumped into a cylindrical heat exchanger inside the digester. The heat exchanger also doubles as a draft tube for use in conjunction with the digester's gas recirculation mixing system, originally designed to continuously mix digester contents.

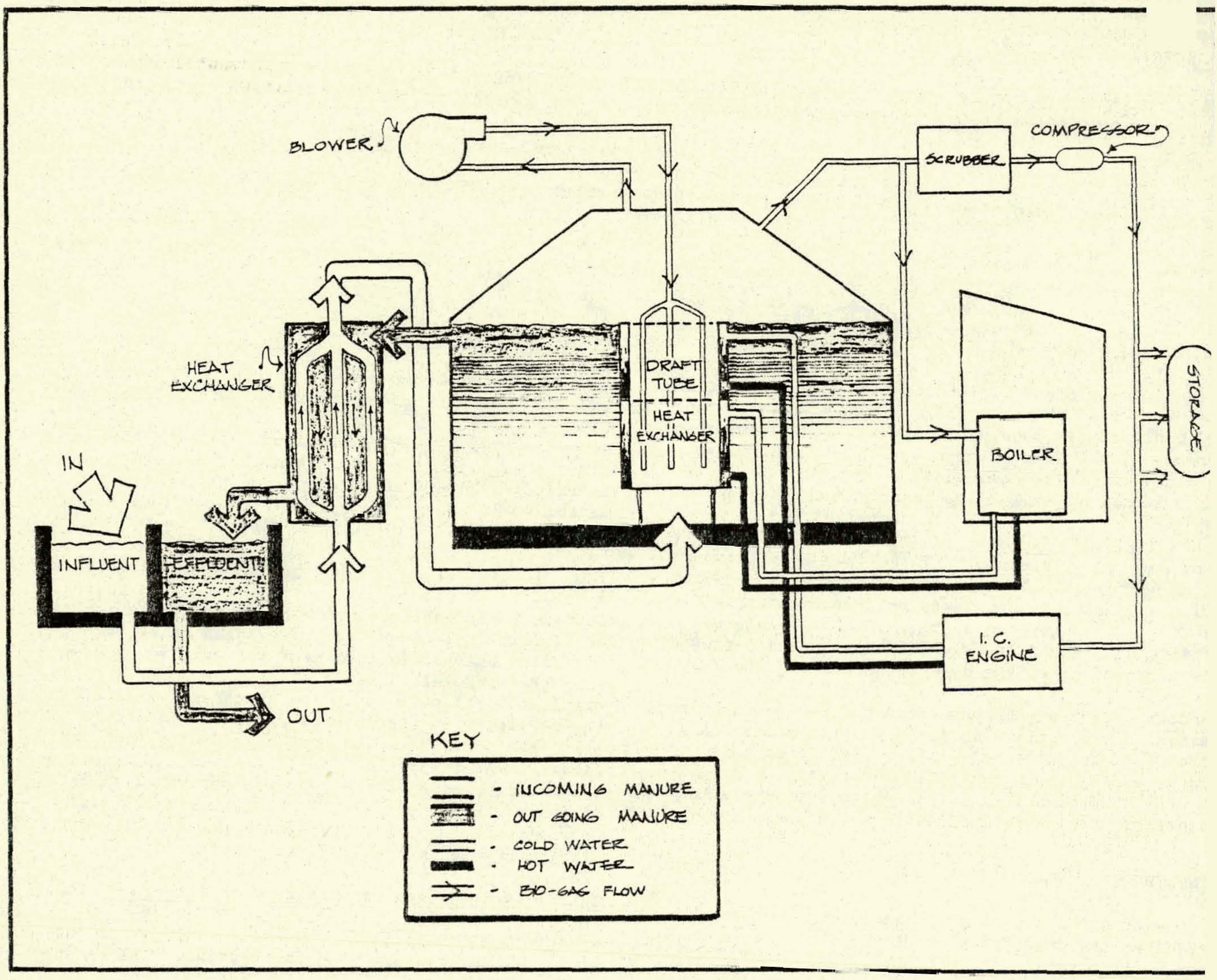


Fig. 1. Schematic of Monroe Anaerobic Digester System

The gas handling components of the Monroe system were modified little from standard sewage treatment gas handling. As a result, this proved to be one of the most expensive aspects of the system. According to the original design gas was to be: (1) burned directly in the boiler for heating the digester; (2) scrubbed and transported to the farm's creamery to produce hot water; or (3) scrubbed and stored for burning in an internal combustion engine to produce electricity in emergency situations. The gas handling system was automated using pressure switches. Only the water traps are manually maintained.

The digester was outfitted with monitoring equipment to assess system performance and energy production. Gas meters were installed to measure gas production and consumption of the boiler, the IC engine, and the flare. Electric meters were installed on pumps, the mixer, and the IC engine. Temperature probes were installed at a variety of locations in the tank to monitor material and heat movement within the tank.

RESEARCH OBJECTIVES

The system has been operated, monitored and evaluated for two years. We have attempted in that time to continually stress the digester and improve its operating characteristics. We have consistently observed that the biological processes associated with the system are very stable and able to withstand much greater stress than originally predicted.

The major research areas in the work at the Monroe facility this year have been:

(1 assess the impact of an increase in the pe total solids of influent loaded into the digester on operator time and net energy production;

(2) To assess the impact of reduction and elimination of tank mixing on electrical consumption, capital cost and gas production;

(3) To test an influent/effluent heat exchanger to decrease gas consumption for digester heating;

(4) To produce electricity with an engine/generator under a variety of load conditions to develop background data on the feasibility of on-farm electrical generation.

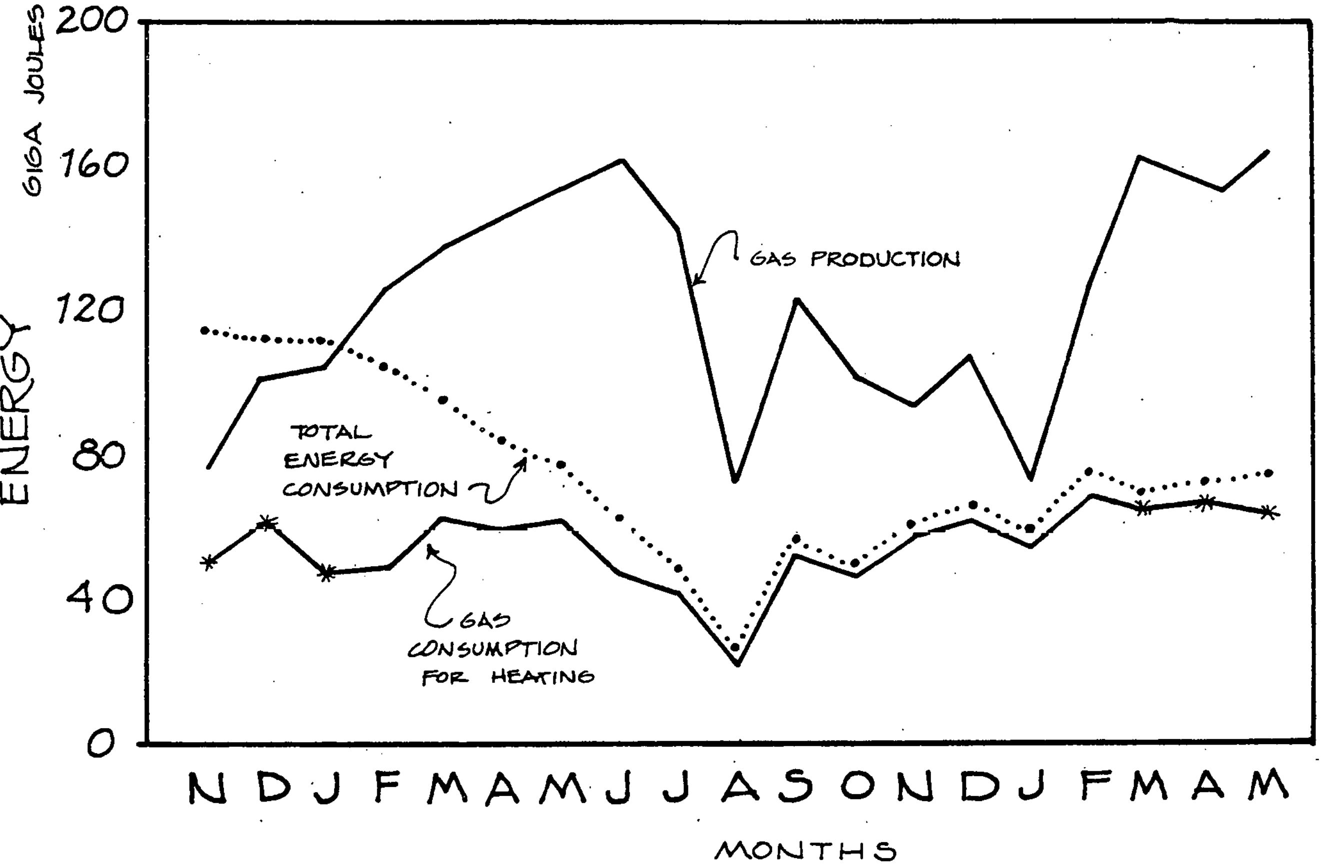
RESEARCH FINDINGS

Percent Total Solids and Plant Operation

During its two years of operation, the percent total solids of the manure loaded has been increased from 4% to 12%. Varying this parameter has had a large impact on many aspects of the system.

The digester tanks were sized based on an assumption that manure could not be loaded at higher than 8% TS and that a retention time of 20 days would be optimum. Based on sewage treatment experience, it was assumed that these limits were necessary for the health of the digester. Once loading began, the impressive stability of the biological parameters led us to increase the loading rate and to increase the % TS of the manure loaded. As a result of these changes, we were able to load all of the manure received into one digester tank. Reducing the needed digester volume by one-half has a great impact on the capital cost of the system since the digester tanks were the most expensive single item of the system.

Gas production has also been positively impacted by increasing the % TS of the influent. The same amount of manure can be loaded with a longer retention time, thereby increasing the gas production. Loading at 12% TS in April 1979 resulted in a production rate of 3.79 ft^3 of biogas per pound volatile solids (VS) added per ft^3 reactor volume, versus 3.52 ft^3/lb VS/ft^3-R added in February 1979, a month that had a comparable loading rate at 10%



* CORRECTED FOR USE OF ENGINE COOLING WATER FOR DIGESTER HEATING
NOTE: ELECTRICAL ENERGY IS CALCULATED @ 11 MEGA JOULES/KWH TO ACCOUNT FOR CONVERSION EFFICIENCY OF CENTRAL STATION GENERATION

Fig. 2. Net Energy for Monroe Digester

solids. (See Table 1) Reduction of water added to the slurry also decreases the system's heat demand since the water must be heated from 50°F to 95°F even though water does not increase gas production.

A significant impact on plant operation has also been made by varying the % TS loaded. When manure was loaded at 4% - 8% TS, lighter materials in the effluent would separate and clog elbows and pipes of the effluent lines. A thick scum blanket would also form in the effluent holding tank, requiring a great deal of agitation to prepare for pumping. These clogging problems could take as long as one hour to solve and were a large source of frustration to operators. After the loading rate was raised to 10% TS and the % TS in the tank rose to greater than 7.5% TS, these clogging problems disappeared. The viscosity of the slurry changed so that the materials stayed in suspension. A number of times in the past year, the % TS in the tank has dropped below 7.5% due to water leaks, and each drop has been accompanied by clogging problems.

Increased electrical demand and operator time associated with influent mixing are the major negative impacts of increasing the % TS loaded. The centrifugal chopper pump has been used to mix the influent by recirculation. When manure was loaded at 8% TS or less, the pump could easily and quickly mix the slurry. As the slurry was increased to 10% and eventually to 12%, the velocity of the manure stream from the bypass pipe decreased and influent mixing time increased significantly. At 8% TS, mixing could be completed in less than one hour with little operator assistance. At 10% TS, mixing time was increased to over an hour and the operator was required to push large clumps of manure into the mixing stream.

When the influent was raied to 12% TS, it became very difficult to thoroughly mix the influent with the present system and it now requires over three hours of pump operation and one-half hour of operator time. (See Fig. 2)

A mechanical system may be able to mix the slurry, but this recycle system cannot. There are tradeoffs between the energy gains that result from increasing the slurry solids loaded and the electrical energy demands of an influent mixing system. The characteristics of the slurry differ so significantly as the % TS are raised that an influent mixing system should be designed for the specific requirements of high solids loading to provide the most cost effective and energy efficient system possible.

Designing a digester system to handle a high % TS would result in savings in capital costs due to the reduction of digester tank volume needed. It will also improve the net gas production of the system by decreasing influent heat demand. However, it is essential that the manure handling system be specifically designed for a thick substrate. Otherwise, the economic benefits gained can be lost to increased electrical consumption and increased operator time.

Table 1. MONROE DIGESTER PERFORMANCE, OCTOBER THROUGH MAY 1979

	loading rate/day lb/ft^3-reactor	ft^3 biogas/ lb VS added	% VS destroye
1977			
Oct	.19	2.72	25.5
Nov	.23	2.64	13
Dec	.29	2.83	18
1978			
Jan	.30	2.75	-
Feb	.41	2.74	26.6
Mar	.37	2.96	19.4
Apr	.37	3.25	22.5
May	.34	3.14	31.8
Jne	.39	3.46	23.2
Jly	.29	3.86	21.6
Aug	.16	2.78	22.3
Spt	.31	3.33	26.4
Oct	.20	4.03	29.4
Nov	.21	3.87	27.5
Dec	.23	3.73	31.5
1979			
Jan	.17	3.52	28.7
Feb	.33	3.52	28.3
Mar	.38	3.47	37.1
Apr	.34	3.79	36.7
May	.37	4.0	

Table 2. % TS OF DIGESTER CONTENTS IN MIXING STUDIE

	% TS						
	Constant Mixing		50% Mixing		33% Mix	17% Mixing	
sample from:	Ja21	Fe18	Mr4	Ap4	Ap29	Je1	Je10
PERIMETER	7.4	8.2	8.2	8.0	8.1	7.3	7.87
top	7.4	8.2	8.2	8.0	8.1	7.3	7.87
	7.5	8.4	8.2	8.0	8.2	7.6	7.82
middle	7.2	8.4	7.8	7.9	7.9	7.5	7.87
	7.2	8.3	8.2	8.1	8.1	7.3	8.10
	7.1	8.3	-	7.9	8.2	7.3	8.17
bottom	7.4	8.3	8.3	8.2	8.7	7.5	7.87
	7.5	8.3	8.5	7.9	8.0	7.5	7.90
INTERIOR							
top	8.5	9.0	8.4	8.8	8.4	8.7	8.6
bottom	8.0	8.5	8.1	8.6	8.4	7.9	8.2

Mixing Digester Contents

Based on the municipal sewage treatment problems with scum formation, the Monroe digester was designed to be continuously mixed. A Roots-type recirculation blower was used in conjunction with an internal draft tube that doubled as the system's heat exchanger. During the first five months of operation in 1977, the blower was run continuously. The electrical demand of the blower was 180 kWh/day, representing 90% of the total electric demand of the system. The blower also required costly rep[illegible] during the time of its operation as well as routine oil changes each week.

nt ttent mixing was investigated in order to ed electrical consumption and equipment wear. aseline mixing studies were performed to determine f solids stratification occurred in the tanks. o samples differed more than .5% TS. (See Table 2) ixing was decreased to 15 minutes on and 15 minutes ff. No increase in solids separation resulted. ixing was decreased to 10 minutes on and 20 minutes ff, then to 10 minutes on and 50 minutes off with o significant stratification and no negative impact n gas production. Stratification tests and operaional experience have shown that solids separation s dependent primarily on the % TS of the slurry.

The effects of convective currents and gas movement on mixing tank contents were investigated. Temperature probes were installed in a variety of locations throughout the tank to monitor the movement of cold manure from loading and the movement of tank contents when warmed by the internal heat exchanger. Blower use was again reduced to loading periods only. Under these conditions, a uniform temperature drop was seen throughout the tank, indicating that the blower effectively disperses the influent during loading. Use of the blower only when loading was continued from May 1978 through early March 1979 with no negative gas production or operational problems.

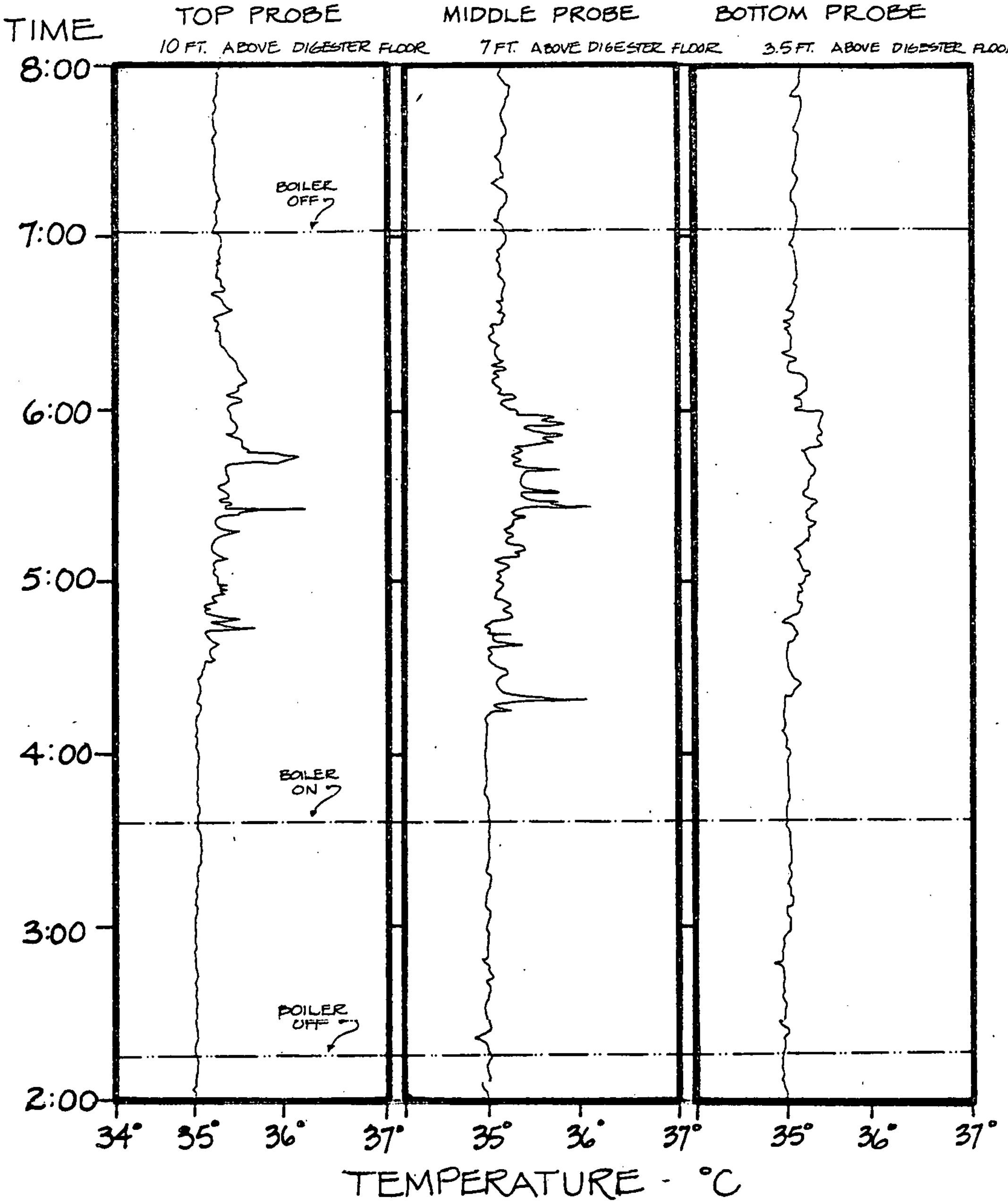


Fig. 3. Temperature Variations Inside the Digester
Due to Boiler Operation

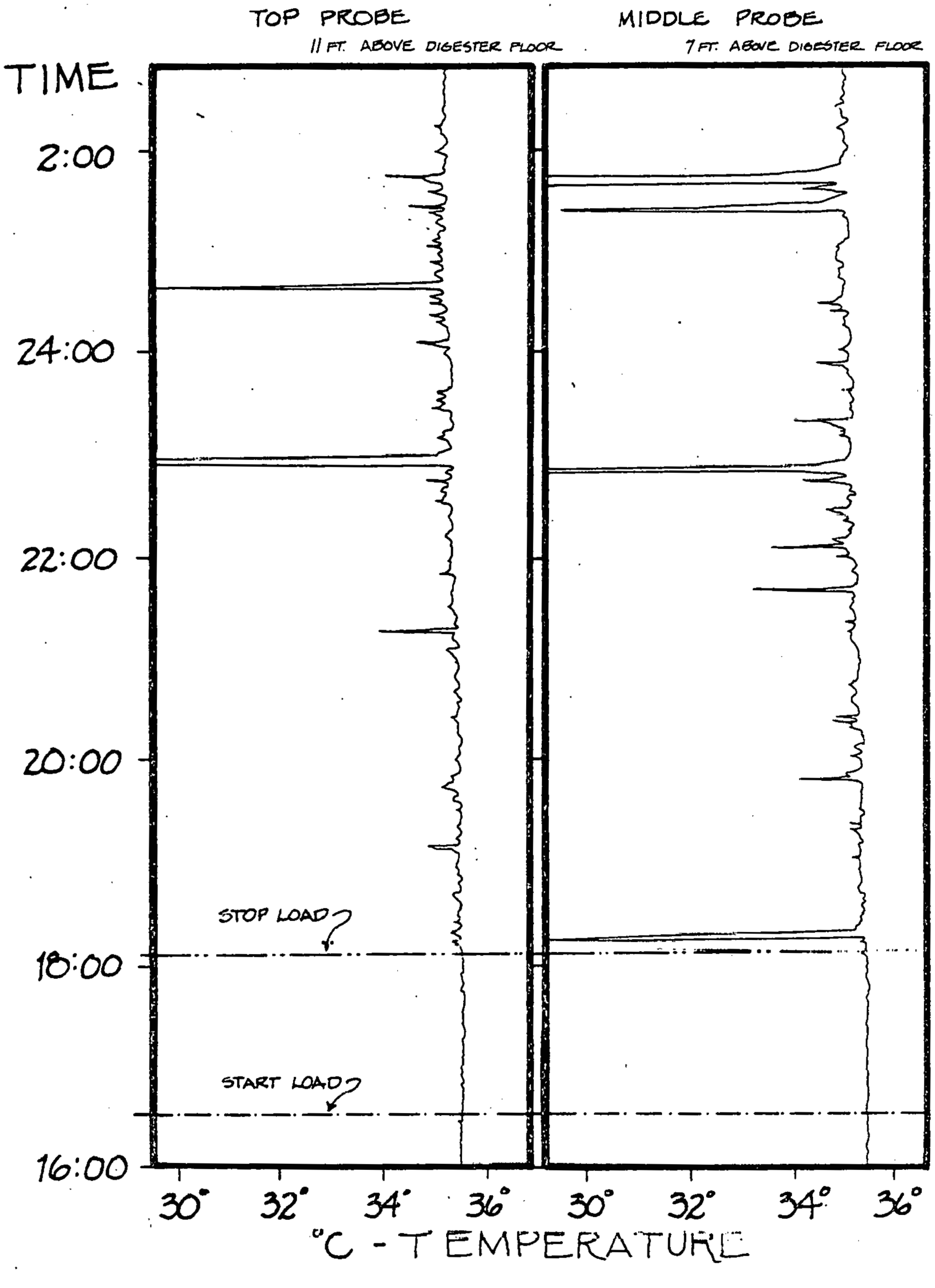


Fig. 4 Temperature Variations Inside the Digester Due to Loading

Mixing was stopped completely on March 6, 1979. Temperature probes in the digester showed that mixing still occurs in the tank without mechanical agitation. Fig. 3 shows movement that is due to convective currents established when the boiler is running. This example is a period from 2 a.m. to 10 a.m. on May 21, 1979. At 2 a.m., the temperature throughout the digester was uniform. It has stabilized after the previous day's loading and the boiler had remained off for most of the night. When the boil[illegible] turned on, agitation could be seen at the three probe points. The spikes on the chart represent an--- that has been heated above the temperature f rest of the tank moving past the stationary rn ... The decrease in the size of the spikes rom the middle to the upper probe indicates that he manure is losing heat as it rises.

he small temperature difference noted by the lower robe may indicate that the manure passing it is eplacing the manure that has been warmed by the nternal heat exchanger. This movement continued hile the boiler was on and decreased after the oiler shut off. Additional agitation may be occuring due to gas movement through the slurry, but e have not yet been able to perfect a flow probe hat can be inserted into our sampling ports and re therefore unable to detect any mixing that ccurs isothermally.

he impact of this natural mixing on freshly loaded anure can be seen in Fig. 4. The temperature of he tank contents was constant before loading. fter loading was completed, sharp spikes of low emperature indicate that cold manure was passing he temperature probes. These probes were located t 7' and 11' above the bottom of the tank. This ixing could be seen for nine hours after loading. ndoubtedly, mixing occurred after this time, but nce the influent temperature was raised to the ame as the tank contents, it could not be detected y the probes.

he use of convective currents to replace mechanical ixing systems can have a great impact on the conomic feasibility of digestion. Not only would t reduce electrical consumption and maintenance ime, but it could also reduce the capital cost of system by as much as 8% and eliminate the ulnerability inherent in having a necessary piece f equipment inside the digester tank.

o maximize convective mixing, more information is eeded to design effective natural digester mixing ystems. The impact of such parameters as % TS, nfluent temperature, tank size, and plumbing conigurations on natural mixing need to be evaluated o determine if mechanical mixing systems can be liminated.

nfluent/Effluent Heat Exchange

ince heating influent represents 80% - 90% of the eating demands for an insulated digester, a great eal of attention has been given in this research ork to the question of recovering heat from effluent for pre-heating influent. A vertical tube-inhell counterflow heat exchanger was designed and nstalled. (See Fig. 5). The original design onsisted of 25 segmented aluminum tubes joined by ubber connectors inside a sheet metal shell. The ystem was operated in February 1976 using a 3% lurry. Severe clogging in the tubes was encountered during loading. In addition, failure of a umber of the rubber connectors resulted in a hort circuiting between the influent and effluent. he heat exchanger was disassembled and the segented aluminum tubes were replaced with straight wall PVC pipe. This reduced the expected transfer, but eliminated the problem of short ircuiting.

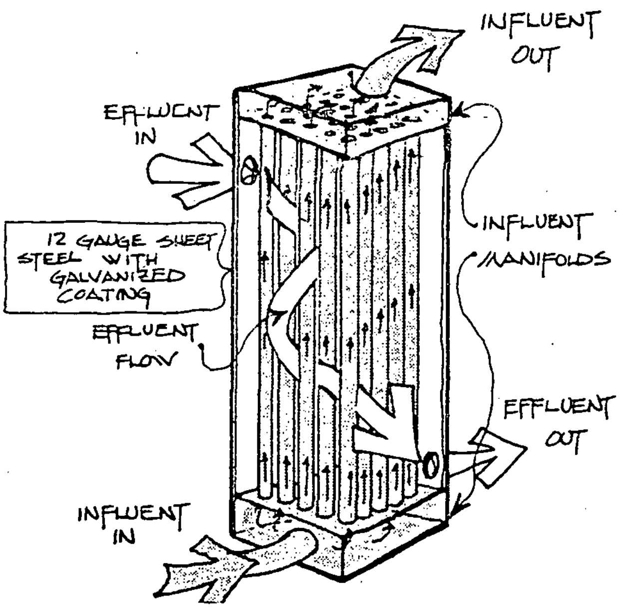


Fig. 5. Influent/Effluent Counterflow Heat Exchanger

Loading through the heat exchanger was attempted again in October 1977. At this time, an 8% slurry was being loaded. Neither of the system's pumps was able to load through the heat exchanger. Wood chips in the slurry prevented the checkvalves on the diaphragm pump from seating correctly and the pump could not be used. The centrifugal pump, with its high flow rate and low pressure was not suitable for loading through the heat exchanger.

A progressive cavity pump was obtained that was able to load high percent solids through the heat exchanger. Influent was loaded at 10% - 12% total solids and the loading rate was varied from 30 gallons per minute to 17 gal/min. The maximum influent temperature rise achieved at any flow rate was 3°F and the difference in performance at various flow rates was not significant. At the percent TS tested a great deal of channeling occurred, as evidenced by distinct hot and cold spots that could be felt on the wall of the heat exchanger.

Manure at these thicknesses has such poor thermal transfer properties that it essentially acted as an insulator in areas where a constant flow was not maintained. Heat exchangers must be designed to take into account the flow characteristics of the material to ensure proper flow and eliminate stagnation at the heat exchanger surface. We are now preparing to test heat exchanger sections of different configurations to evaluate how to optimize influent/effluent heat exchange.

While manure loaded at high % TS has a positive effect on system net energy and operation performance, it complicates the already difficult problem of influent/effluent heat exchange. Additional work needs to be done to design and test influent-to-effluent heat exchange and effluent-to-mixing water heat exchange, as well as testing the feasibility of using effluent as a source of low grade heat.

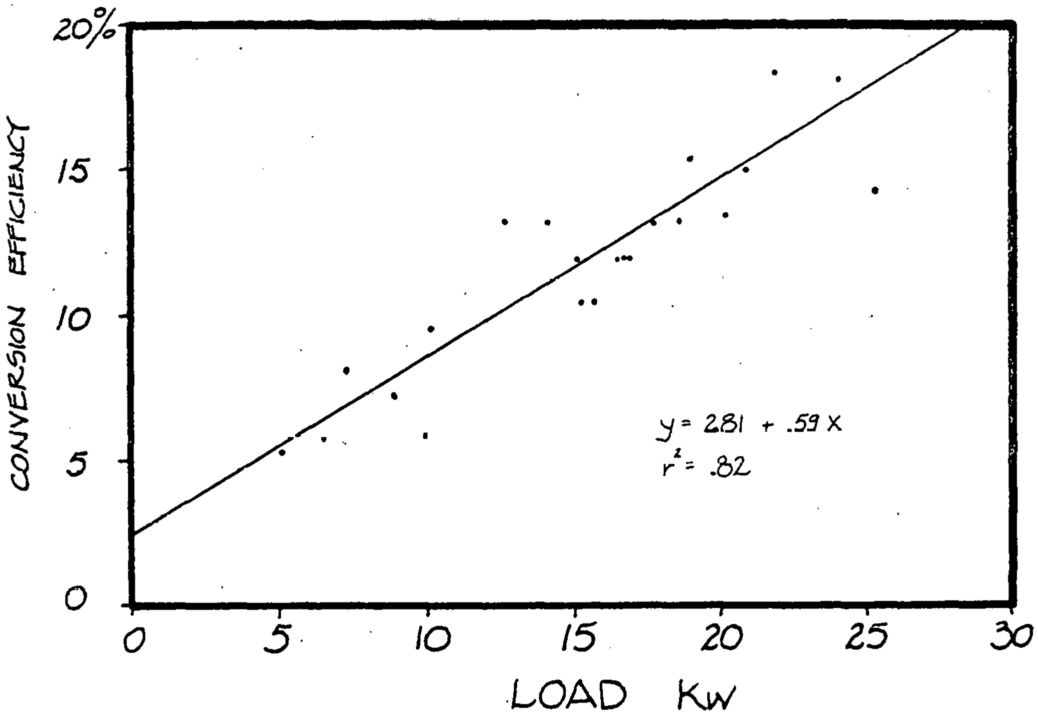


Fig. 6. Conversion Efficiency for Electrical Generation

Engine/Generator

An internal combustion engine with a 40 kW (peak) generator was installed as a part of the original demonstration project. The purpose of this installation was to provide emergency back-up electricity for the creamery and milking operations. Consequently, it was sized to meet peak electrical needs of the creamery and not to be compatible with daily gas production rates. The IC engine was operated December 1977 through January 1978 and March through May 1979. A resistance electrical load was hooked up to the engine that could be varied from 40 kW to 6 kW.

The engine/generator set is rated at 40 kW and 23% efficiency, based on its performance running under full load conditions with propane. The system must be derated for lower BTU fuels. As a consequence, the engine would not run at a load greater than 25 kW. This represents a capacity loss of 37%. Fig. 6 shows the electrical conversion efficiency under various load conditions. The efficiency varied linearly over the 6 kW to 25 kW range tested. Based on our test results, running the engine/generator at the full load of 25 kW would result in an electrical conversion efficiency of 18%.

When the engine is run, its coolant water is used to heat the digester contents. The engine coolant water was used during the two coldest months of last year and provided more than enough waste heat to meet the heat demands of the digester. This improved the overall system efficiency by displacing gas used for heating. If electrical generation is included with a digester system, it should be integrated with a system that can recover the waste heat for use elsewhere. This will significantly improve both the system's net energy and economic feasibility of electrical generation.

CONCLUSIONS

Anaerobic digestion is feasible and workable at farm scale. The work remaining is to refine digestion systems to improve their net energy performance, their reliability, and their ease of operatio The biological performance of the Monroe digester has been remarkably stable. Attempts to stress the system such as increasing loading rate, increasing percent total solids loaded and decreasing mixing have all resulted in an improved net energy. The biological stability and handling characteristics of dairy manure differ so significantly from municip sewage that digesters designed on the municipal model such as ours will be oversized, inefficient an prohibitively expensive. Unlike digestion for municipal sewage treatment--and, for that matter, most facilities that provide energy--farm scale digesters cannot pass costs on to a captive market. Digesters must therefore be designed within the constraints of farm economics and farm operations.

The concurrent goals of minimizing capital cost, electrical requirements and operator time while maximizing gas production offer an attainable engineering task. The information generated from full scale and pilot scale operations must be used to develop simple systems that can be used into farming operations. Operating demonstrations of reliable, economical and commercially available systems is essential for anaerobic digestion to meet its potential capacity to provide a significant amount of energy to the agricultural sector.

PILOT PLANT DEMONSTRATION OF AN ANAEROBIC, FIXED-FILM BIOREACTOR FOR WASTEWATER TREATMENT*

R. K. Genung and W.W. Pitt, Jr.
Chemical Technology Division
Oak Ridge National Laboratory
Oak Ridge, Tennessee 37830

Abstract

In response to the need for major near-term decreases in the energy intensiveness of contemporary technologies, Oak Ridge National Laboratory (ORNL) has been engaged in the development and demonstration of a pilot-plant wastewater treatment facility based on an anaerobic, fixed-film bioreactor. The bioreactor employs a process which consists of attaching microorganisms to stationary packing material and passing liquid wastes upward through the unit for continuous treatment by biophysical filtration and anaerobic fermentation. The process has been demonstrated using municipal sewage with a bioreactor designed to process 5000 gpd. Treatment of other wastewaters such as industrial discharges is presently being considered. The economic advantages of the process depend on the elimination of operating energy requirements associated with the aeration of aerobic-based processes and with the significant decrease of sludge-handling costs required with conventional activated sludge treatment systems. Methane production and recovery is also emphasized in the design of this innovative system.

With a view toward enhancement of technology transfer, the unit was designed during the summer of 1976 as a joint venture between ORNL and the Norton Company (Akron, Ohio). It was installed with the cooperation of the Norton Company and the city of Oak Ridge in the late fall of 1976, and was operated on a continuous basis for approximately two years with minimal downtime.

Data from this pilot plant operation and conceptual designs for large-scale systems based on the data will be presented. Economic comparisons between this new technology and conventional activated sludge systems are made.

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*Research sponsored by the Division of Building and Community Systems, U.S. Department of Energy under contract [illegible]405-eng-26 with the Union Carbide Corporation.

NOTES

Session V C

CONVERSION OF FOREST RESIDUE TO A METHANE-RICH GAS

H. F. Feldmann, S. P. Chauhan, K. T. Liu,
B. C. Kim, P. S. Choi, and H. N. Conkle
BATTELLE
Columbus Laboratories
505 King Avenue
Columbus, Ohio 43201

ABSTRACT

The objectives of this program are to reduce the cost of wood gasification to a gas that is suitable for either synthesis or a fuel gas that is interchangeable with natural gas. This will be accomplished by catalyses of the gasification reaction and developing a novel reactor system for biomass gasification.

Promising catalyst candidates previously identified in bench-scale screening studies are wood ash and calcium oxide. These catalysts are now being studied in a continuous reactor system which simulates conditions in a commercial gasification system. Results of gasification experiments with pure hydrogen, hydrogen/steam mixtures and pure steam will be described.

The novel multi-solid reactor system should allow significant economic benefits to be achieved by eliminating the oxygen required for gasification, minimizing front end feed preparation and increasing gasifier throughput. A pilot plant multi-solid gasifier is now being designed which will demonstrate these advantages.

Preliminary economic feasibility studies reported previously indicate that wood gasification by this process should be cost competitive with other supplemental clean fuels.

DESCRIPTION OF TASKS

This program consists of the following three tasks.

- Task 1 - Determination of the Effect of Gasification Parameters for both Catalyzed and Uncatalyzed Wood
- Task 2 - Design, Construction, and Operation of a Novel Multi-Solid Fluid-Bed Gasifier that can handle a wide range of feed material at higher throughputs than conventional reactor systems.
- Task 3 - Process Engineering Support

JECTIVES/COST AND PERFORMANCE TARGETS

he overall objective of this program is to develop a gasification process that will allow the production of a clean transportable gas from wood or other biomass materials at a price competitive with alternate energy supplies. The gas produced can either be used as a fuel for industrial or utility purposes that is interchangeable with natural gas or fuel oil or used for the synthesis of a variety of products ranging from ammonia to gasoline.

The cost criteria used as a basis for determining the potential competitive position of wood gasification are:

SNG from Coal	\$4.00/MM Btu
No. 2 Heating Oil	\$2.70/MM Btu
LNG	\$3.00/MM Btu

The program goal will be achieved by:

(1) Enhancing the reactivity of the wood by catalyzing it with either calcium oxide or wood ash. These materials are widely available, relatively cheap, and have been identified as increasing wood gasification rates in bench-scale experiments.
(2) Developing a novel reactor system based on a proprietary contacting scheme - the so-called multi-solid fluid bed (MSFB).

A preliminary cost feasibility study indicates that wood gasification using the MSFB gasifier should allow the economic objectives to be achieved.

APPROACH

The approach to achieving the program objective is described below for each of the program tasks.

Task (1) Process Chemistry of Catalyzed Gasification of Wood

The effect of controllable gasification parameters on conversion and product distribution is being determined with emphasis on the effects of previously identified gasification catalysts as well as in-situ methanation catalysts. These tests are being conducted in a continuous gasification reactor capable of operating over a complete range of conditions that are of practical interest.

Specific information being generated in this task will include:

- Data for detailed process heat and material balances.
- The determination of critical process parameters and their effects including the evaluation of wood ash and calcium oxide gasification catalysts.
- Determine the advantage of an in-situ methanation catalyst. This concept would use a methanation catalyst as the dense-phase fluid-bed medium to enhance methane production in the reactor.

The continuous gasifier used in these experiments is shown in Fig. 1 and is capable of operating at pressures up to 1000 psig and temperatures to 2000 F. Thus, it can be used to evaluate any conditions that could be considered commercially feasible.

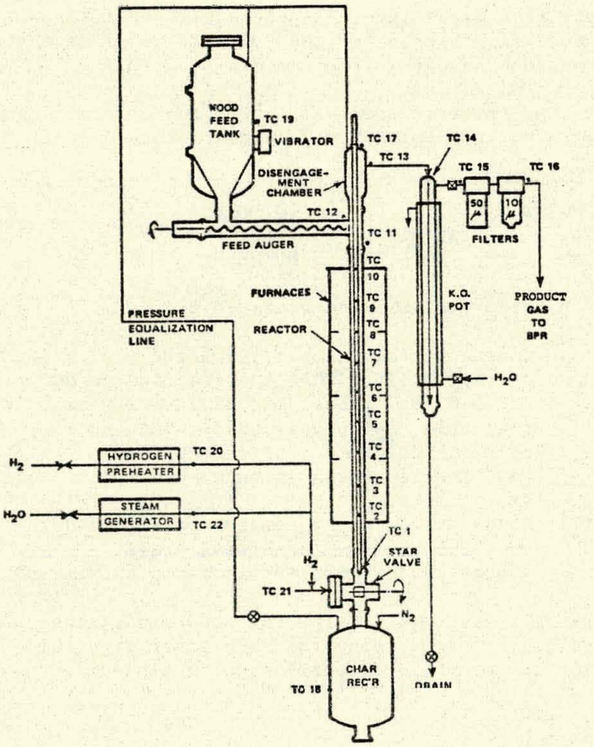

Fig. 1. Continuous Gasification Unit

An initial test series was conducted with the following types of wood pellets.

(1) Woodex^R Pellets--a mixture of hardwood and bark provided by Tennessee Woodex.
(2) Whitewood Pellets--a mixture of hardwoods provided by California Pellet Mill.
(3) Pellets made in our pelletizer from softwood sawdust.

The first two pellets worked well in the gasifier while enough disintegration of the softwood pellets, made in our lab, occurred to cause plugging of the reactor.

Key parameters studied in this reactor include:

- Wood Type/Particle Size
- Gasifying Agent - Steam/Hydrogen and Mixtures
- Wood Residence Time
- Gas Flow Rate
- Pressure
- Temperature

Task (2) Design, Construction and Operation of a Multi-Solid Fluid-Bed (MSFB) Wood Gasification Unit

The overall MSFB concept shown in Fig. 2 consists of a dense-phase fluid-bed gasifier through which flows an entrained phase (sand for example) to transport the heat required for gasification. The entrained phase flowing through the dense-phase fluid bed creates bubble-free fluidization, as well as increased heat and mass transfer in the bed. The transfer of char and sand from the gasifier to combustor is from an entrained phase rather than from the fluid-bed phase. Thus, with coarse wood chips, the conversion level in the MSFB can be considerably lower than that in the entrained phase which contains attrited char particles. Since the dense-phase bed material is different than the entrained material, it can be selected to allow the MSFB gasifier to operate at extremely high velocities without requiring that it operate in the so-called "fast fluid-bed" mode.

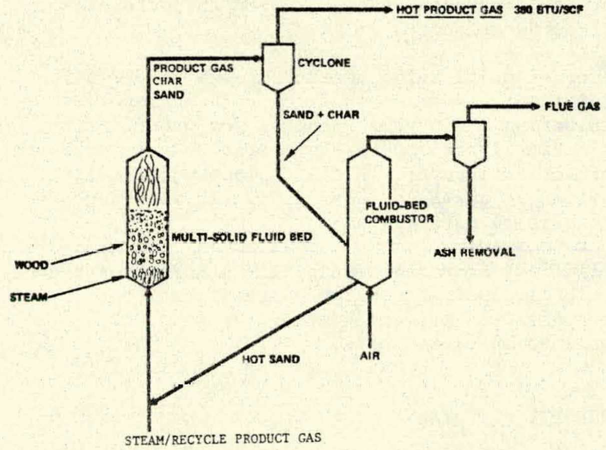

Fig. 2. Schematic of MSFBG Process

The utilization of this unique gasification system offers the following four major advantages over existing gasification systems.

(1) Elimination of the need for an oxygen plant. This improvement alone can reduce plant investment requirements by approximately 50 percent over a process using a fixed-bed slagging bottom gasifier.
(2) The ability to handle a wide range of wood feed from sawdust to chips without front end preparation and with minimal predrying which reduces investment and increases plant thermal efficiency.
(3) The capability of operating at very high solids throughputs because of the high

velocities at which the MSFB system operates. This allows more compact reactor systems further reducing investment requirements.

(4) The production of a lower CO_2 content gas which reduces distribution cost as well as the cost of acid gas removal.

This task was formally initiated May 1, 1979. Operating data from the MSFB gasifier will provide the basis for a commercial reactor design and a detailed cost assessment.

A sketch of the MSFB unit to be constructed is shown in Fig. 3.

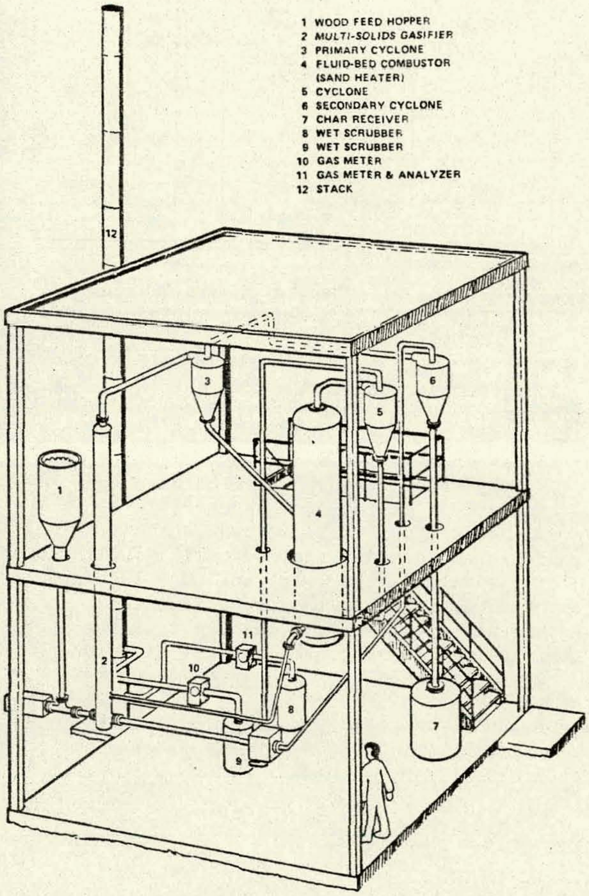

Fig. 3. Multi-Solid Fluid-Bed Gasification Process Development Unit

Task (3) Process Engineering Support

This task is dedicated to the translation of experimental data into commercially feasible process concepts and the determination of their economic benefits.

I[illegible] a previously reported "base" case cost feasi[illegible]ty evaluation of the MSFB gasifier it was estimated that an intermediate-Btu gas* could be produced at a price competitive with alternative forms of clean fuels. These results are summarized in Fig. 4 where the height of the bar depends on the cost of the wood which was assumed available at prices from \$10 to \$20/green ton. Also shown in Fig. 4 are the prices of other supplemental fuels that are either currently in use or contemplated for introduction into the market. This demonstrates that wood gasification could be currently competitive with No. 2 heating oil and LNG for 1000 (oven dry) tpd plants.

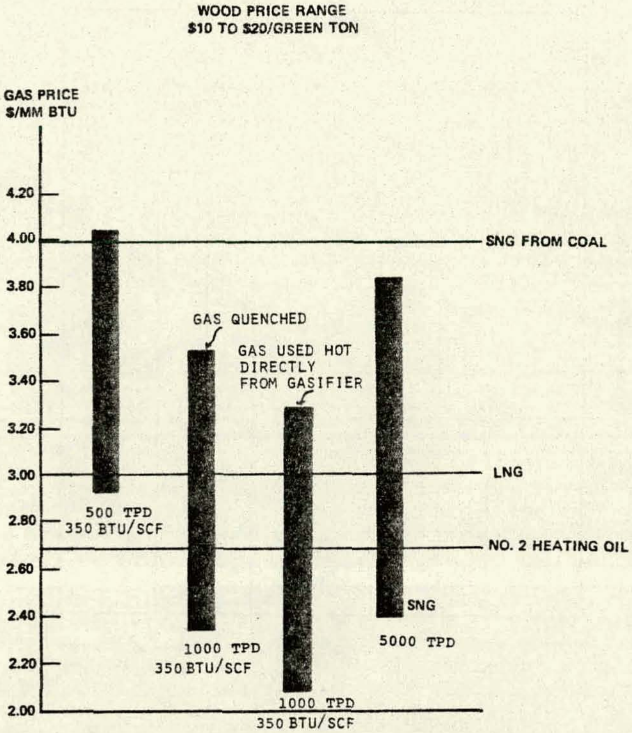

Fig. 4. Effect of Plant Size, Wood Price, Gas Heating Value, and Gas Quenching on Average Price of Gas from Wood

This task will resume in earnest when more data is available from the 2.8 inch I.D. reactor that is currently operating and a detailed conceptual design done when operating data from the MSFB gasifier is available. The prospects for substantial economic benefits over existing technology are great. For example, a recent economic evaluation[1] of an oxygen blown Purox^R system for producing medium-Btu gas from wood indicates that 60 percent of the total capital investment is for an oxygen plant and 10 percent of the annual operating costs are for electricity purchases for the oxygen plant.

KEY RESULTS/ACCOMPLISHMENTS

Some of the key results generated under Task (1) are listed below.

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*A as having a heating value typically above 300 Btu/SCF and which for most industrial purposes is interchangeable with natural gas.

- Steam is a more effective gasification agent for wood than hydrogen because steam gasification proceeds at a higher rate than hydrogasification and results in a greater net Btu recovery in the product gas. This is clearly illustrated by Fig. 5 which shows the increase in carbon conversion as the fraction of steam in the hydrogen/steam feed gas mixture is increased.

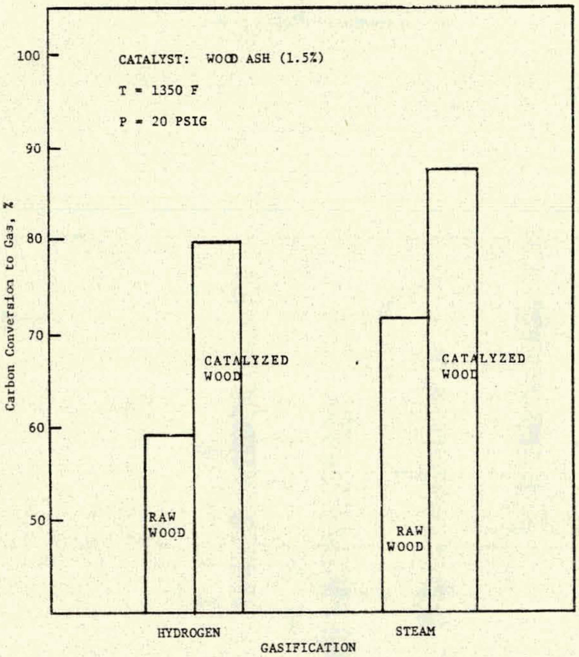


Fig. 5. Effect of Steam and Hydrogen on Carbon Conversion Level and Net Gaseous Btu Yield - 20 psig, 1350 F

The enhancement of the net Btu yield (defined as Total Heat Content of Product Gas-Total Heat Content of Feed Gas/Lb of Wood Fed) with steam compared to hydrogen is shown in Figure 6.

- Wood ash and calcium oxide are very effective catalysts enhancing both the carbon conversion level and the net gaseous Btu yield. Wood ash is effective in both hydrogen and steam gasification atmospheres while calcium oxide seems more effective in steam atmospheres than in hydrogen atmospheres. The effect of catalyses with wood ash on carbon conversion and net Btu yield is shown in Figures 5 and 6, respectively.

  In a commercial process, wood ash is available from the combustion of wood char to produce plant steam requirements and the heat for gasification. The wood ash would be recycled to achieve the optimum ash/wood feed ratio.

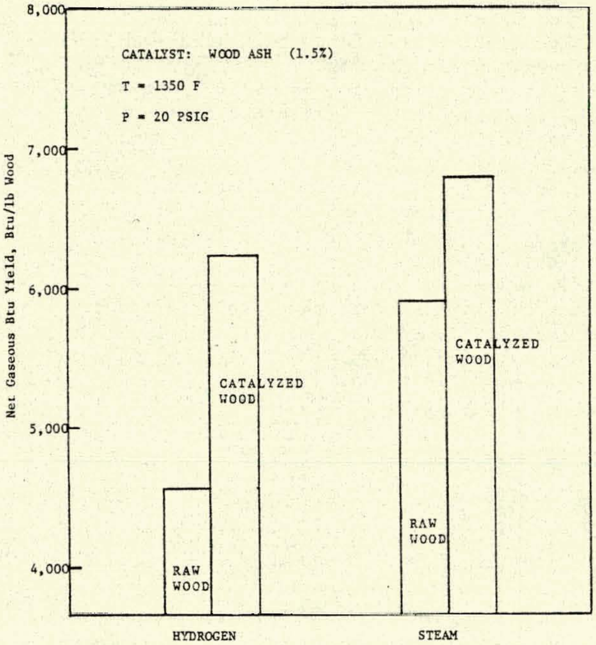


Fig. 6. Effect of Catalyst on Net Gaseous Btu Yield in Hydrogen and Steam Gasification

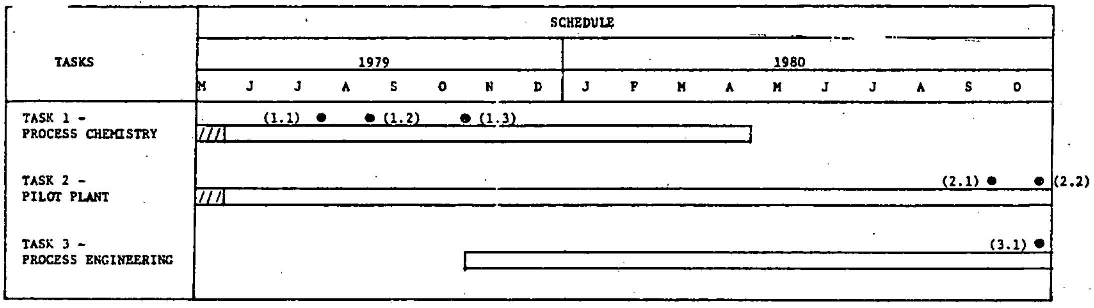
- Methane concentrations in excess of that predicted by thermodynamic equilibrium were achieved over the entire range of H_2/steam ratios.

  For example, with feed gases ranging from pure hydrogen to pure steam methane concentrations ranged from about 20 to 30 volume percent while the methane concentration predicted assuming thermodynamic equilibrium of the methanation reaction ($CO + 3H_2 \rightarrow CH_4 + H_2O$) qould be on the order of 1 to 2 percent. We therefore believe that the use of an in situ methanation catalyst is unnecessary.

- Design of the MSFB gasification unit was initiated May 1, 1979, and is proceeding on schedule.

FUTURE PLANS

A simplified schedule for the existing program is shown in Fig. 7 where the decision points to be reached during the present phase of the program. Assuming the experimental work demonstrates the economic potential of the technology, subsequent phases of the work would be aimed at rapid commercialization.



● PROGRAM DECISION POINTS

1.1 Should a Gasification Catalyst be Used?
1.2 Should Life Studies on Methanation Catalyst Initiated?
1.3 Are Temperature/Pressure(s) Practical for Commercial Operation?
2.1 Is MSFBG Technically Attractive?
2.2 What is the Next Logical Scale-Up Step for MSFBG System?
3.1 Is Wood Gasification via MSFBG Economically Attractive?

Fig. 7. Schedule for Existing Program

REFERENCES

[1] Desrosiers, R. E., "Process Designs and Cost Estimates for a Medium Btu Gasification Plant Using a Wood Feedstock, SERI/TR·33·151, (February 1979).

NOTES

CONVERSION OF BIOMASS MATERIALS TO GASEOUS PRODUCTS

Donald E. Garrett
Garrett Energy Research & Engineering Co.
911 Bryant Place, Ojai, California 93023

DOE Contract EY-76-C-03-1241
Phase II Start November 1978

ABSTRACT

The Garrett Energy Research & Engineering biomass gasification process is centered around a multiple hearth furnace of special design. The multiple hearth furnace was chosen because of its ability to process difficult-to-handle materials--even materials that cannot be processed in conventional air-blown biomass gasifiers. The stages of this multiple hearth furnace are as follows: biomass drying in a stream of hot flue gas, indirectly heated pyrolysis--the pyrolysis gas contains no N_2--at elevated temperature, combustion of part of the pyrolytic carbon, and finally a water gas stage in which the remaining carbon is gasified in a steam/air atmosphere.

When moist manure is fed to the pyrolysis hearth, a substantial part of the pyrolytic carbon is gasified there. A mathematical model of the low temperature water gas reaction is presented and used to correlate some pilot plant data.

OBJECTIVE

The objective of this work is to demonstrate that difficult-to-handle biomass materials can be economically gasified in a multiple hearth furnace of special design.

CONTRIBUTION

If successful, the GERE multiple hearth reactor will be able to process biomass materials that existing air-blown gasifiers cannot.

ORGANIZATION

The people in Garrett Energy Research & Engineering Co. who work on this project are as follows:

President	Donald E. Garrett
Biomass Project Manager	Ritchie D. Mikesell

GERE Headquarters, Ojai, California

Pilot Plant Supervisor	Dinh Co Hoang
Pilot Plant Chemist	David M. Kawano
[illegible]nician	Edward Andrada

GERE Pilot Plant, Liquid Chemical Corp.
Hanford, California

PROCESS DESCRIPTION

A flow diagram is shown on Fig. 1. The GERE PDU is a multiple hearth furnace of special design. The multiple hearth furnace was chosen because it is capable of handling difficult-to-handle materials--materials whose ash melts, agglomerating materials, materials that are wet, stringy, or that have low bulk densities.

The GERE PDU has four hearths--direct contact drying, pyrolysis, combustion, and water gas--and these hearths are fitted with solids valves in their downcomers. These valves pass solids but very little gas, so the hearths are effectively isolated from each other.

The pressure in the pyrolysis hearth is slightly positive. The drying hearth pressure is slightly negative, and the combustion hearth pressure is more negative, perhaps 20 inches of water. Oxygen-containing flue gas from the combustion hearth cannot therefore intrude into the pyrolysis hearth and create a dangerous situation there.

The biomass feed material is loaded into a bin which in turn feeds an elevated screw conveyor. Hot, aerosol-free pyrolysis gas flows countercurrently through the jacket and exchanges heat with the incoming biomass. The biomass is heated up as it moves through the conveyor, but very little of its moisture is removed.

At the top of the PDU, the warmed biomass falls onto the direct contact drying hearth where rabble teeth rake it across the perforated baseplate to the downcomer. Hot flue gas flows into the manifold and passes up through the bed of drying solids moving across the baseplate. The flue gas temperature is limited to about 200°C so that incipient pyrolysis and/or combustion does not occur in the drying hearth, and its flow rate is limited to about 100 SCFM so that excessive solids entrainment does not occur.

The partially dried biomass passes through a rotary solids valve into the pyrolysis hearth. A small flow of clean pyrolysis gas is injected into the solids downcomer below the solids valve. This prevents tarry pyrolysis gas from getting into this cold downcomer and plugging it. As the solids are raked across the baseplate heated from the combustion hearth below, they heat up, lose their moisture, and then pyrolyze. Some of the residual pyrolytic carbon then reacts with steam which comes partly from the moisture in the feed and partly from the pyrolysis reaction itself. In general, we desire to steam-gasify as much pyrolytic carbon as possible in the pyrolysis hearth, and so the feed to the pyrolysis hearth is only partially dried.

The pyrolytic char passes through the pyrolysis hearth downcomer and an internal solids valve of the sliding gate type into the combustion hearth below. The solids valve is operated by an automatic pneumatic driver which is outside the hearth. This valve must be simple, cheap, and it must work reliably in the high temperature environment of the combustion hearth. Only enough of the carbon is burned in the combustion hearth to maintain a minimum temperature level (600°C) in the pyrolysis hearth above; the rest of the carbon is fed to the water gas hearth below.

Another internal sliding valve similar to the one between the pyrolysis and combustion hearths regulates the flow of char into the water gas hearth. Three mols of steam and about 0.7 mols of oxygen, together with the nitrogen from the air, per mol of carbon are preheated to 800°C and injected through the water gas manifold. The water gas and combustion reactions are counter-balanced thermally so that the reactor temperature maintains itself at about 800°C. The ash falls through an external rotary solids valve and out of the process. The combustion and water gas hearths are lined with fire bricks.

The pyrolysis gas leaves the pyrolysis hearth at about 600°C and contains heavy tar and oily aerosol material. This gas is cooled to about 250°C in the inner pipe of a vertical double pipe heat exchanger. Hot flue gas is the coolant. Hot flue gas is used so that the inside surface of the inner tube does not get colder than the solidification temperature of the heavy tar. The heavy tar runs down the vertical wall of the inner tube into a heated trap below.

The cooled pyrolysis gas is now filtered through a fixed char bed to remove its aerosol content. This filtration is done at about 200°C, well above the dew point of the pyrolysis gas. There are two parallel filters, operated one at a time, and the spent char is recycled through the process.

Finally, the cooled and clean pyrolysis gas is passed through the conveyor jacket where it loses sensible heat to the incoming biomass. The pyrolysis gas is sufficiently clean that the heat transfer surface does not become fouled.

The incoming air for the water gas reaction is preheated by heat exchange against hot flue gas from the combustion hearth. This heat exchanger is a double pipe with a 3 inch outer tube and a 1½ inch inner tube with 24 longitudinal fins. The air passes through the jacket. The steam for the water gas reaction is superheated in a coil of 1 inch tubing wound around the inside wall of the combustion hearth. Cold tap water is deionized and fed to this coil.

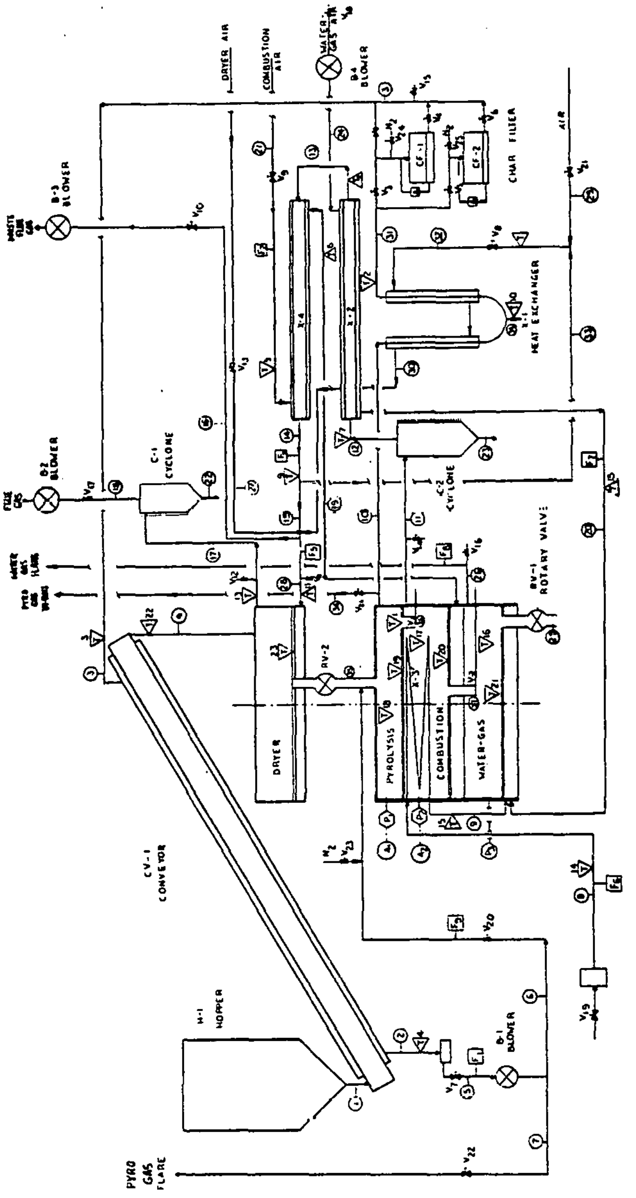


Fig. 1. GERE Biomass Gasification Process

Flue gas from the combustion hearth first passes through a cyclone. Past experience with the PDU combustion hearth suggests that we should expect little or no ash entrainment in the flue gas.

The clean flue gas then exchanges heat with the incoming water gas air. The water gas air flow rate is relatively small, so the flue gas doesn't cool much in this heat exchanger.

The flue gas then exchanges heat with the incoming combustion air. The air is preheated to about 600°C, while the flue gas is cooled to about 200°C.

hi at exchanger is also a double pipe with a inch outer pipe and a 1½ inch inner pipe with 4 longitudinal fins.

old air is then drawn into the flue gas to bring ts temperature down to about 150°C so that it an be used to cool the raw pyrolysis gas. In the yrolysis gas cooler, the flue gas and pyrolysis as exchange heat counter-currently. The outgoing lue gas leaves the heat exchanger substantially otter than 200°C, so more cold air is drawn into t to bring it down to this temperature.

00 SCFM of the flue gas at 200°C is injected into he manifold of the direct contact dryer, and the emaining flue gas is wasted. This wasted flue as could, of course, be used in a second dryer. n the direct contact dryer, the gas cools and ecomes humid. It may also entrain some of the ines, so the gas leaving the dryer is made to ass through a cyclone before it is rejected.

OST AND PERFORMANCE TARGETS

he GERE multiple hearth PDU has a diameter of four eet. We hope to be able to process 6 tons manure 50 wt % moisture/day in this reactor. The economics of the process have not been worked out yet. t may be possible to use carbon steel instead of tainless steel for the combustion and water gas earths. It also seems likely that simple, relatively cheap sliding gate valves can be used etween the stages.

INDINGS

substantial fraction of the pyrolytic carbon s steam-gasified in the pyrolysis hearth at elatively low temperatures and short solids esidence times when the incoming manure is oist, say 20 - 40 wt %. Pyrolysis experiments ere done in a single hearth bench scale reactor n order to generate design data for the PDU. he data showed that pyrolytic carbon had been team-gasified at char temperatures as low as 00°C at char residence times of 10 to 20 minutes.

ccording to Finson, et al. [1], the reaction equence is

$$\underset{\text{Active Site}}{C} + H_2O \underset{k_{-2}}{\overset{k_2}{\rightleftharpoons}} \underset{\text{Complex}}{C(O)} + H_2 \qquad (1)$$

$$C(O) \xrightarrow{k_5} CO \qquad (2)$$

t low temperatures, the absorption/desorption step s much faster than the complex breakdown step, so he BET surfaces should be saturated with C(O). n this case the gasification rate is just

$$-\frac{1}{[C]}\cdot\frac{\partial [C]}{\partial \tau} = k_5 \cdot \sigma\beta \qquad (3)$$

$$-\frac{1}{[C]}\cdot\frac{\partial [C]}{\partial \tau}$$ is often called char reactivity.

Note that when the BET surface is saturated with the C(O) complex, the steam concentration has no direct effect on the char reactivity.

A model of BET surface development gives the result

$$\frac{\sigma\beta}{(\sigma\beta)_0} = \frac{1}{(\sigma\beta)_0 + \{1 - (\sigma\beta)_0\}\exp\left(-\frac{k_5\tau}{\sigma}\right)} \qquad (4)$$

When this is inserted into Equation (3), the result is

$$Y = \frac{[C]}{[C]_0} = \frac{\exp\left(-\frac{k_5\tau}{\sigma}\right)}{(\sigma\beta)_0 + \{1 - (\sigma\beta)_0\}\exp\left(-\frac{k_5\tau}{\sigma}\right)} \qquad (5)$$

$(\sigma\beta)_0$ is the fraction of all the carbon atoms present in the virgin char that are exposed to the steam on the BET surfaces. At low temperatures, the steam-carbon reaction is sufficiently slow that mass transfer resistance in the gas phase is unimportant. For a continuous process in which the char bed is well-mixed, Equation (5) becomes

$$Y = \frac{1}{1 + \dfrac{(\sigma\beta)_0 \cdot \left(\frac{k_5\tau_R}{\sigma}\right)}{\left[(\sigma\beta)_0 + \{1 - (\sigma\beta)_0\}\exp\left(-\frac{k_5\tau_R}{\sigma}\right)\right]}} \qquad (6)$$

Equation (6) with $(\sigma\beta)_0 = 1$ was used to correlate our bench scale pilot plant data. For this case

$$\frac{1}{Y} - 1 = \frac{k_5^{*}\tau_R}{\sigma}\cdot\exp\left(-\frac{E}{T_c}\right) \qquad (7)$$

The Arrhenius plot is

$$\ln\left[\frac{1}{\tau_R}\cdot\left(\frac{1}{Y} - 1\right)\right] = \ln\frac{k_5^{*}}{\sigma} - \frac{E}{T_c} \qquad (8)$$

The data correlation is shown on Fig. 2. The result is

$$\frac{1 - Y}{Y} = 19381\cdot\frac{V_c}{W_c}\cdot\exp\left(-\frac{12065}{T_c}\right) \qquad (9)$$

For example, at $T_c = 800^0C = 1073.2^0K$, and $\tau_R = 20$ min, Equation (9) predicts that about 84% of the pyrolytic carbon will be steam-gasified. It appears that the steam-carbon reaction is catalyzed by the presence of volatile matter in the char, and therefore it proceeds at reasonable rates at relatively low temperatures.

When pyrolysis is done in a steam atmosphere, the energy content of CO is shifted to H_2 through the water gas shift reaction.

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad (10)$$

The heating value of the resulting gas is slightly lowered as Reaction (10) is driven to the right.

The aerosol tar content of pyrolysis gas can be removed in a char filter. This makes possible heat recovery from the pyrolysis gas stream. When the pyrolysis hearth feed is moist and the pyrolysis gas contains a great deal of steam, heat recovery becomes more important to the thermal efficiency of the process.

A simple automatic valve can be used between the stages to isolate them from each other. These valves are of the sliding gate type, are driven from outside the reactor, and must work reliably at high temperatures.

RAW MATERIAL AND ENERGY REQUIREMENTS

100 T/D Manure Plant

Input	100 T/D Manure @ 40 wt. % moisture. Net heating value = Heat of combustion of dry manure - Heat to evaporate moisture Net heating value = 5.57 (10^8) BTU/day
Electricity for Blowers and Valves	∿40 kg = 0.05(10^8) BTU/day heating value
Output	
Pyrolysis Gas	635 SCFM @ 310.5 BTU/SCF (dry, with CO_2). Heating value = 2.84(10^8) BTU/day
Producer Gas	None
Waste Water	5.1 GPM

$$\text{Overall thermal efficiency } \frac{2.84(10^8)}{5.62(10^8)} = 50.5\%$$

ECONOMICS

A final economic analysis has not been completed.

FUTURE ACTIVITIES

(a) To complete the present contract.

(b) To propose a follow-on contract.

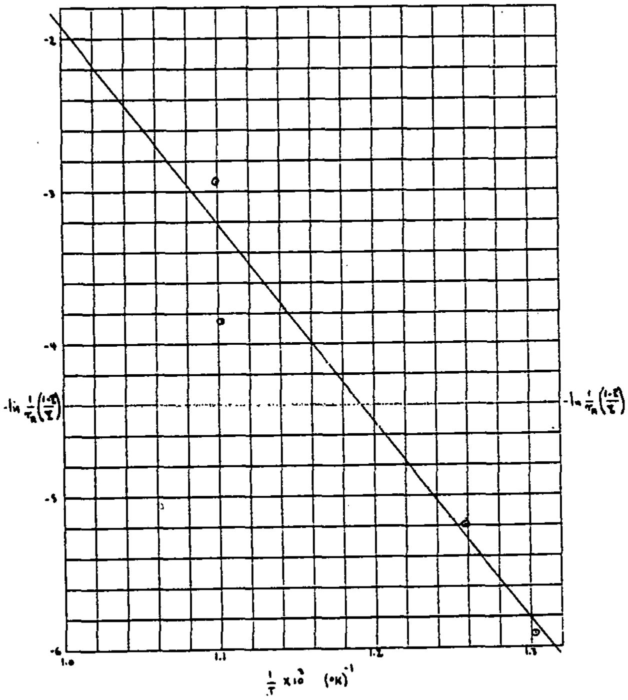
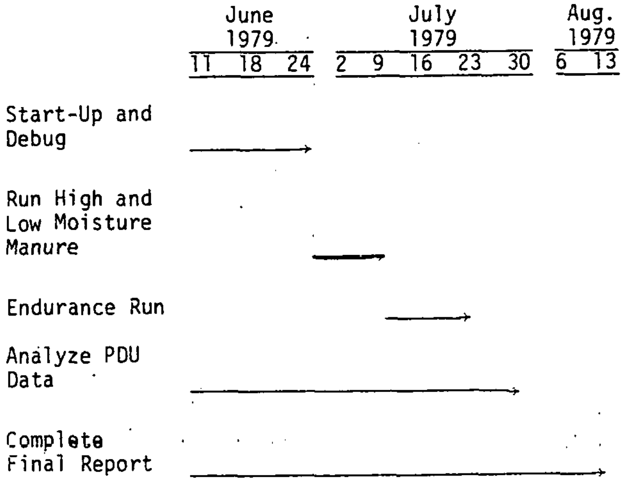


Fig. 2. Low Temperature Water Gas Reaction Kinetics

WORK SCHEDULE



NOMENCLATURE

$[C]$	Carbon concentration, g C/cm³ char
$[C]_o$	Initial carbon concentration, g C/cm³ char
$E = E^*/R$	Emperical constant, °K
E^*	Activation energy, cal/mol

V_c	Char volume, cm^3
W_c	Char volumetric flow rate, cm^3/min
$Y = \frac{[C]}{[C]_o}$	Fraction carbon remaining
β	BET surface area, cm^2 BET/mol carbon
σ	Carbon concentration in the BET surface, mols C/cm^2 BET
τ	Time, min
$\tau_R = V_c/W_c$	Char residence time, min

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NOTES

CATALYTIC CONVERSION OF BIOMASS TO FUELS

K. K. Ushiba, I. Mahawili, M. C. Cooper and R. L. Garten

Catalytica Associates, Inc.
3255 Scott Blvd., Suite 7-E
Santa Clara, California 95051

ABSTRACT

This study concerns a systematic assessment of the role of catalysis and new developments in catalytic reactor technology for the thermochemical conversion of biomass to gaseous and liquid fuels. In this report the physical and chemical properties of biomass are compared to coal as a basis for assessing the utility of coal conversion concepts for biomass. In addition, the role of catalysts in the steam gasification of carbon is examined with regard to the implications for gasification of biomass. Recent studies demonstrate that known catalysts (alkaline earths, Group VIII metals) for the carbon-steam do not increase reaction rates by a lowering of the activation energy but by increasing the concentration of catalytic sites in contact with the carbon source. Maintaining maximum contact between the carbon source and the catalyst throughout gasification is therefore necessary to achieve the greatest rate enhancement. This generally requires undesirably high catalyst concentrations which necessitates recovery and reprocessing of the catalyst. The most desirable catalyst is one which enhances the rate by lowering the activation energy and which could be utilized in low concentrations. Such materials have not been reported to date.

INTRODUCTION

The dwindling supply of indigenous petroleum in the United States and the increasing dependence on foreign imports has provided incentive for the development of other sources of energy. In this regard, coal has been and continues to be the most intensely studied alternative to petroleum. Extensive catalytic science and technology are being applied by both government and industry to economically convert this substance to clean-burning gas and liquid fuels. Another potential fuel source in which interest has escalated during this decade is biomass which, in contrast to coal, represents a renewable supply of energy via photosynthetic processes. The utilization of catalytic systems in biomass conversion processes has received some attention. However, no assessment of the potential impact of catalysis and of new developments in catalytic reactor technology has been undertaken for thermochemical biomass conversion processes for the production of gas and liquid fuels. The primary objectives of this project, therefore, are:

- to provide a systematic assessment of the role of catalysis in thermochemical gasification and liquefaction,
- to establish the relationship between potential biomass conversion concepts and the catalytic concepts currently under development in other areas, such as coal conversion,
- to identify promising catalytic systems that could be utilized to reduce overall costs of fuels production from biomass materials.

The approach utilized in this study involves a review of the patent and technical literature concerning concepts for the thermochemical conversion of coal to gas and liquid fuels, and an identification of the major technical problem areas. This background provides the basis for an assessment of the potential applicability of the various concepts to biomass which, because of the marked differences in physical and chemical properties relative to coal, may significantly reduce the problems encountered with coal. Limitations and problems which are specific to biomass are also identified. Finally, the most promising technologies for the thermochemical conversion of biomass to fuels will be identified, and the problems which must be solved to develop the identified technologies are discussed. It should be emphasized that the focus of this study is on the chemistry and reactor engineering aspects of thermochemical conversion processes rather than on total systems and detailed economic analysis. Thus, the recommentations concerning technologies for future developments are based on an assessment of the most promising catalytic and reactor engineering concepts and a judgement as to their economic feasibility.

The project is divided into four major tasks:

Task 1	A review of the literature on thermochemical conversion concepts for coal and biomass
Task 2	Identification of major technical problem areas in biomass conversion processes
Task 3	Selection of most promising technologies for biomass conversion
Task 4	Recommendations for future R & D

Task 1 which involved a review of chemical and engineering concepts for the numerous gasification, hydrogasification, pyrolysis and liquefaction technologies has been completed. The focus of recent efforts on this project has concerned various aspects of Tasks 2 and 3 which are discussed in this report. These aspects include a comparison of the physical and chemical properties of biomass relative to coal and a discussion of key problems associated with the catalytic steam gasification of biomass. The latter discussion draws on recent fundamental studies which illustrate the mode of catalytic action in altering the rate of the carbon-steam reaction and considers the implications of the results for biomass gasification.

CHEMICAL AND PHYSICAL PROPERTIES OF BIOMASS AND COAL

In order to provide a perspective for the potential advantages of biomass relative to coal for synthetic fuels production, a comparison of their chemical and physical properties was made. In this section a brief summary of the key differences is presented as a basis for assessing the utility of coal conversion concepts for biomass and the suitability of biomass as a feedstock for synthetic fuels production.

Table 1 provides a general comparison of the physical and chemical properties of biomass and coal. Clearly there are marked differences which appear to favor biomass over coal as a feedstock for synthetic fuels. The H/C ratio of biomass is especially favorable approaching values typical of petroleum derived fuels. This is illustrated by the comparison in Figure 1 which shows that the H/C ratio of biomass is comparable to that of fuel oils. Figure 1 also provides a comparison of typical O/C ratios of biomass with various coals. The oxygen content is, of course, maximum with biomass producing the lowest Btu content of the materials compared (Figure 2). The implications of the elemental compositions for upgrading are that conversion of coal to liquid fuels requires an increase in the H/C ratio either by the addition of hydrogen (hydroliquefaction) or the removal of carbon (pyrolysis). Conversion of biomass to liquids, on the other hand, may be achieved by removal of oxygen, an approach which forms the basis of the liquefaction project currently underway in Albany, Oregon.

TABLE 1

GENERAL COMPARISON OF THE PHYSICAL AND CHEMICAL PROPERTIES OF BIOMASS WITH COAL

	Coal	Biomass
H/C	0.2 - 0.8	1.6
O/C	0.007 - 0.25	0.68
Ash	7-12 wt.%	0.2 - 3 wt.%
S	0.8 - 3.4 wt.%	Traces
N	1-16 wt.%	Traces
H_2O	Bit. ∿ 2% S-Bit. ∿ 10% Lignite ∿ 30%	Bark ∿ 0% Wood ∿ 10% Bagasse ∿ 50%
Density	1-2(Lig.)-1.75 (Anthr.)	0.4(S-Wood)-0.6 (H-Wood)
Volatile Matter (%)	9(Anthr.)-36 (Bit.)	∿ 87

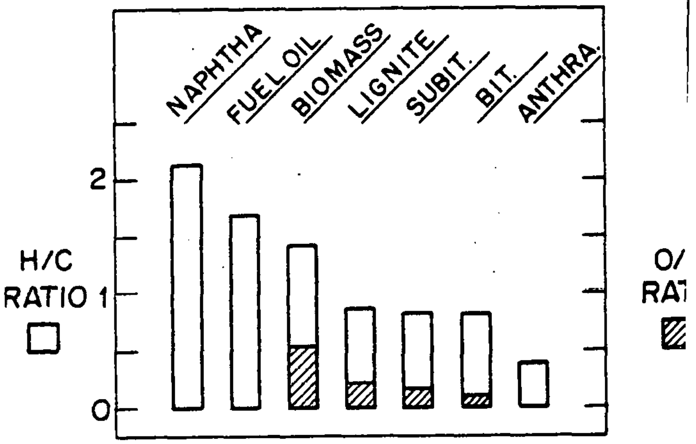

Figure 1. Comparison of the H/C and O/C ratios of biomass with petroleum derived products and coal.

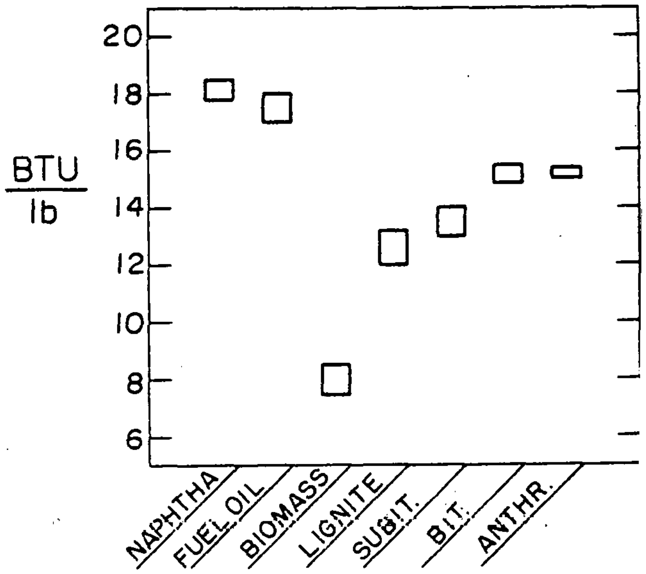

Figure 2. The energy content of biomass relative to petroleum derived products and coal.

In gasification, the elemental composition of biomass coupled with its higher reactivity and larger fraction of volatile matter also appears favorable in that the steam, oxygen and feed requirements would be lower relative to coal.

Perhaps the most attractive aspect of the chemical properties of biomass is the low concentration of impurities, S, N, and ash. Figure 3 provides a general comparison of S, N, and ash levels of biomass, coals and petroleum derived products. The low concentrations of impurities in biomass significantly reduces the heteroatom removal requirement relative to processing of coal derived fuels. In addition, the low impurity concentration broadens the scope of potential catalytic systems which could be applied for synthetic fuels production. This is particularly true for poison sensitive catalysts which are used in combination with gasification catalysts in single step processes to generate particular products such as CH_4, H_2, CO or synthesis gas (1).

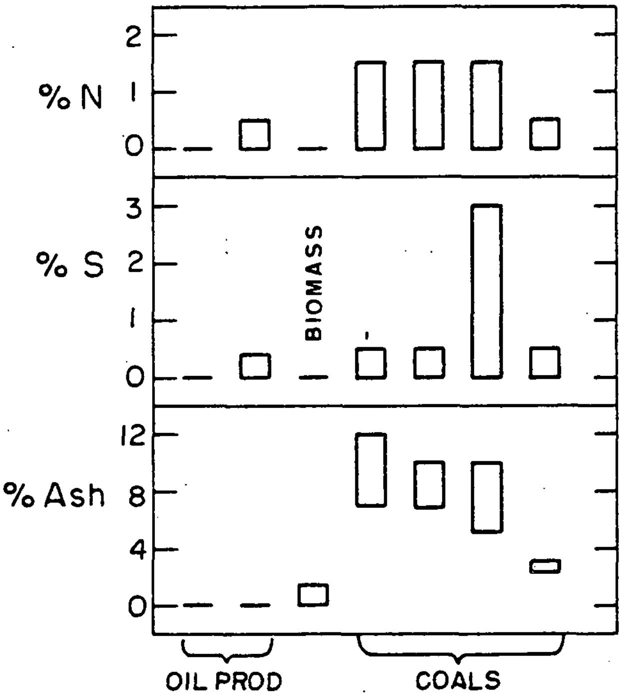

Figure 3. Impurity levels in biomass relative to other fuel sources.

Although the chemical properties of biomass, in general, are quite uniform and attractive for synthetic fuel production, the physical properties vary considerably depending on the biomass resource, e.g. wood chips, hulls, straws, stalks, etc. Such materials are not easily pre-processed by grinding, for example, to the small particle sizes which are attainable with coal. Pre-processing and feeding of biomass, therefore, are important considerations in choosing conversion technologies. Conversion processes, which require little pre-processing and accept a wide variety of feed geometries and particle sizes appear desirable. Alternatively, it is likely that several technologies applied on a regional basis and designed to handle feedstocks with a specified range of physical properties will evolve.

THE ROLE OF CATALYSTS IN STEAM-GASIFICATION OF BIOMASS AND COAL

General Remarks

The steam-gasification of carbonaceous materials can be schematically represented by the diagram in Figure 4. As the temperature increases, the first processes to occur are pyrolysis or devolatilization reactions which produce volatile matter and ash-containing char. With regard to pyrolysis, biomass exhibits much greater reactivity than most coals as is illustrated by the comparison in Figure 5. Pyrolysis of cellulose materials is initiated at relatively low temperatures (300-500°C) and a large portion of the feed (70-90 wt %) is converted to volatile matter depending on the heating rate (2). The higher the heating rate, the less char is produced. In fact, Lewellen et al. (3) have reported complete volatilization of cellulose at very rapid heating rates of $>10,000^{\circ}$C/sec. With coals, pyrolysis requires higher temperatures than biomass (Figure 5) and the portion of volatile matter produced is much lower ($<\sim 50$ wt %).

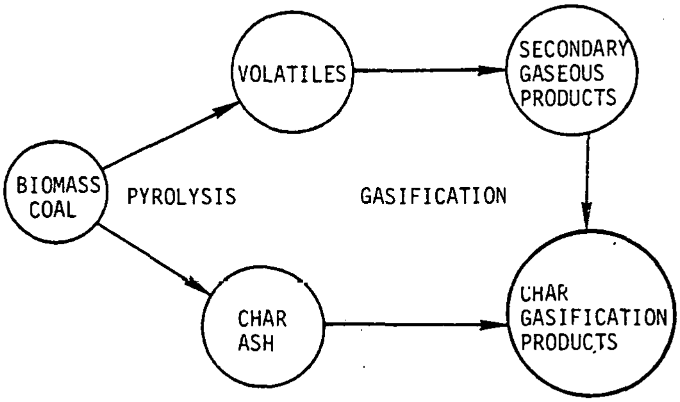

Figure 4. Schematic diagram representing steps in steam-gasification.

At temperatures above $\sim 600^{\circ}$C, secondary reactions of the devolatilization products occur. These processes may be influenced by the addition of appropriate catalysts to direct the reactions towards particular products (1). In the absence of a catalyst the secondary reactions of the volatilization products are similar to steam-cracking reactions (2).

At the highest temperatures, $>700^{\circ}$C, the pyrolytic char undergoes steam gasification to produce

H_2, CO and CO_2 via the following reactions:

I $C + H_2O \Leftrightarrow CO + H_2$ (carbon gasification)

II $CO + H_2O \Leftrightarrow CO_2 + H_2$ (shift reaction)

III $C + CO_2 \Leftrightarrow 2CO$ (Boudouard reaction)

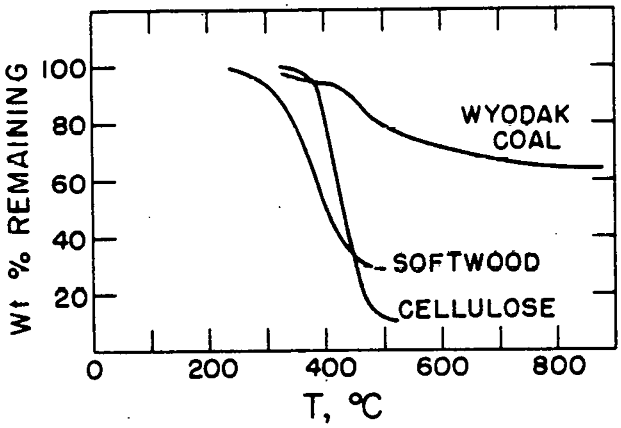

Figure 5. Comparison of the pyrolysis reactivity of cellulosic materials and coal(2)

The high temperatures required to promote the carbon reactions (I and III) reflect the slow rates of these reactions and the desirability of using catalysts to enhance the reaction rates.

For steam-gasification processes in which methane is the desired product, the chemical dilemma is an inefficient thermal balance. This is apparent from the following reactions:

Gasification	$\Delta H_{298°K}$ (kcal/mole)	Reaction Temperature
$2C + 2H_2O \Leftrightarrow 2CO + 2H_2$	+ 62.76	1000°C(1800°F)
Shift		
$CO + H_2O \Leftrightarrow CO_2 + H_2$	- 9.83	300°C(600°F)
Methanation		
$3H_2 + CO \Leftrightarrow CH_4 + H_2O$	- 49.27	400°C(700°F)
$2C + 2H_2O \Leftrightarrow CH_4 + CO_2$	+ 3.66	

Thus, the sum of the reactions which give rise to the formation of CH_4 are nearly thermoneutral. However, the water-gas shift and methanation steps are carried out at much lower temperatures for thermodynamic reasons and catalyst maintenance. Therefore, the exothermic heat generated in these two steps cannot be efficiently utilized at the higher temperature for steam-gasification. The potential solution to the problem is catalytic gasification at 800-900°F. Then the heat from the shift and methanation steps could supply that required for the gasification reactions. This possibility of a "single step" gasification has been a primary motivation for catalytic steam gasification.

To have a "single-step" gasification reactor, the gasification, water-gas shift and methanation reactions must be catalyzed simultaneously. This requires a bifunctional catalyst. For the steam gasification reaction, alkali metal oxides and carbonates have been found to be among the best catalysts. The water-gas shift and methanation reactions are usually catalyzed by metal catalysts such as Ni or Fe. Consequently, combinations of Ni and other transition metals with the alkali metal oxides or carbonates have been explored as "single-step" gasification catalysts (1, 4-6). Massive amounts of catalyst are usually required for acceptable activity with coal which contains significant quantities of catalyst poisons such as sulfur. With biomass, however, the concentrations of potential catalyst poisons, particularly sulfur, are significantly lower which may provide incentive for the development of methanation catalysts that are unaffected at the low sulfur concentrations typical of biomass.

Because the carbon-steam reaction is the slowest process in gasification of chars derived from biomass, it is useful to focus on catalysis of this reaction. A number of recent studies have provided important insight regarding the role of the catalyst in steam gasification. These studies and their implications for biomass conversion are discussed in the following section.

Catalyzed Steam-Gasification of Carbon

Much of the earlier work on the carbon-steam reaction was in conflict in that different authors reported that the same catalyst enhanced, had no effect or even suppressed the rate of gasification (7). This circumstance was undoubtedly due to improper attention to the influence of heat and mass transfer effects and to the ambiguities associated with the reporting of rate data. The results of the majority of the earlier work did not contribute much to the understanding of the role of the catalyst in steam-gasification.

In a recent series of reports (7-10), Shelef and co-workers have carefully examined the effect of a variety of catalysts in the steam gasification of pure graphite and coal chars. These authors measured gasification rates in the presence and absence of catalysts under conditions which eliminated mass and heat transfer effects. Only under such conditions can the effect of catalysts be quantitatively established. In addition, Shelef et al compared gasification rates on a specific surface area basis (quantity of carbon gasified per unit time per unit area) rather than on a un weight basis. The importance of making comparisons on a unit area basis is illustrated by the data in Table 2. When compared on a weight basis, the gasification rates at the extremes differ by a factor of 1750 whereas on a unit area basis the difference is only a factor of 15.7. Clearly the difference in reactivity of the materials is due primarily to the large variation in surface areas. The higher specific area reactivity of the Brannkoble was shown to be due to the catalytic effects of native ash. The data in Table 2 indicate that false conclusions concerning catalytic effects may be reached when data is compared on a weight basis particularly if the addition of a catalytic agent produces a modification of the surface area of the material to be gasified.

TABLE 2

COMPARISON OF THE STEAM-GASIFICATION OF COAL CHARS AND GRAPHITE AT 850°C (11)

Coal	Bet Area (m^2/g)	$\mu g\ g^{-1}\ min^{-1}$	relative rate	$\mu g\ min^{-1}\ m^{-2}$	relative rate
Braunkohle	436	2100	1750	4.7	15.7
Pittsburgh Seam	407	280	233	0.7	2.3
Graphite	3.6	1.2	1	0.3	1.0

Other important aspects of the work by Shelef et al concern the mode of catalytic action. These authors compared the activation energies and specific area rates of the carbon-steam reaction in the presence and absence of catalysts. The results are summarized in Figures 6 and 7. Figure 6 shows that the activation energy of the carbon-steam reaction is not significantly affected by the addition of alkaline earth catalysts, Ca, Sr and Ba. The rate of gasification on a specific area basis, however, increases substantially in the presence of these catalysts with Ba exhibiting the greatest enhancement. Similar results are obtained with catalysts from the Group VIII elements (Figure 7). In this case activation energies appear to be decreased somewhat, however, Shelef et al concluded that this was due to the onset of mass transport effects rather than a catalytic lowering of the activation energy. For the Group VIII metals much greater enhancements in specific area gasification rates were obtained (~7600 relative to the uncatalyzed reaction) with rhodium exhibiting the highest activity. It is unlikely that expensive catalysts such as rhodium could even be seriously considered as a gasification catalyst.

The implications of the results illustrated in Figures 6 and 7 become apparent from the Arrhenius equation:

$$\text{Gasification Rate} = A \exp(-E_a/RT)$$

where A is the pre-exponential factor which is related to the concentration of catalytic sites and E_a is the activation energy. The most desirable effect of a catalyst is to decrease E_a which leads to an exponential increase in rate. Rate is also enhanced by increasing the concentration of catalytic sites but the effect in this case is linear. The data in Figures 6 and 7 indicate that the activation energy of the steam-gasification is unaffected by the catalysts examined to date and that the enhancement in rate is primarily due to an increase in the pre-exponential factor, i.e., the concentration of catalytic sites. This implies that the catalyst does not affect the rate determining step of the carbon-steam reaction but may increase the concentration of reactive species which participate in the rate determining step. In a simplified scheme, the carbon-steam reaction can be envisioned as two steps, viz:

- Activation of H_2O

$$H_2O \longrightarrow 2H_{(a)} + O_{(a)}$$

- C-C Bond Rupture (rate determining)

$$C\text{-}C + O_{(a)} \longrightarrow CO + C$$

In this scheme the catalyst increases the concentration of the species $[O_{(a)}]$ but does not alter the energetics of the rate determining step involving C-C bond rupture. Thus, the rate of gasification is increased without altering the activation energy as observed experimentally.

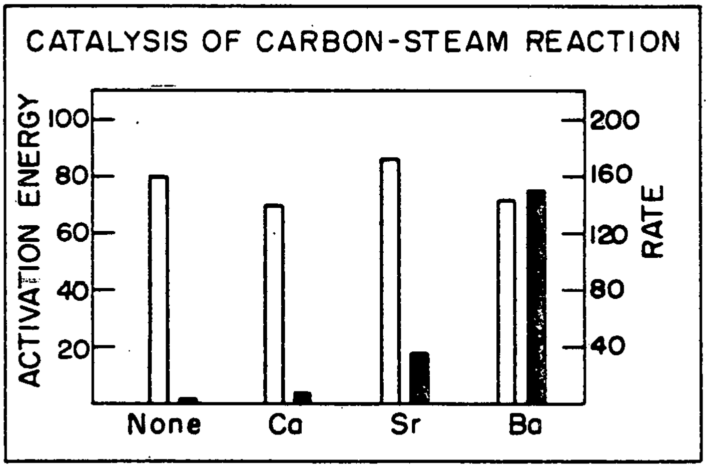

Figure 6. Comparison of activation energies (kcal/mole) and rates ($\mu g/m^2/min$) for catalyzed and uncatalyzed steam-gasification of graphite. Data pertain to reaction at 850°C and 0.3 atom % catalyst concentration (10).

The fundamental studies of Shelef and co-workers have significant implications concerning catalysis of the carbon-steam reaction. The results indicate that the largest catalytic enhancement of the reaction rate is achieved when contact between the catalyst and the carbon is maximized. Indeed, Shelef et al found from careful characterization studies by physical techniques that the efficacy of Ba as a catalyst relative to Ca and Sr was due to more effective wetting of the carbon (10). The Ba catalyst tended to spread uniformly over the surface of the carbon during the reaction whereas the other catalysts were less effective in this regard. If the catalyst effectively wets the carbon during the gasification reaction then the amount of catalyst necessary to achieve the maximum effect will depend on the surface area of the carbon as shown in Figure 8. For high surface area carbons very large catalyst concentrations are required so that recovery of the catalyst is necessary. Clearly, the most desirable catalyst is one which would enhance the rate of the carbon-steam reaction by lowering the activation energy rather than increasing the pre-exponential factor. Catalysts which lower the activation energy, in the absence of mass transport effects, could be used in exceedingly low concentrations reducing or alleviating the recovery problem.

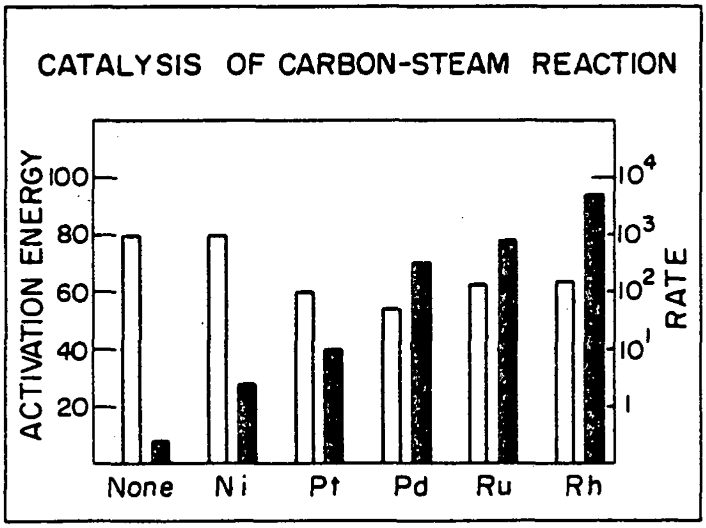

Figure 7. Comparison of activation energies (kcal/mole) and rates ($\mu g/m^2/min$) for catalyzed and uncatalyzed steam gasification of graphite. Data pertain to reaction at 850^oC and 1 wt.% catalyst concentration (7).

If catalysts which lower the activation energy cannot be found, then gasification technologies which provide for maximum catalyst-carbon contact throughout the course of gasification or which do not form char may be attractive. Technologies which may fulfill these criteria are currently being examined and will be discussed in future reports.

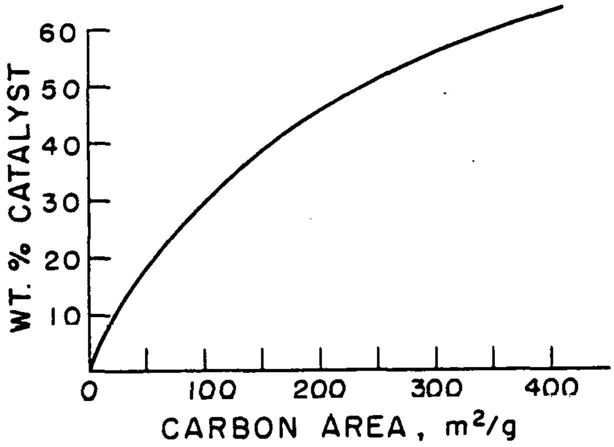

Figure 8. Relationship between catalyst concentration required to achieve monolayer catalyst coverage (maximum catalyst-carbon contact) and carbon surface area. Example is presented for a Barium catalyst assuming area covered by one Ba atom is $7.6\ \text{Å}^2$.

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RESEARCH AND EVALUATION OF BIOMASS RESOURCES/CONVERSION/UTILIZATION SYSTEMS

Yong Kee Ahn, Richard J. Brumberg, Herbert T. Chen, Eric T. Nelson, and Richard P. Stringer
Gilbert Associates, Inc., P.O. Box 1498, Reading, Pennsylvania 19603
and
Richard C. Bailie
West Virginia University and Environmental Energy Enginering Inc.
Morgantown, West Virginia 2757

ABSTRACT

The objective of the project is to develop a linear programming model to inform the regional energy planner of the most profitable biomass resources and conversion facilities available to satisfy specific fuel needs.

To produce this model, data will be generated on representative biomass feedstocks; market analyses of biomass resources and commercial fuel needs; and economics of thermochemical conversion facilities.

Thermochemical conversion profiles will be generated from representative biomass materials. The profiles will be developed from Thermal Gravimetric Analyzer (TGA) runs under various conditions of temperature, pressure, and particle size and in the presence of catalysts. Gasification, pyrolysis, and direct combustion will be the modes of conversion.

The biomass thermochemical conversion profiles will be used to develop biomass-to-fuels process models. These models will be verified using a Process Development Unit for a series of runs. During the initial phase of the two-year project, the linear programming model has been developed, biomass materials selected, and standard procedures for the TGA analysis developed.

INTRODUCTION

The objective of the project is to develop a linear programming model to inform the regional energy planner of the most profitable biomass resources and conversion facilities available to satisfy specific fuel needs.

To produce this model, data will be generated on representative biomass feedstocks; market analyses of biomass resources and commercial fuel needs; and economics of thermochemical conversion facilities.

Thermochemical conversion profiles will be generated for representative biomass materials. The profiles will be developed from Thermal Gravi[illegible]ric Analyzer (TGA) runs under various condi[illegible]ons of temperature, pressure, and particle size and in the presence of catalysts. Gasification, pyrolysis, and direct combustion will be the modes of conversion.

The biomass thermochemical conversion profiles will assist in developing biomass-to-fuels process models. These models will be verified using the project's Process Development Unit for a series of runs. Combinations of reactor types (fluid bed, entrained bed, and fixed bed) and gasification modes (oxygen/steam and air/steam) that can produce medium Btu gases will be preferentially selected for model development, since medium Btu gas can be further processed to yield various commercially usable liquid and gaseous fuels or chemicals.

The commercial fuels which can be derived from biomass will include: 1) for transportation - gasoline and Fischer-Tropsch liquid; 2) for the chemical industry - methanol, ammonia, and Fischer-Tropsch chemicals; 3) for residential and commercial - synthesis natural gas and Fischer-Tropsch fuel oil; 4) for industrial - low and medium Btu gas; and 5) for electric utility - combustion products, low and medium Btu gas, and methanol.

Finally, a linear programming model shall be developed for evaluating specific biomass feedstocks and the conversion processes required to supply a given volume of fuel under given market conditions. The data banks of market conditions and commercial-scale process economics will be used as input to the model.

The basic need for the project arises from the variety of available combinations of biomass feedstocks and biomass-to-fuels conversion processes available for satisfying specific fuel needs. The task of selecting the most profitable routes for satisfying these fuel needs on a regional and seasonal basis creates a complex allocation problem which can be most efficiently solved by a linear allocation modeling program and a digital computer. Linear programming is a general purpose mathematical technique for allocating resources and operating capabilities in order to obtain a definite objective.

The value of the allocation model can be explained by showing how it was used to solve a relatively simple allocation problem involving only three biomass feedstocks (corn stover, furfural residue, and wood residue) which were available to satisfy a demand for methanol fuel for power generation by combined cycle, for gasoline as a blending stock, and for chemical grade methanol [1].

EXAMPLE CONCLUSION
OPTIMUM FEEDSTOCK ALLOCATION POLICY

		Allocation of Feedstock i to Consuming Sector j, MMBtu/Hr		
		j = 1	j = 2	j = 3
i	Feedstock	Electric Utility	Transportation	Chemicals
1	Corn Stover	0	0	0
2	Furfural Residue	0	5.75	1.75
3	Wood	15.83	26.52	0

The table shows that the furfural residue alone was sufficient to satisfy the demand of chemical grade methanol. This is due to the fact that all the available, least expensive feestock (furfural residue) was used to manufacture the most expensive product (chemical grade methanol). The transportation sector requires the second most expensive product (methanol for gasohol), and any furfural residue left over after chemical grade methanol. The balance of the transportation grade methanol was supplied by wood residue, the second least expensive feedstock. All electric utility demand was satisfied by wood residue, and no corn stover was used.

This is a very simple example involving only three feedstocks and three products manufactured by a similar process. The complexity of the allocation model would be compounded as the number of feedstocks and potential products is increased. For example, if there are 20 feedstocks and 10 products, then there are up to 200 paths along which energy can flow to satisfy a specific demand. Furthermore, there are an infinite number of ways in which different amounts of energy can be allocated to these 200 paths in order to satisfy that same demand. Some allocations will yield more net profit than others, but it is not practical to select the most profitable approach without the help of a computer. Usually, that help is made available in the form of a linear programming software package.

In order to develop this comprehensive biomass allocation program, a large amount of data must be generated. Also, a mechanism must be built into the combined documentation and data storage systems which will enable for the data to be updated. The data must be generated in five key areas: market analysis, feedstock characterization, engineering economics, PDU verification, and biomass allocation model. At first, the market analysis shall develop supply and demand information about approximately 100 woody and herbaceous biomass feedstocks selected on the basis of their abundance and potential value as fuels. This analysis will be for feedstock amounts by region and season in tons/year, Btu/lb of the feedstock, and feedstock costs in \$/million Btu. It will generate data on the demands for biomass derived fuels by product and market sector and on the selling price in \$/million Btu for the fuels by region. In addition, four national regions shall be selected based on an overlapping availability of abundant biomass resources and the demand for biomass derived fuels. Biomass resources selected during the market analysis shall be characterized in a variety of thermal environments including gasification, pyrolysis and combustion. The feedstock characterization work which will take place on both a Thermal Gravimetric Analyzer and a Process Development Unit shall be used to develop process models. The models and existing data will then be used for deriving costs for manufacturing biomass derived fuels. Market analysis and the manufacturing cost data will both be used as input to the biomass allocation program. A detailed discussion of the key task areas and the progress which has been made to data follows.

MAJOR TASKS

System Allocation Program

Project Integration - Data from other tasks of the project concerning market analysis of biomass resources and biomass derived fuels as well as the engineering economics data for commercial scale conversion processes shall input the system allocation program. The program will create a mechanism for allocating most profitable biomass resources and conversion processes in order to satisfy specific fuel needs.

Approach - The computer system being used for the development of the Biomass Feedstock Allocation System is an IBM 370 Model 158 which has a model 3158-3 435 CPU with three million bytes of storage. The direct access storage subsystem (which is used for the biomass sytem data base) consists of twenty-two IBM 3330 disk storage units. The operating system currently in use is MVS (release 3.7E) with the time sharing option, and the computer programming language being used in FORTAN IV H.

In general, each projected demand for energy from biomass for a region can be satisfied in an infinite number of ways. One such allocation is shown in Fig. 1.

In order to determine whether one allocation is better than another it is necessary to employ some measure of the relative worth of each feasible energy allocation. Since the net profit expected is a generally accepted measure, it was selected for use in this project.

For the purpose of this overview, each process (for example, the process generating product 1 from feedstock 5) may be viewed as a single

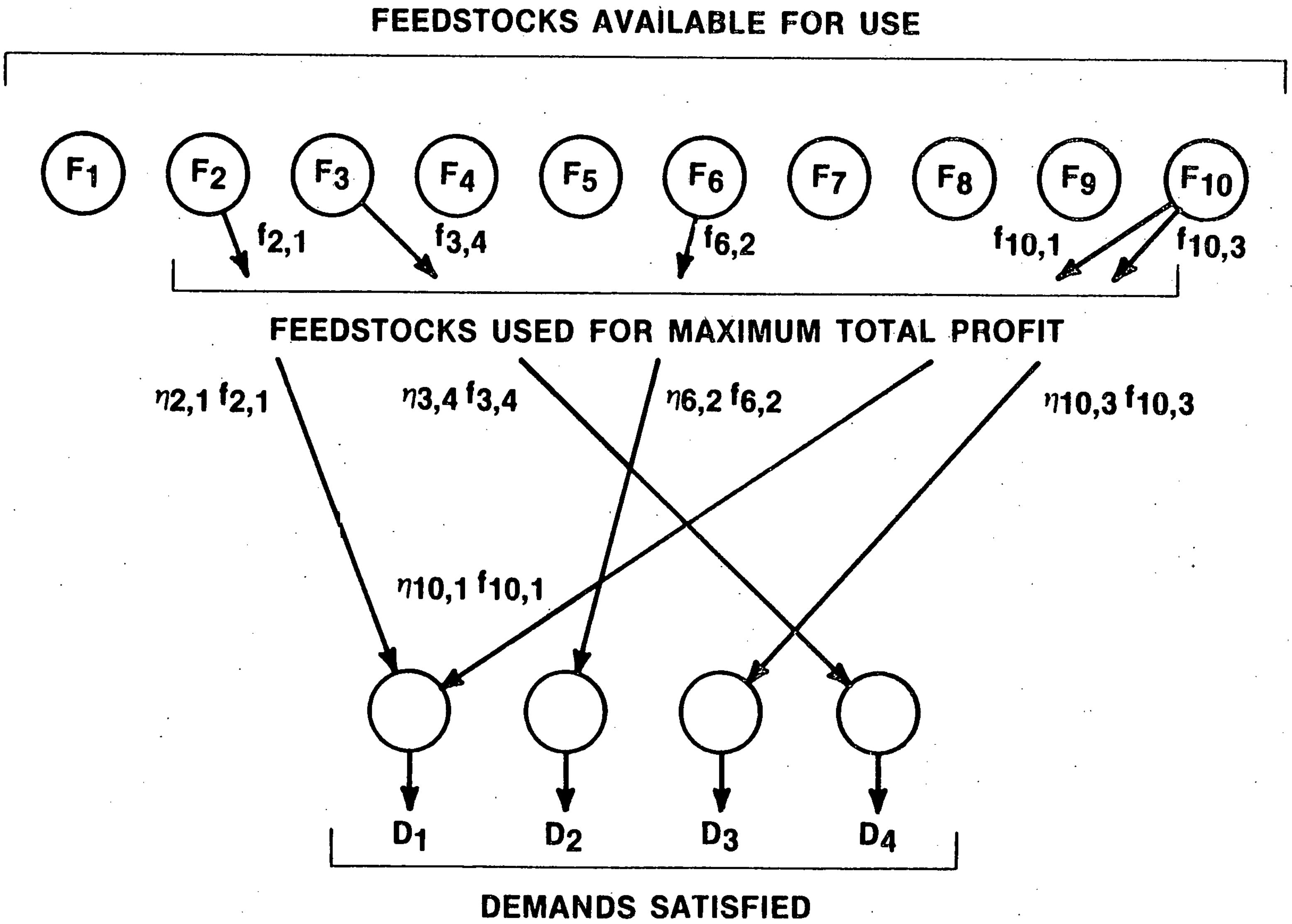

Fig. 1 A Typical Allocation

process with a single efficiency and a single manufacturing cost.

Assuming that this process has an overall efficiency of $N_{5,1}$ and manufacturing cost of $M_{5,1}$ \$/MMBtu, and that the product sells for S_1 \$/MMBtu, it follows that the profit associated with energy stream $f_{5,1}$ can be exaressed as

$$P_{5,1} = (S_1 - M_{5,1})N_{5,1}f_{5,1} \quad \$/YR$$

Similarly the profit (or loss) associated with all other energy streams in the system can be expressed in a similar manner. As a result, the total net profit for the selected system of feedstocks and energy demands can be expressed as

$$P = \sum_{j=1}^{jmax} \sum_{i=1}^{imax} (S_j - M_{ij})N_{ij}f_{ij} \qquad (1)$$

In order to proceed with a quantitative analysis of alternative distributions, it is necessary to write an energy balance for each feedstock and each demand. For a given feedstock (for example, feedstock 2) it follows that energy can flow to more than one process (or sequence of processes) but the availability of the feedstock is limited as follows:

$$F_2 \geq f_{2,1} + f_{2,2} + f_{2,3} + f_{2,4} \qquad (2)$$

Also it follows that a given demand (for example, for product 3) it follows that

$$D_3 = N_{1,3}f_{1,3} + N_{2,3}f_{2,3} + \dots + N_{6,3}f_{6,3} \quad (3)$$

Owing to the fact that equation 2 is an inequality, a slack variable (W_2), which corresponds to the amount of feedstock unused, must be added to the right hand side. When this is done for all of the feedstock equations, the result is a system of m equations in n variables.

Given such a system of m equations in n variables with n-m of those variables set to zero, it follows that a basic feasible solution can be found if the m variables occurring in the various energy balances are linearly independent and non-negative. The method usually employed to solve this type of problem is called the simplex method. This method is being used in the project.

Results and Discussion - Although an optimum profit is sought each time the biomass allocation program is executed, the actual numerical estimate of the profit is less important than the resulting allocation. The profit is merely used as a means of evaluating one distribution as opposed to another. From a users point of view the program executes the following sequence of functions:

1. For a given region and season, find the quantities of energy available for each of the following feedstocks.

   (User inputs feedstock numbers)

2. For the same region and for the following products and market sectors, retrieve the appropriate energy demand and selling price information.

   (User inputs market sector and product numbers)

3. For the user specified feedstock and demands, retrieve all efficiency and manufacturing cost information from the system data base.

   (User makes no entry)

4. Solve the resulting allocation problem.

   (User makes no entry)

5. Display the resulting allocation. At this point the program prints out the results shown below.

```
Solution Found  Feedstock   Feedstock
   Feedstock    Available     Used       Product
    Number     (MMBTU/YR)  (MMBTU/YR)    Number

      47       0.1050E+08  0.2235E+07      3
      45       0.6000E+08  0.9931E+07      2
       9       0.6000E+08  0.2108E+08      5
       3       0.3500E+08  0.3377E+08      6
```

A preliminary version of the above allocation program is presently being checked out using simulated data. During the next few months, programs to input actual data will be developed together with additional programs to output the final results.

The preliminary system will then be documented, a users manual will be written and extensive tests using both actual data and simulated data will be carried out. Further development and enhancement of the system will continue throughout the remainder of the project.

Biomass Characterization

Project Integration - The biomass characterization data is to serve as input to the process modeling and engineering economics which will be used to estimate \$/million Btu for biomass derived fuels. The characterization of biomass species according to their behavior under thermal stress is necessary to specify and design reactor systems for conversion to alternative fuel products.

Approach - The curve shown in Fig. 2 will be experimentally determined using a combination of

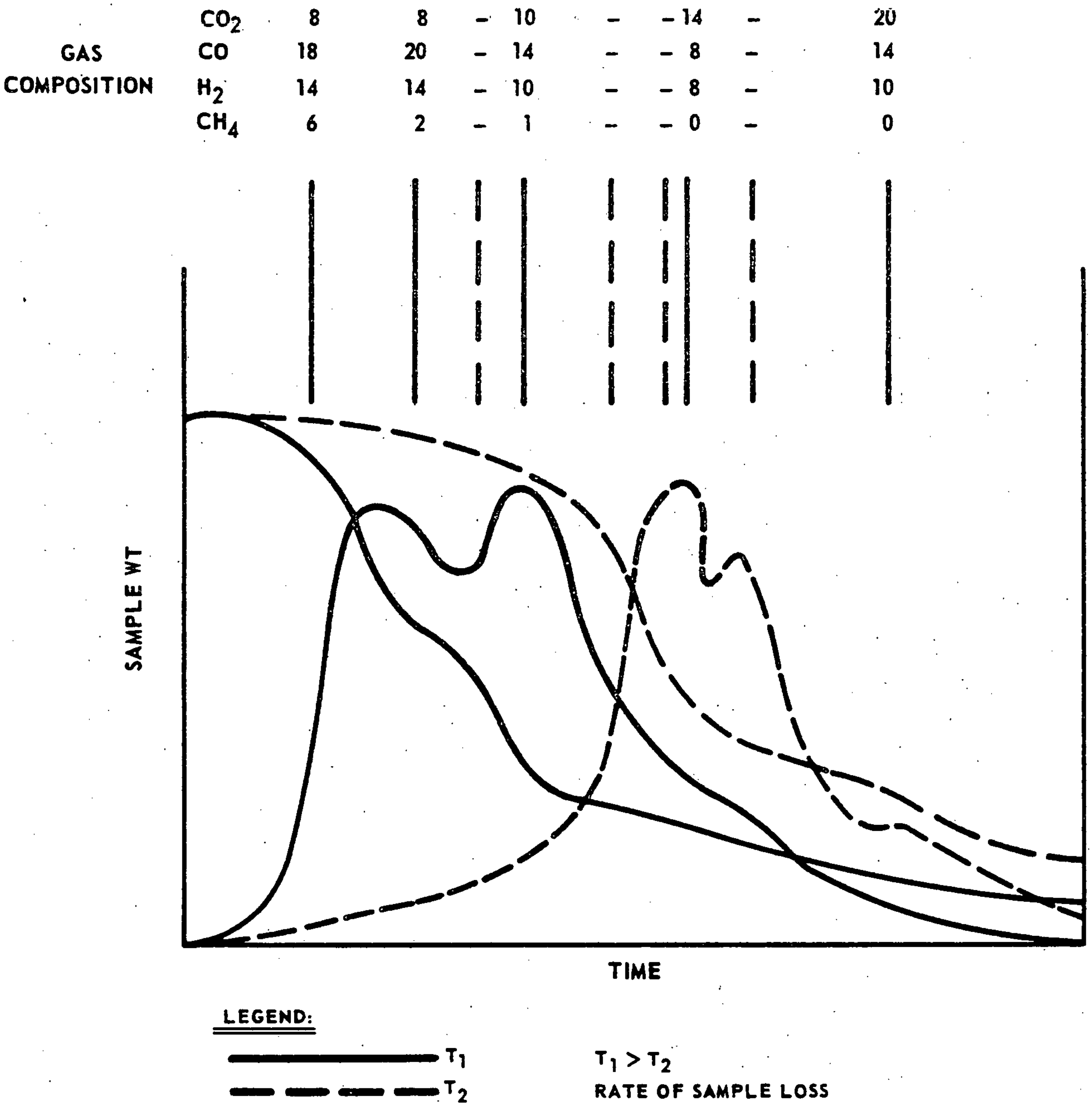

Fig. 2 Hypothetical Biomass Thermal Profile

TGA - Thermal Gravimetric Analyzer

EGA - Effluent Gas Analyzer

Two separate Thermal Gravimetric Analyzers (TGA) are used, one for atmospheric studies and one for pressurized studies. Both are of the Thompson type. A sample is prepared and lowered into the hot reactor zone, held at constant temperature. The weight and rate of weight loss is recorded. At atmospheric conditions the following variables will be studied:

Variable	Range	
Temperature	$900^{\circ}F$-$1800^{\circ}F$ (Three selected)	
Atmospheres	100% CO_2 100% N_2 100% Steam Air/Steam Air O_2/Steam 100% H_2	
Catalysts	Ni base Limestone Dolomite	Metal oxide Ash from biomass conversion

Non-catalytic pressurized studies will use the same temperature range with pressures up to 300 psi and a selection of atmospheres.

In addition to the measurement of weight loss characteristic, the gas will be analyzed for CH_4, CO_2, H_2, CO, O_2 as a function of time. This will be done by taking a series of gas samples from an exhaust gas manifold at selected time intervals during the active portion of the reaction. The gas is collected in plastic bags for subsequent analysis in a gas chromatograph.

Before any specie can be analyzed it must be collected and prepared. Once received, sample preparation involves chipping, grinding, oven drying and pelletizing.

In addition to the biomass thermal profile, the samples are characterized using the traditional moisture, proximate, ultimate and constitutent analysis.

As a preliminary screening of biomass candidates, the five listed below have been selected as prime candidates to receive early consideration.

1. White Oak
2. Eastern Hemlock
3. Loblolly Pine
4. Bagasse
5. Rice Straw

A tentative decision regarding a "standard sample" has been made. The choice was based upon preliminary runs in the TGA. Use of 0.8 g pill shaped sample, nominal dimensions in diameter and thickness was selected because: 1) it could measure weight accurately on TGA; 2) it could be easily prepared; 3) its reaction time was not too rapid to measure accurately and take gas samples, and; 4) it promotes a reaction occurring through faces rather than on edges/ends.

<u>Results and Discussion</u> - The term biomass represents a wide range of materials with distinctly different characteristics. In terms of physical appearance it is not difficult to separate trees, from corn stover, from sugar cane, from shrubs. While the physical appearance is widely different, it has been accepted in most analyses that the behavior of these varied biomass materials will be essentially the same under a high temperature environment. This similarity has not been established to be true. Since the thermal-chemical processes have as a common step the breakdown of biomass by exposure to high temperatures, it is essential to establish how various biomass materials behave when exposed to thermal stress, high pressures, different reaction gases, and catalysts. This is important for a meaningful assessment of biomass feed materials as candidates for a wide range of fuel products to be made from a variety of processes.

The data is to provide a general system of characterization that will be useful in: 1) assessment of various processes, 2) diagnosing unusual behavior of an operating system, and 3) screening potential biomass candidates for a given fuel application.

The ability to use the output from this program to predict the relative behavior of various biomass materials will be tested in a series of P.D.U. verification studies, discussed separately.

The present methods to characterize the behavior of various biomass materials include:

1. Moisture determination - This is most often reported and is ususally used to reduce the biomass yield to tons of dry biomass/ unit area.
2. Heating value - Needed for material behavior calculations.
3. Ultimate anlaysis - Needed for material behavior calculations.
4. Proximate analysis - Determines the percent of the dry biomass that is devolatized under one set of thermal conditions.
5. Constitutent analysis - Determines extractables, cellulose, hemi-cellu-cellulose, and legnin fractions.

While all of these are useful, only the proximate analysis gives any clue as to the behavior of us biomass materials when exposed to elevated temperatures. These methods of characteriziation do not provide sufficient information to select or design reactors using biomass as a feedstock.

The present studies undertake to provide additional information that is needed by characterizing the behavior of biomass according to a biomass thermal profile (BTP) which consists of:

1. The fractional weight loss of biomass as a function of time.

2. The rate of fractional weight loss as a function of time.

3. The constitutent gas generation as a function of time (gases analyzed are CH_4, CO_2, CO, O_2, H_2).

These BTP will be established for:

1. Various temperatures

2. Various reactive and non-reactive atmospheres

3. Pressures to 300 psig

4. Presence of catalysts

Fig. 2 provides two hypothetical curves at two temperatures. The curves give the weight and rate of weight loss against time. The vertical lines show the time that gas samples are taken and analyzed. The numbers shown are gas compositions identified to the left. To avoid clutter, the gas analysis for the lower temperature run is not shown.

The significant output from this study is the Biomass Thermal Profiles for a wide range of biomass materials. This work has not progressed to the stage where we can report how the various biomass materials differ when exposed to thermal stress under various atmospheres.

Engineering Economics for Commercial Scale Processes

Project Integration - The allocation model will require data on the regional manufacturing cost for biomass derived fuels in $/million Btu. Conversion efficiencies and manufacturing costs will be generated for each conversion path. This information will be developed with the aid of process models, feedstock cost data and data already in the literature.

Approach - Process Modeling: In this task, we shall be largely concerned with the problems of predicting the reaction behavior of biomass thermochemical processed by models that are simple enough to be of practical value yet [illegible]orous enough to accurately represent the [illegible]ction system. While there has been sustantial research on coal conversion reactions, little has been done for biomass conversion. Existing coal conversion models will be modified with the rate data from TGA and PDU tests.

The combination of three types of gasifiers with two types of gasification media, results in the following six process systems:

1. Entrained bed for oxygen/steam gasification

2. Entrained bed for air/steam gasification

3. Moving bed for oxygen/steam gasification

4. Moving bed for air/steam gasification

5. Fluidized bed for oxygen/steam gasification

6. Fluidized bed for air/steam gasification

The approach taken here is to examine the current status of coal conversion models; to analyze the specific characteristic behavior of biomass conversion; and to select, with necessary modification, a cost effective model capable of predicting the conversion results.

Economics: From the models and the literature data, overall material and heat balances are developed, and thermal efficiencies and economics for commercial scale plants to produce the fuels or chemicals shown in Fig. 5 are to be estimated.

The costs are developed using three levels of feedstock costs and three levels of plant capacities. Cost sensitivity curves as a function of the feedstock costs are then developed.

Results and Discussion - A search for coal computer programs which can be modified to simulate biomass conversion is underway. The computer simulation programs developed in West Virginia University have been identified and a test run will be initiated in the near future. The published data have been compiled and work is underway to select the baseline process configuration for use with the process design and economics.

Model Verification

Project Integration - The Process Development Unit will be operated in several "major" reactor configurations to verify or alter the process model previously developed.

Approach - The Process Development Unit will be modified to accommodate operation as

1. A fluid bed system
2. A falling bed system
3. An entrained bed system
4. A packed bed (moving) system

where the atmosphere in the reactor may be varied to provide

1. Air
2. Steam

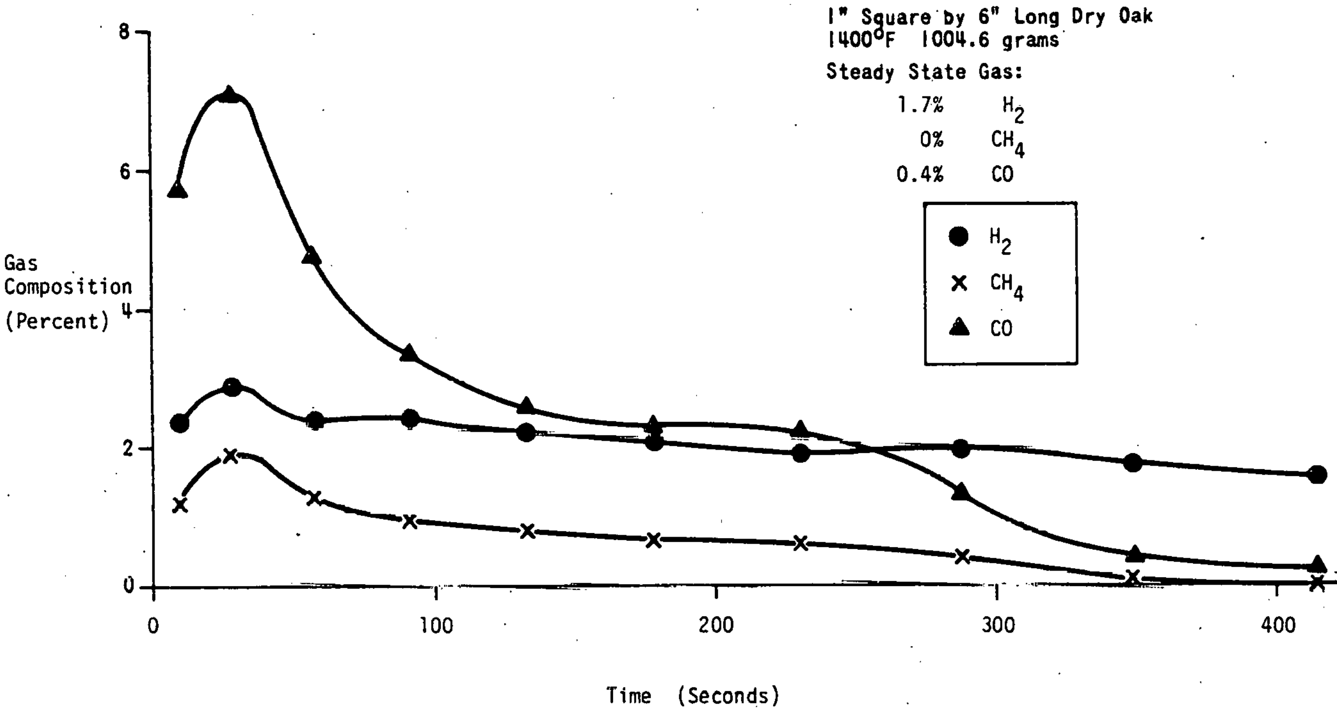

Fig. 3 Typical Gas Analysis Results

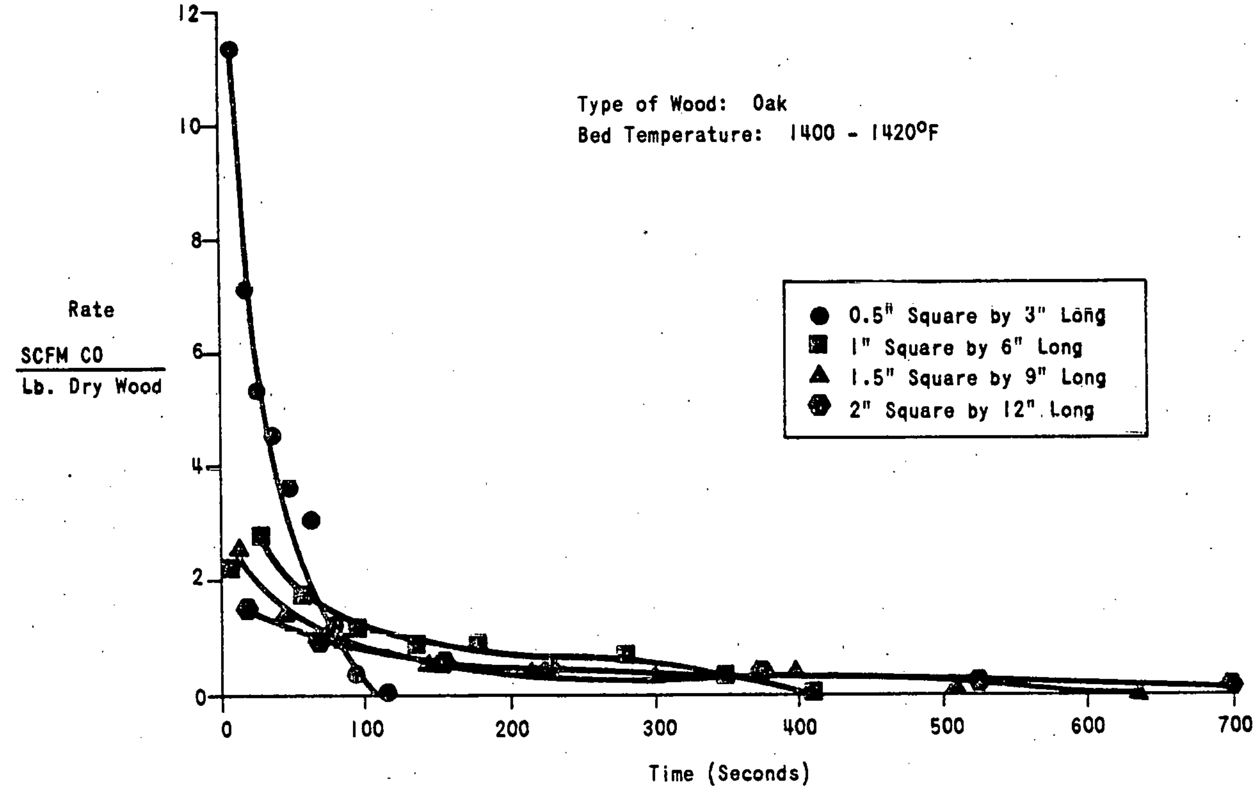

Fig. 4 Effect of Particle Size on Carbon Monoxide Production

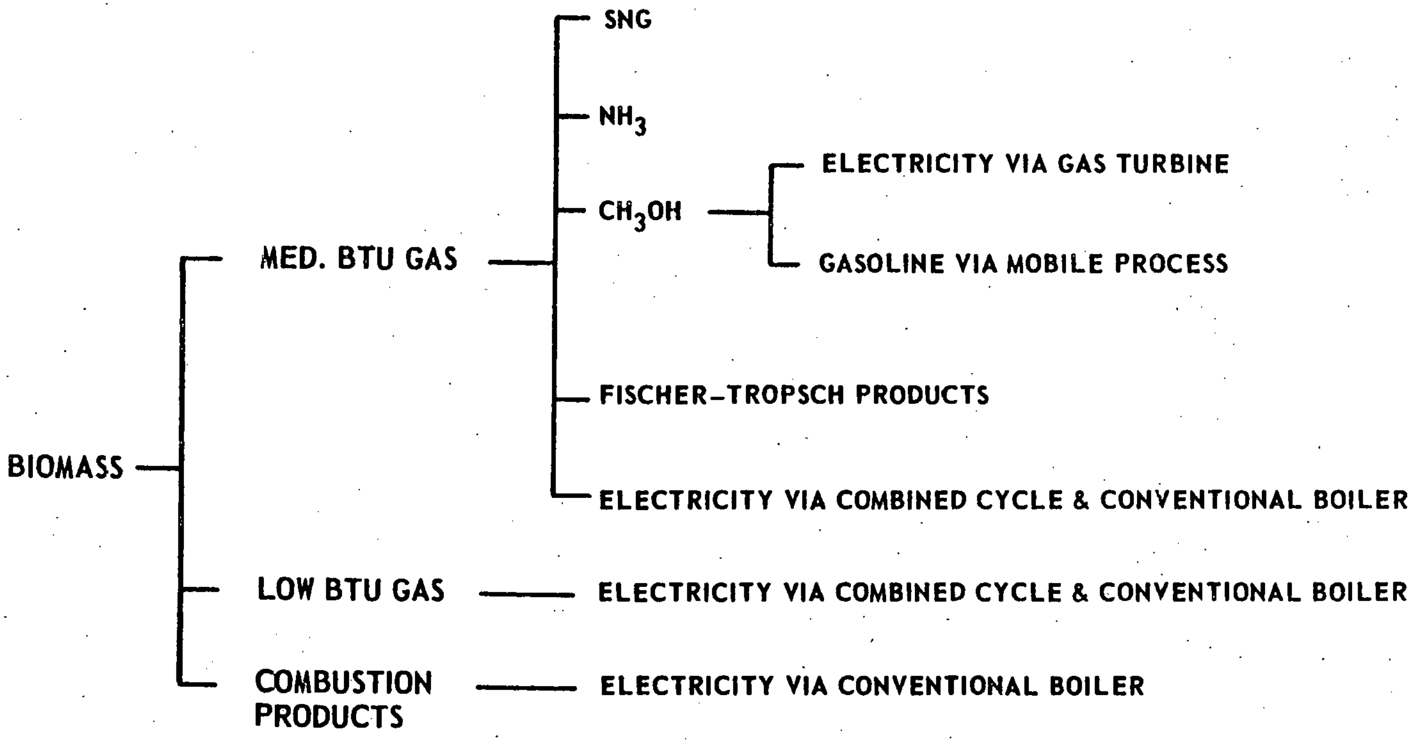

Fig. 5 Thermochemical Conversion Pathways For Fuels From Biomass

3. Steam-Oxygen
4. Steam-Air
5. Pyrolysis gas

The following series of studies and experiments will be carried out once the system has been modified.

1. Experimental verification of process models -- These will consist of runs 2-3 hrs. in duration where the PDU will be operated as

   a. Combustion Unit
   b. Pyrolysis Unit
   c. Gasification Unit

   using the following reactor types

   a. Fluid bed
   b. Entrained bed
   c. Free fall bed
   d. Packed bed (moving bed)

2. Experimental verification of process models using catalysts. These are similar to the runs made in item 1, but will include a catalyst to see of predicted gas splits are achieved. These will be done in the fluid bed mode.

3. Demonstration runs -- Based upon the first two items, some conditions will be selected for sustained runs of 24-40 hrs. where 1-2 tons of biomass are consumed.

4. In addition to the experiments shown above, the PDU will be used as a high-temperature heat source where the effect of particle size will be studied.

In all the studies, the amount and quality of gas produced will be the major output sought. The major portion of the gas data is obtained from an on-line gas chromatograph and will provide for CH_4, CO, H_2, O_2 and N_2 measurements.

The PDU will also be used to test the effect of particle size on the gas produced. The system is run as a fluid bed. The gas passing upward is adjusted to the desired temperature and contains no oxygen. This is done by adding steam to the gas produced from stoichiometric combustion of natural gas (it therefore contains significant H_2O along with N_2 and CO_2). The system is allowed to come to thermal equilibrium. A fixed amount of wood of a fixed size is then added as a pulse to the bed and the composition of gas exiting the unit is measured at varying time intervals over the course of the reaction.

Results and Discussion - The preliminary results from the tests used to evaluate the effect of particle size and gas quality and quantity are provided for

1. Oak -- representing a hard wood
2. Fir -- representing a soft wood

Some typical results are shown in Figs. 3-4 and Tables 1 and 2.

Fig. 3 -- Typical output from an experiment. Note: The reason the compositions do not return to zero is the steady-state value for H_2 and CO.

Fig. 4 -- Typical curve showing the effect of particle size at a temperature of 1400°F.

Table 1 - Shows the effect of temperature on both composition and yield for 1 inch fir blocks.

Table 2 - Shows the effect of size on both composition and yield for oak wood at 1400°F.

TABLE 1

	Compositions			Yield
	H_2	CH_4	CO	SCF/lb
1150°F	16.1	18.8	65.1	10.5
1420°F	24.4	17.7	57.9	13.8
1620°F	28.9	15.0	56.1	20.4

TABLE 2

	H_2	CH_4	CO	SCF/lb
0.5" Oak	20.5	18.2	61.3	12.6
1.0" Oak	21.4	18.4	60.2	12.8
1.5" Oak	25.3	17.7	57.1	13.1
2.0" Oak	27.7	17.5	54.8	11.8

1" Oak
Combustible Gases

	H_2	CH_4	CO	SCF/lb
1400°F	21.4	18.4	60.2	12.8
1600°F	23.7	19.9	56.3	15.8

The results shown in the Tables and Figures above are preliminary. Additional runs have been made since these reported above and they do not provide the same clear patterns shown. Further effort is necessary to substantiate the results presented.

M[illegible] Analysis

Project Integration - The market analysis data base will be used as input to the linear programming activities. The linear programs will require data for both market demands for fuels derived from biomass and feedstock supply for materials used in converting biomass to various energy forms.

Approach - The market analysis is being conducted in order to develop data bases that will include data on the regional and seasonal availability of biomass materials. Specific information obtained for each biomass will be: 1) feedstock amount by region and season in tons/year, 2) Btu/lb, and 3) feedstock cost in $/million Btu. The biomass materials include woody and herbaceous biomass species. Several of the criteria employed in selecting the materials are quantities available and proximity to user markets.

The markets for biomass derived fuels are to be identified. To date, most of the biomass resouce analyses have emphasized identification of biomass materials suitable for conversion to specific fuels and the conversion technologies available. This work will supplement these studies by examining the problem from the perspective of market demand. The question posed would be: What demands exist in the market for the fuels derived from biomass resources and what are the characteristics of these fuels? This study will identify and examine the major market issues for fuels that can be produced from biomass materials via thermochemical conversion.

The market analysis is being conducted in a discrete set of steps as follows:

- Review existing biomass market oriented studies.
- Identify a large number of biomass materials that could be priority candidates for conversion to energy forms. Collect a set of these materials for laboratory TGA analysis.
- Develop a set of regions for testing data.
- Determine the regional and seasonal supply for the biomass material selected.
- Develop a cost data base for the biomass materials to be characterized via the laboratory analysis.
- Determine demands for the biomass derived fuels for the following market sectors:
    - Transportation
    - Electric utility
    - Residential/commercial
    - Chemical and Allied products

Results and Discussion - During the first phase of the program, the market analysis project has been instrumental in assembling the following:

A list of approximately ninety biomass materials have been identified. The list will be expanded to include the major domestic woody biomass species and a number of priority agricultural materials that have large quantities of residue and could be converted to biomass derived energy forms.

A set of four regions is being developed in order to test a range of market supply and demand situations. To date, a preliminary selection of the four regions has been made. The selection criteria and data are being reviewed and these regions are being added to the study data base. Each of the regions will consist of a grouping of two to three states. There will be one region for each of the following sections of the United States: northeast, southeast, midwest, west.

The overall study will incorporate a market oriented analysis. The characterization of the fuels and the existing market infrastructures are examples of two portions which must match on the supply and demand sides respectively. This portion of the program will continue to develop characterizations of the appropriate market sectors and supply these sector data on a regional basis. These outputs will serve to provide the appropriate data to test the models under development in the study.

SUMMARY

Data are being generated in the laboratory and from the literature concerning the thermochemical behavior of biomass and the market for biomass resources and biomass derived fuels. These data will be used as inputs for a linear programming model which can be used as a tool by the energy planner to select the most profitable biomass resources and conversion facilities available to satisfy specific fuel needs.

ACKNOWLEDGEMENTS

This work is conducted under Contract Number ET-78-C-02-5022 with the Department of Energy.

REFERENCE

[1] Ahn, Y.K. "A Biomass Allocation Model - Conversion of Biomass To Methanol," A Paper Presented at the American Chemical Society Division of Petroleum Chemistry, Honolulu, Hawaii, April 1-6, 1979

NOTES

COAL GASIFICATION TECHNOLOGY: WOOD FEEDSTOCK RETROFIT POTENTIAL CONTRACT NO. ET-78-C-02-4862

Alvin G. Keene and Andrew C. Nyce
Gorham International Inc.
Gorham, Maine 04038

INTRODUCTION

Gorham International Inc. undertook a 14 month study contract on June 1, 1978 to determine which state-of-the-art and advanced coal gasifiers are technically and economically suitable for gasification of wood feedstocks to produce low, medium, and high Btu gas and synthesis gas and to specify the various modifications necessary to convert these gasifiers for use with wood residues.

The primary study tasks are to (1) critically review patents and published literature, (2) conduct 40 to 50 in-depth on-site interviews with gasifier developers, users, and manufacturers, (3) obtain information on specific gasification technologies from over 200 firms and organizations, (4) collect prior art and operating data on wood gasification, (5) estimate operating cost, capital equipment costs, gas costs, and gasification rates for coal gasifiers operating on wood feedstocks, (6) identify advantages and disadvantages of using wood feedstocks, (7) evaluate the sensitivity of capital costs and product costs to plant size, (8) determine wood type, size. shape and moisture content most suitable for specific coal gasifiers, (9) identify barriers to implementation of wood gasification plants and (10) develop implementation strategies.

Gorham will deliver the following information in the final report: (1) a narrative discussion and assessment of the thrust of coal gasification technology, (2) specifications, schematics, process flow diagrams, narrative descriptions, and wood gasification rates for coal gasifiers modified for use with wood feedstocks, (3) specifications, schematics, and process flow diagrams covering the specific modifications required to convert coal gasifiers for use with wood feedstocks, (4) tabular and graphical presentation of comparative capital equipment costs, operating costs, and costs ($/mm Btu's) for low. medium, and high Btu gas and synthesis gas produced from coal and from wood feedstocks, (5) recommendations for wood feedstocks and wood feedstock preparation and handling equipment, (6) comparative costs for the materials of construction for coal and wood gasifiers, (7) wood gasifier implementation plan, (8) barriers to implementation plan and (9) alternative strategies to overcome these barriers.

RESULTS TO DATE

Gorham International has completed approximately [illegible] percent of the contract program designed to [illegible]ermine the feasibility of retrofitting coal gasifiers for use with wood residues.

The program efforts completed to date include: (1) critical review of U.S. and foreign patents and published literature related to gasification of carbonaceous materials, (2) on-site interviews with over 40 gasifier developers users and manufacturers, (3) telephone interviews with over 75 firms active in coal and biomass gasification technology development and commercial operations, (4) identification of the primary advantages and disadvantages relative to utilization of wood feedstocks in fixed bed, fluidized bed, entrained flow and molten bath coal gasifiers, and (5) establishment of retrofit assessment criteria.

This report presents the basic descriptions and categories of coal gasification technologies included in the study program. Each technology is being evaluated for wood retrofit potential according to the following criteria:

Technical Retrofit Assessment Criteria

- Ability to accept wood feedstocks
- Gas output (Btu/hr)
- Gasification thermal efficiency
- Primary design application
- Product gas composition
- Materials of construction
- Environmental controls

Economic Retrofit Assessment Criteria

- Capital equipment cost
- Wood feedstock costs ($/MM Btu)
- Plant operating costs
- Economies of scale
- Ultimate gas cost ($/MM Btu)
- Development costs for commercialization

Technical Retrofit Assessment Criteria

Ability to Accept Wood Feedstock. Ideally the gasifier should accept state-of-the-art wood feedstock such as chips, pellets, sawdust, flour, etc. without the need for redesign of the feed system or other major parts of the overall gasifier system.

Replacement of the feed system or other parts of the gasifier system with state-of-the-art components or systems to enable retrofit with wood is considered acceptable.

Gas Output. There should be no significant derating of the gasifier when wood is used as a feedstock. For the same size gasifier, the Btu per hour of product gas should be equal to or greater than when the gasifier is operated on coal, lignite, or peat.

Gasification Thermal Efficiency. The overall thermal efficiency of the gasifier defined by the total energy contained in the wood feedstock divided into the total usable energy output should be equal or greater than when the gasifier is operating on coal, lignite or peat. The total energy equation would include the energy required in wood preparation, steam raising, electricity to operate various equipment components, etc.

Primary Design Application. Each gasifier will be evaluated for its potential to be operated with wood for specific end use applications such as (1) industrial on-site or over-the-fence industrial fuel gas, (2) gas turbine combined cycle power generation, (3) petrochemical feedstock. Some coal gasifiers have been designed for highly specific applications and for specific types of problem coals and must be carefully evaluated to determine if optimum operation can be obtained with wood feedstocks.

Product Gas Composition. The compositions of the gas including particulates should not provide any problems with respect to operation of the gasification plant and clean-up system nor with respect to specific end use of the gas. For example, if the gas is to be used to produce methanol, then the CO/H_2 ratio should be within acceptable limits and the CO_2 content should not be so high that special gas clean up problems are encountered.

Materials of Construction. State-of-the-art gasifiers appear to have no particular materials problem with coal and are expected to operate equally as well using wood as a feedstock. On the other hand, the advanced and second generation coal gasifiers have a host of serious materials problems directly related to the chemical and physical make up of coal.

State-of-the-art materials and construction techniques must be applicable to coal gasifiers retrofitted for use with wood.

Environmental Controls. The processing plants and systems for particulate, gaseous and aqueous clean up must be state-of-the-art systems which have been tried and proven either coal or wood gasification, direct combustion systems or other industrial applications.

Economic Retrofit Assessment Criteria

Capital Equipment Cost. The capital equipment cost of the gasifier, wood preparation, oxygen plant and environmental control system should not exceed the cost for a comparably sized coal gasification complex with the same Btu/hour output. There should be no penalties for using wood as the feedstock. The cost of harvesting, transporting and preparing the wood will be considered separately.

Wood Feedstock Costs ($/MM Btu). The cost of wood feedstocks on a $/MM Btu basis ready for feeding to a coal gasifier should not exceed the corresponding cost for coal. This cost will depend upon whether the gasifier is being operated over-the-fence or at great distance and upon whether the gasification plant is located in the center of the wood supply. For example, it would make a great deal of sense to locate a gasifier/power generation complex in the woods to lower the cost of transportation. This concept would require cooperation of the utilities with respect to using their power transmission lines. In the case of medium Btu gas, it can be mixed with high Btu gas and transported hundreds of miles from wooded to non-wooded regions, thereby decreasing the cost of transporting wood long distances.

Plant Operating Costs. The cost of operating the gasification plant on wood should be less than or equal to the operating costs of a coal plant. The labor force and other operating expenses should be equivalent in a wood or coal gasification plant.

Economies of Scale. There must not be economic penalties for scale down of the gasifier output. Some advanced single gasifiers or multi-gasifier plants require 5000 tons/day for economic operation. It appears that 1000 oven dry TPD of wood is the largest practical sized gasifier plant which could be kept supplied with wood on a 365 day per year basis.

Ultimate Gas Cost ($MM/Btu). The cost per million Btu's of low and medium Btu gas should be the same or lower for wood feedstocks compared to coal. The feedstock costs of gathering, transporting and preparation should not exceed those for coal unless other offsetting cost decreases are realized with wood feedstocks.

Development Costs for Commercialization. In the case of advanced and second generation coal gasifiers or state-of-the-art gasifiers requiring major redesign, the cost of bringing a specific gasifier to commercial status should be consistant with DOE guidelines and projected budgets for research, development and demonstration programs.

Coal Gasification Technology Categories

The following brief listing of specific coal gasification technologies presents the categories and systems included to date in the study program.

Each of these technologies has been evaluated and assigned a preliminary rating for its wood retrofit potential. Currently, these preliminary assessments are being reviewed by system developers, operators and others active in the gasification field. Based on the preliminary nature of current results and the ongoing review, no quantitative assessment results are included in this paper.

The results of the final wood retrofit potential rating will be utilized to select those coal gasification technologies with the greatest probabili[illegible] of commercially utilizing wood feedstocks. The selected systems are undergoing more detailed re [illegible] it analysis as discussed briefly in the remaining sections of this paper.

	TPD	(MTPD)
Fixed Bed Gasifiers		
Commercial:		
- Wilputte	30	(27.2)
- I.F.E./Swindel Dressler	60	(54.4)
- Riley Morgan/Riley Stoker	60	(54.4)
- Wellman-Galusha/McDowell-Wellman	60	(54.4)
- I.G.I./Bobcock Contractors	100	(90.7)
- Lurgi	800	(725.6)
Pilot Scale:		
- DOE/Merc Stirred Bed	20	(18.1)
- Gegas/General Electric	24	(21.8)
- British Lurgi Slagging	80	(72.6)
Entrained Flow Gasifiers		
Commercial or Semi-Commercial:		
- Babcock & Wilcox (Slagging)	400	(362.8)
- Koppers-Totzek (Slagging)	860	(780.0)
- Ruhrgas Vortex/Ruhrgas A.G.	NA	NA
Pilot Scale:		
- Bell Aerospace Textron	12	(10.9)
- Flash Hydropyrolysis/Rockwell International	14.4	(13.1)
- Texaco (Slagging)	15	(13.6)
- Combustion Engineering	120	(108.8)
- Bi-Gas/Bituminous Coal Research Inc. (Slagging)	150	(136.1)
- Foster Wheeler (Slagging)	480	(435.4)
Fluidized Bed Gasifiers		
Commercial:		
- Winkler	1100	(997.7)
Pilot Scale (over 25 TPD):		
- Union Carbide/Battelle	25	(22.7)
- CO_2 Acceptor/Conoco	30	(27.2)
- Hydrocarbon Research	30	(27.2)
- Hygas/IGT	70	(63.5)
- Synthane/Perc	72	(65.3)
- Cogas	80	(72.6)
Pilot Scale:		
- Bituminous Coal Research	1.2	(1.1)
- Exxon Gasification	5	(4.5)
- U-Gas/IGT	6	(5.4)
- Westinghouse	12	(10.9)
Molten Bed Gasifiers		
Pilot Scale:		
- Atomics International-Molten Salt	24	(21.8)
Commercial:		
- Otto-Rummel Slag Bath	264	(239.4)

Gasification Technology Performance Evaluation.

[illegible]d on the previously reported practical analyses those gasification technologies judged to have the best retrofit potential for wood feedstocks are being evaluated for operational performance.

Gasification technology selection for this analysis is based primarily on (1) manufacturer/developer interest in operation with wood fuel, (2) apparent retrofit difficulty and associated costs, (3) feedstock form and size characteristics, (4) scale of operation, and (5) heating value of product gas. The performance of selected gasifiers will then be compared using simple stoichiometric and equilibrium calculations that have been found suitable for coal gasification. These analyses methods will consider the effects of wood moisture and reactivity, gasifier type (fixed, fluidized and entrained beds) and carbon efficiency. Predicted parameters such as (1) throughput, (2) product heating value, (3) cold gas efficiency and (4) hot gas efficiency should provide a suitable basis for comparing the potential of those gasification technologies most adaptable to utilization of wood feedstocks. The cold gas efficiency is a reflection of the chemical enthalpy of the product gas; while the hot gas efficiency considers both the chemical and sensible enthalpy of the product gas. The predicted throughputs and heating values will enable identification of industrial process applications amenable to utilization of specific retrofitted coal gasification technologies. The cold and hot gas efficiencies will identify those gasification systems most suitable for production of medium and low Btu gases, respectively.

The proposed calculations will proceed in the following manner:
(1) Determination of the oxider/fuel ratio corresponding to the maximum theoretical cold gas efficiency.
(2) Determination of the maximum throughput for a given carbon efficiency.
(3) Determination of the gasifier temperature via an energy balance.
(4) Determination of the product mole fractions by assuming water gas equilibrium at the calculated gasifier temperature.
(5) Determination of the product gas heating value
(6) Determination of the coal and hot gas efficiencies.

In these calculations, the effect of moisture content and the possibility of operation without steam addition (particularly for low Btu gas) will be considered. Also, to facilitate comparisons with prior analyses of biomass and solid waste gasification systems using wood feedstocks, the raw wood feedstock will be a fifty percent moisture "standard wood" as defined by SRI International.

Upon completion of the above analysis, a preliminary implementation program and schedule will be developed for each gasification technology determined to have significant potential for retrofit with wood feedstocks. The implementation plan will cover necessary technological and nontechnological efforts required for the selected systems to progress from their current state-of-the-art to commercialization using wood feedstocks. In addition to the standard technical parameters related to actual system operation, a preliminary assessment of financial risk, political, institutional, regulatory, organizational and governmental requirements and potential barriers will be carried out for specific gasification systems. The ultimate result of the above discussed evaluation procedure will be a projected implementation plan and program schedule for the coal gasification technologies with the best retrofit potential with wood feedstocks.

Primary Coal Gasification Technology Applications

Coal gasification technologies have been developed and designed for primary usage in specific applications. Thus, the process applications considered in the original system development have influenced the resultant mode of operation. The four basic operational modes generally considered are (1) atmospheric air blown, (2) pressurized air blown, (3) atmospheric oxygen blown, and (4) pressurized oxygen blown. Normally, product gas from air blown systems are classified as low Btu while oxygen blown systems are labeled medium Btu.

The use of wood feedstocks in coal gasification systems result in a fuel product similar to that resulting from coal. The following comparison of air and oxygen blown systems using wood and coal as feedstocks indicate the resultant product compositions:

Air Blown Producer Gas Composition
Wood versus Coal

	(% of Product Gas)	
Component	Coal	Wood
CO	30	26
H_2	15	14
CH_4	2	2
CO_2	3	9
N_2	50	49

Oxygen Blown Producer Gas Composition
Wood versus Coal

	(% of Product Gas)	
Component	Coal	Wood
CO	38	45
H_2	38	35
CH_4	3.5	0.35
CO_2	18	19.5
N_2	2.5	0.15

Based on the published and stated design objectives and applications data, the following is a summary of primary applications for the more prominent coal gasification technologies. The primary design applications represent usage with little or no modification. Based on the similarity of gas compositions shown above, units retrofitted to use wood feedstocks should be applicable for use in the primary design applications associated with coal feedstocks.

Gasifier Type and Developer/Marketer	Primary Design Application(s)
Fixed Bed	
Lurgi-Dry Ash	Synthesis Gas (Chemicals & SNG) Combined Cycle Power Generation
Lurgi - Slagging	Synthesis Gas (Chemicals & SNG) Large Scale Fuel Gas
GEGAS/General Electric	Combined Cycle Power Generation
Riley Morgan/Riley Stoker	Small Scale Fuel Gas
IFE/Swindell Dresser	Small Scale Fuel Gas
Wellman Galusha/ McDowell-Wellman	Small Scale Fuel Gas
Wilputte	Small Scale Fuel Gas
IGI/Babcock Contractors	Small Scale Fuel Gas
MERC Stirred Bed/DOE	Small Scale Fuel Gas
Fluidized Bed	
Union Carbide/Battelle	Synthesis Gas-Chemicals
Westinghouse	Combined Cycle Power Generation
Exxon	Synthesis Gas
Hydrocarbon Research	Small Scale Fuel Gas
Winkler/Davy Powergas	Combined Cycle Power Generation
HYGAS/IGT	Synthesis Gas - SNG
U-Gas/IGT	Combined Cycle Power Generation
BCR-Low Btu	As Above
CO_2 Acceptor/Conco	Synthesis Gas - SNG
COGAS/Cogas Dev. Co.	Synthesis Gas - SNG
Synthane/PERC	Synthesis Gas - SNG
Entrained Flow	
Flash Hydropyrolysis/ Rockwell International	High Btu Gas
BI-GAS/Bituminous Coal Research	Synthesis Gas - SNG
Combustion Engineering	Large Scale Fuel Gas
Ruhrgas Vortex	Low Btu Fuel Gas
Foster Wheeler	Combined Cycle Power Generation

Gasifier Type and Developer/Marketer	Primary Design Application(s)
Entrained Flow Continued	
Koppers-Totzek	Synthesis Gas-Chemicals
Texaco	Synthesis Gas-Chemicals
Bell Aerospace Textron	Low Btu Gas
Babcock & Wilcox	Synthesis Gas-Chemicals

While many of the above gasification technologies have been proposed for uses other than the primary design objectives, the above indicated initial design applications are factors which must be considered when making a final selection of coal gasification technologies for potential use with wood gasification. Thus, the anticipated and target end uses for wood gasification systems should be selected and marketed with the appropriate coal gasification systems' primary design applications to maximize the potential for retrofit success.

Implementation Criteria

Prior to commercialization, retrofitted coal gasification technologies will face several hurdles and potential barriers to implementation. Thus, this contract includes an analysis of potential implementation barriers, formulation of strategies to overcome these barriers and ultimate development of a systematic plan of commercialization which will address key barriers and result in a time and cost effective commercialization route.

Many of the variables which must be reviewed and addressed prior to commercialization of retrofitted coal gasifiers are similar to those for standard coal or biomass gasification. Thus, the approach to development of this implementation plan will include a critique of both coal and biomass gasification programs and attempt to select complementary strategy development utilizing the optimum mix of current and proposed approaches for both coal and biomass.

At this point, only preliminary efforts have been conducted in the implementation planning area with concentrated efforts scheduled upon selection of coal gasification systems and of primary end use technology applications.

The principal items of concern relative to implementation barriers can be grouped into four broad categories as follows:

(1) Technical Risk
(2) Economics
(3) Institutional
(4) Environmental

(1) Technical Risk. The technical risk associated with all alternative energy sources is a significant barrier to implementation and is a major factor in the increasing dependence on foreign energy [illegible]plies. However, wood retrofitted coal gasifi[illegible]ion technologies may be able to capitalize on conventional coal gasification progress and overcome this factor more rapidly than biomass systems developed from ground zero. One factor favoring wood retrofitted coal gasifiers is the use of a generally favorable feedstock in systems already proven on a more difficult feedstock. While we do not envision retrofitted systems becoming commercial prior to the original coal units, they could be commercialized in a parallel program provided proper development and implementation programs are formulated and instituted in a timely manner.

In addition to the technical risks associated with gas production, synthetic gas utilization represents considerable uncertainty and concern to potential gas users.

Thus, specific technological programs to demonstrate the routine long term operability of alternative energy production systems of all types are essential to overcome technical risk barriers.

(2) Economics. As with any new technology, economic feasibility is an essential element prior to capital fund commitments. Currently, the major deterrent to implementation of alternative energy sources are production of required energy at a competitive price. While the desire for energy independence and long term cost projections can be used to illustrate the desirability of a "construct now" philosophy, the current higher cost per unit of energy makes it difficult, if not impossible, to obtain industrial commitments of required capital funds.

Based on the uncertainty of future energy pricing and the unstable national energy policy, solid economic projections are difficult to make. However, programs designed to define and eliminate technical risk can aid in development of comparative economics by providing more solid design/operational data and by enabling complete project design based on hard data.

(3) Institutional. The institutional factors relative to the integration of new energy production sources into the conventional energy picture represent a complex picture. While some traditional sales have started to bend slightly, the rigid, bureaucratic regulatory agencies and utility companies move very slowly and are careful to protect their operational domains. Relationships between buyers and sellers coupled with regulatory controls are an area of much study and proposed new relationships which will play a major role in the speed of new energy source commercialization over technical and economic barriers to be overcome. Thus, preliminary attention and preplanning are essential if these factors are not to cause inordinate delays once energy suppliers/users are satisfied that technical and economic feasibility has been established.

(4) Environmental. Any process oriented industry is faced with environmental concerns and energy production via any means is no exception. While technical and economic concerns are caused by regional environmental controls, these must be considered as a part of the overall development and

implementation process to avoid surprises at the commercialization stage. The primary impact of environmental requirements will be the increased cost of installing and operating facilities required to meet standards. Probably the most difficult environmental related problem to deal with is the public involvement in site selection and approval. Again, this is now a way of life and proper preparation and education programs are essential to cope with the public concern.

While the above brief discussions only present the gross problems and not solutions, they highlight the major classes of barriers which will be addressed in development of the implementation plan and strategy for wood retrofitted coal gasification technologies during the latter phase of this contract program.

SUMMARY

The project schedule now calls for submittal of a draft final report by July 31, 1979. The major task efforts underway which will result in development of the originally stated program deliverables include:

(1) Final selection of coal gasification systems with greatest potential for retrofit with wood feedstocks.

(2) Gasification system performance evaluation (operational evaluation based on preliminary kinetic assessment).

(3) Techno-economic analyses of selected gasification systems based on results of item 2 and selected technology applications.

(4) Development of wood retrofitted coal gasification technology implementation plan.

(5) Identification of barriers to implementation and development of strategies to overcome identified barriers.

This project's results will enable the Biomass Energy Systems Branch to effectively utilize the results of research, development, demonstration and commercial coal gasification programs. The potential for related technology transfer and applications to aid in the utilization of biomass to provide an increased portion of our national energy demand will be defined in a manner to facilitate real world implementation programs.

BIOMASS BASED METHANOL PROCESSES

Edward I. Wan
John A. Simmons
Joseph D. Price
Tien D. Nguyen
Science Applications, Inc.
1764 Old Meadow Lane
McLean, Virginia 22102

ABSTRACT

This work involves a comprehensive research program to evaluate and develop biomass to methanol processes. The three major components of the biomass to methanol system are the availability of biomass feedstocks, the thermochemical conversion of biomass to methanol fuels, and the distribution and market analysis of methanol fuels.

Initial efforts were undertaken to identify and evaluate the regional potential of biomass resources and to assess the market for methanol fuels. The near-term and long-term biomass availability and methanol supply and demand data were synthesized in an effort to prioritize the suitable methanol production areas in the country. In the conversion process analysis, detailed evaluation of each major processing step was attempted. Three commercially available oxygen-blown gasification processes were analyzed and compared in terms of their feasibilities for producing methanol synthesis gas from biomass. These processes include the fixed bed Wellman-Galusha and Purox gasifiers, and the entrained-bed Koppers-Totzek gasifier. Various synthesis gas conditioning and cleaning processes were also evaluated and selected according to the specific gasification application. Two commercially available methanol synthesis processes plus one advanced liquid-phase methanol system were also analyzed. Empirical and theoretical process models were developed so that parametric analysis and economic sensitivity can be determined. Based on these analyses, an optimized process design based on 1000 tpd biomass throughput has been derived. Detailed material and energy information of this base design were used for economic sensitivity analysis.

The results of this study have indicated that currently the cost of methanol production from biomass ranges from 60 to 90 cents per gallon depending upon regional feedstock costs, the conversion plant size, the gasification process employed, and overall process energy integration and optimization. In order to achieve higher conversion efficiency and competitive economics, several alternative biomass to methanol production concepts were formulated. This involved the rporation of an advanced biomass gasification ept, simplification of synthesis gas modification processes, optimization of methanol synthesis, and the adaption of hybrid synthesis gas schemes.

INTRODUCTION

Science Applications, Inc. (SAI) is conducting a comprehensive engineering system study to assess various thermochemical processes suitable for converting biomass to methanol and other alcohol fuels. The overall objectives of this program are:

- Develop alternative thermochemical processes for the production and introduction of methanol and other alcohol fuels from biomass.
- Develop a systems model to assess the economic feasibility of the production and marketing of alternative alcohol fuels.
- Assess the technical and economic feasibility of biomass based alcohol as a fuel and as a chemical feedstock.
- Develop recommendations for RD&D plan to minimize the economic impacts of the introduction of biomass based alcohol as a new fuel into our national economy.

During the past few years, much interest has been generated in the possibility of producing alcohol fuels from renewable biomass resources. As a result, many new and improved biomass gasification concepts and catalytic conversion technologies have emerged. These new technologies may play a significant role in improving the methanol production economics, and can also provide conversion process alternatives for the production of higher alcohols and gasoline products. Thus, the development and integration of new conversion processes could be one of the determining factors in the economic production of chemicals and fuels from biomass.

In view of the need for transfer and integration of the current and new technologies to biomass applications, SAI has proposed a three-phase thermochemical alcohol fuel research program. These are:

Phase I - Technical and economic feasibility study of current technology for the production of methanol from biomass;

Phase II - Conceptual process design and economic assessment of advanced and innovative thermochemical conversion processes for producing alcohol fuels and gasoline from biomass;

Phase III - Pilot plant studies and process demonstration of commercially feasible advanced alcohol fuels and gasoline production processes.

The Phase I program, the current biomass to methanol program, is directed toward the assessment of the application of conventional technology to methanol production. The remaining paper will present a summary of results of our Phase I program delineating the technical and economic feasibilities of producing methanol fuel from biomass resources utilizing the currently available technologies.

The system approaches and analysis steps utilized in the Phase I program in terms of three specific topic areas are outlined as follows:

Biomass Resources Analysis

- Characterization of regional biomass resources.
- Projection of near-term and long-term biomass resources.
- Establish criteria for screening potential biomass production areas.

Methanol Fuel Market Analysis

- Evaluate existing and potential methanol end uses.
- Methanol market penetration analysis.
- Comparison of potential methanol demand and supply.

Conversion Process Analysis

- Assessment of the State-of-the-Art conversion technology.
- Development of alternative conceptual process configurations.
- Comparison of conceptual biomass to methanol processes.

The significant results of the analysis of the biomass based methanol processes are summarized in the following:(1)(2)

Biomass Resource Analysis

- In the near term (before 1990), the major source of biomass will be agricultural and silvicultural residues (Table 1). This represents approximately 423 million dry tons of biomass per year (equivalent to 7 quad of energy) in the United States. The West North-Central region will have the largest available quantity, followed by the South Atlantic region, and then the East North-Central region.
- In the long term (after 1990), additional biomass could be available from biomass energy farms, where total national production could reach 265 million dry tons per year (Table 1). The regions which have the greatest potential for production from energy farms are the south and southeast regions of the United States.
- The examination of residue resource characteristics has shown 27 suitable methanol conversion sites (Table 2). These sites have different resource availabilities ranging from 560 to 12000 ODT/day capacity. Similarly, the 10 representative methanol sites based on energy farm resources are shown in Table 3.
- The estimated delivered costs for biomass feedstocks suitable for methanol conversion range from 1.7 to 4.5 \$/MMBtu (29 to 71 \$/dry ton) for residue type biomass and from 1.4 to 2.2 \$/MMBtu (24 to 37 \$/dry ton) for energy farm type biomass. (Fig. 1).

Methanol Market Analyses

- Methanol consumption in the U.S. totaled 1150 million gallons in 1975. Methanol requirements for traditional chemical uses are estimated to grow at annual rates of 7.7 percent during the years 1970-1980 and at 5.4 percent during the years 1980-2000.
- Estimates of potential future markets for methanol vary, but most indicate a high potential demand. By 1985, demand for methanol in applications such as peak power generation, automotive fuel, single cell protein, steel manufacture and sewage treatment could be ten times as large as the demand for methanol in the manufacture of establsihed chemicals.
- The Mid-Atlantic, South Atlantic, East North-Central and West South-Central regions are expected to gain the largest share of the methanol market between 1985-2000. If methanol is blended to 10 percent in gasoline, a total market of 13.44 billion gallons per year is projected by 1985. Other uses for methanol as a fuel would result in a still greater demand (see Table 4).

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(1) "Evaluation of the State-of-the-Art of Biomass Based Methanol Process", Progress Report No.1, August 14, 1978, Science Applications, Inc.

(2) "Development of Biomass to Methanol System Algorithms and Models", Progress Report No.2, March 25, 1979.

Table 1

PROJECTED REGIONAL BIOMASS RESOURCE AVAILABILITY

	Existing Resource Base							Potential Biomass Farms			Total Projected		
	Forestry	Agricultural			Total								
	1985 - 2000	1985	1990	2000	1985	1990	2000	1985	1990	2000	1985	1990	2000
NE	14.1	0.3	0.3	0.3	14.4	14.4	14.4	3.5	7.0	14.0	17.9	21.4	28.4
MATL	16.8	3.3	3.4	3.5	20.1	19.4	20.3	6.5	13.0	26.0	26.6	32.4	46.3
SATL	54.3	10.0	10.2	10.8	64.3	64.5	65.1	35.0	70.0	126.0	99.3	134.5	191.1
ENC	21.9	37.8	38.6	40.2	59.7	60.5	62.1	16.2	32.4	64.8	75.9	92.9	126.9
ESC	39.6	6.5	6.6	6.9	46.1	46.2	46.5	18.9	37.8	71.4	65.0	84.0	117.9
WNC	14.1	97.0	99.0	103.0	111.1	113.1	117.1	13.2	26.4	52.8	124.3	139.5	169.9
WSC	24.5	23.0	23.5	24.4	47.5	48.0	48.9	36.0	72.0	122.4	83.5	120.0	171.3
MT	14.0	21.0	21.4	22.9	35.0	35.4	36.9	0	0	0	35.0	35.4	36.9
PAC	4.3	20.6	21.0	21.9	24.9	25.3	26.2	3.5	7.0	14.0	28.4	32.3	40.2
Total U.S.	203.6	219.5	224.2	233.9	423.1	426.8	437.5	132.8	265.6	491.4	555.9	692.4	928.9
MM Acres	500	400	400	400	900	900	900	16.2	32.4	32.4	916	932	932
Percent of U.S. Land Area	22	18	18	18	40	40	40	0.7	1.4	1.4	40	41	41
Equivalent(1) Energy (10^{15} Btu)	3.5	3.7	3.8	4.0	7.2	7.2	7.5	2.2	4.5	8.3	9.5	11.7	15.8

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(1) Based on 17 million Btu per dry ton of Biomass

Table 2: Representative Methanol Conversion Sites Based on Biomass Residue Availability and Costs

Site #	State	Plant Location(1)	Percent(2)		ODT/Day		Typical Transportation (Miles)	Delivered Cost(3)	
			Crop	Forestry	100%	25%		(1980$/ODT)	($/MMBtu)
1	Maine	Piscataquis Co.	18	82	3300	800	60	42(3)	2.5
2	Illinois	Woodford Co.	100	0	49000	12000	80	60	3.5
3	Indiana	Delaware Co.	99	1	7600	1900	45	56	3.3
4	Ohio	Wood Co.	98	2	6300	1600	40	41	2.4
5	Wisconsin	Dane Co.	99	1	5300	1300	35	56	3.3
6	Iowa	Iowa Co.	100	0	12000	3000	35	63	3.7
7		Palo Alto Co.	100	0	34000	8500	75	68	4.0
8	Kansas	Gray Co.	100	0	16000	4000	55	56	3.3
9		Barber Co.	100	0	19000	4900	65	60	3.5
10	Minnesota	Clay Co.	100	0	10000	2500	50	57	3.4
11		Stearns Co.	100	0	5500	1400	45	56	3.3
12	North Dakota	Sargent Co.	100	0	8500	2100	45	56	3.3
13		Stutsman Co.	100	0	10500	2600	40	55	3.3
14		Ward Co.	100	0	11600	2900	45	56	3.3
15		Hettinger Co.	100	0	3900	1000	25	53	3.1
16	Florida	Hendry Co.	100	0	6500	1600	30	54	3.2
17	Georgia	Chatham Co.	26	74	3800	960	40	30	1.8
18	Alabama	Clarke Co.	93	7	10000	2500	50	71	4.2
19	Arizona	Maricopa	100	0	2600	660	65	76	4.5
20	Colorado	Morgan Co.	100	0	6900	1700	60	48	2.8
21	Idaho	Minidoka Co.	100	0	6000	1500	65	39	2.3
22	Idaho	Canyon Co.	35	65	2200	560	45	38	2.2
23	Montana	Chouteau Co.	100	0	9400	2300	80	41	2.4
24		Flathead Co.	10	90	5100	1300	75	35	2.0
25		Broadwater Co.	36	64	2200	560	40	29	1.7
26	California	Trinity Co.	13	87	16000	4100	80	41	2.4
27		Fresno Co.	88	12	10000	2500	75	37	2.2

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[1] Listed County name is assumed to be the site of the conversion plant which uses the feedstock from that county and adjacent counties selected by SAI.

[2] Based on site-specific data from SRI's - An Evaluation of Agricultural Residues as an Energy Feedstock - A Ten-Site Survey and county by county inventory of residues. Multi-county aggregation, and transportation costs corrected for residue density and distance by SAI.

[3] Weighted average for major residues. Estimated by SAI using low moisture residues.

Table 3

REPRESENTATIVE METHANOL CONVERSION SITES BASED ON
ENERGY FARM BIOMASS AVAILABILITY AND COSTS (average radius:50 miles)*

				ODT/Day		Near-Term		Long-Term	
Site #	State	Plant Location	Acres	Present	Future	$/ODT	$/MMBtu	$/ODT	$/MMBtu
1	Wisconsin	Rusk	439000	6600	13000	42	2.5	28	1.6
2	Missouri	Polk	248000	6000	11000	35	2.0	25	1.5
3	Louisiana	Sabine	325000	8900	16000	29	1.7	24	1.4
4	Georgia	Hancock	297000	9000	16200	33	1.9	24	1.4
5	New England	Windsor	123000	1900	3800	46	2.7	31	1.8
6	Washington	Lewis	129000	3900	7800	34	2.0	26	1.5
7	Illinois	McLean	64000	1200	2400	57	3.3	37	2.2
8	California	Kern	39000	1200	2400	46	2.7	35	2.0
9	Florida	Levy	299000	9100	16300	30	1.8	25	1.5
10	Mississipi	Perry	299000	9100	16300	30	1.8	25	1.5

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* Not maximum radius. One tenth of suitable permanent pasture, forest, range, rotation hay and pasture, hayland and open land formerly dropped. County data from MITRE, Silvicultural Biomass Farms, Vol.III, 1977. Multi-county aggregation and delivered cost adjusted for distance by SAI.

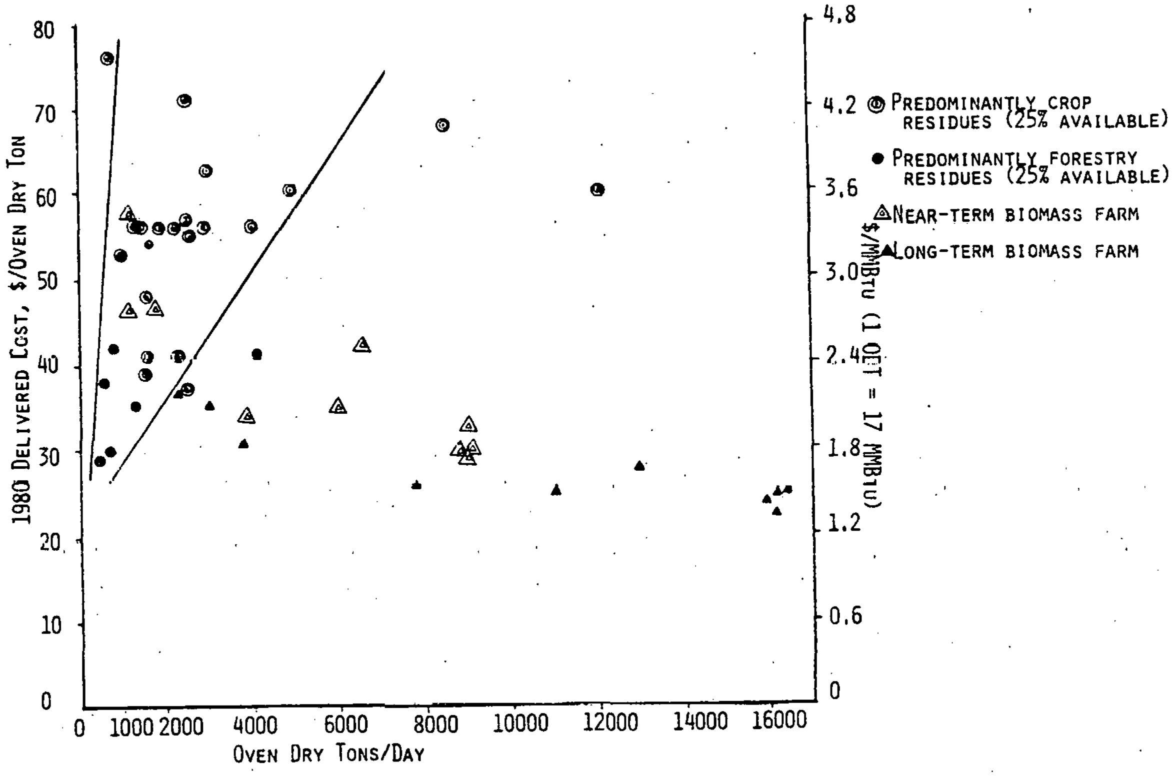

Figure 1

BIOMASS FEEDSTOCK ECONOMICS

Table 4

POTENTIAL METHANOL MARKET SHARES BY REGION, 1985-2000 (billion gallons per year)

	Scenario 1					Scenario 2				Scenario 3			
	10 Percent Blend with Gasoline					44 Percent Gasoline Displacement & 30 Percent Diesel Displacement(2)				Multi-Market Penetration Scenario Gasoline, Diesel, Distillate and Residual(3)			
Region	1985		1990	1995	2000	1985	1990	1995	2000	1985	1990	1995	2000
New England	0.81	(0.79)(1)	0.67	0.61	0.59	4.13	3.75	3.68	3.81	6.26	5.83	5.73	5.83
Mid Atlantic	1.59	(2.25)	1.31	1.18	1.14	8.80	8.10	7.44	8.43	13.64	13.07	12.54	13.67
South Atlantic	2.09	(2.76)	1.72	1.53	1.48	11.83	10.87	10.87	11.27	13.57	12.70	12.73	13.31
East North Central	2.71	(3.06)	2.24	2.02	1.96	14.94	13.70	13.60	14.18	18.36	17.63	18.01	18.76
East South Central	0.84	(1.14)	0.68	0.60	0.57	4.89	4.47	4.39	4.56	5.34	5.00	4.99	5.20
West North Central	1.31	(1.52)	1.06	0.92	0.87	7.75	7.09	6.96	7.20	8.70	8.08	8.06	8.43
West South Central	1.81	(1.83)	1.46	1.24	1.17	10.01	8.96	8.50	8.60	10.60	9.56	9.08	9.18
Mountain	0.57	(0.88)	0.46	0.41	0.40	3.58	3.36	3.41	3.63	4.02	3.78	3.82	4.04
Pacific	1.72	(3.55)	1.46	1.38	1.38	9.91	9.48	10.03	10.86	11.08	10.80	11.48	12.38
All Regions	13.44	(17.78)	11.06	9.89	9.56	75.84	69.78	68.81	72.54	91.57	86.45	86.44	90.80

Petroleum Fuel Displacement (Million Barrels Per Year)

	Scenario 1					Scenario 2				Scenario 3			
Gasoline	208.1	(275.1)	171.2	153.1	147.8	915.6	753.4	673.8	650.4	915.6	753.4	673.8	650.4
Diesel	-		-	-		187.5	237.3	284.0	342.6	187.5	237.3	284.0	342.6
Distillates	-		-	-		-	-	-	-	156.9	168.1	179.8	187.9
Residuals	-		-	-		-	-	-	-	17.2	16.4	15.6	14.7
Equivalent Crude Oil(4)	218.5	(288.9)	179.8	160.8	155.2	1158.3	1040.3	1005.7	1042.6	1341.1	1234.0	1153.2	1255.4

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(1) Numbers in parentheses are based on FEA estimates, they are shown here for comparison purposes. Other projections were derived from predicted petroleum demands by the SRI energy model.

(2) Assumed 70 percent of gasoline vehicles use 20 percent methanol blend, and 30 percent of gasoline vehicles switch to pure methanol. Also assumed 30 percent of diesel demand projected by SRI to be displaced by methanol.

(3) Assumed to be methanol requirement for scenario 2 plus requirement equivalent to 30 percent of distillate market and 5 percent of residual market.

(4) Equals total volume of displaced petroleum products plus 5 percent assumed refinery losses.

- Three scenarios were developed to assess the future methanol requirements based on:

  1) 10% automotive gasoline displacement
  2) 44% displacement of the automotive gasoline market and 30 percent displacement of the automotive diesel market.
  3) Same as scenario (2) plus 30% displacement of distillate and 5% of the residual oil market.

- Scenario (1) methanol requirements can be satisfied reasonably by a future biomass availability, equivalent to 20 percent of projected total biomass resource in the near-term (1985-1990) (Table 5).

- Scenario (2) and Scenario (3) requirements cannot be satisfied by the projected availability of U.S. biomass resources. It is estimated that a 10 to 20 percent displacement of automotive gasoline is a reasonable target.

Conversion Process Analysis

The results of conversion processes analysis can be broken down as follows:

Biomass Gasification

- Biomass materials which are gasified by partial oxidation in the conventional manner will produce a synthesis gas with an insufficient quantity of hydrogen for optimum methanol synthesis. This may necessitate costly gas treatment steps including hydrocarbon removal, carbon dioxide removal and CO shifting with steam.

- In a low temperature gasification processes, a substantial quantity of CH_4 or other hydrocarbons along with oil and tar formation may result. The cleaning and conditioning of such a gas product require a costly steam reforming or cryogenic separation process.

- In the near term (before 1990), large scale gasification technologies that have been developed for coal and solid wastes could be adapted for the production of methanol from biomass. Among the commercially available gasification processes, the oxygen-blown Koppers-Totzek entrained bed gasifiers, the oxygen blown Wellman-Galusha fixed-bed gasifiers and Purox gasifiers are compared in Table 6.

- In the long term, advanced catalytic gasification processes may be a significant improvement in the production of synthesis gas from biomass. Analysis indicates that advanced catalytic and hydro-gasification of biomass could produce a hydrogen-rich synthesis gas suitable for optimum methanol synthesis.

Status of Methanol Conversion Technology

- Methanol synthesis technology is a highly developed, competitive field with several commercially proven processes as represented by the ICI and Lurgi processes.

- Based on comparison of commercially proven technology, the Lurgi process has a slightly higher process thermal efficiency than the ICI process due primarily to the higher steam energy recovery. However, the capital cost of Lurgi's isothermal reactor system is higher than the ICI's quench type reactor system. (Table 7)

- Methanol processes operating at higher pressures have a lower process thermal efficiency because of their higher feed gas compression energy requirements.

- A recently developed liquid phase methanol process has the highest process efficiency and shows potentially lower capital costs than current commercial processes. This system employs an inert liquid to absorb the heat of reaction and thereby increases the conversion rate and improves the waste energy recovery.

- The current methanol synthesis processes have a process efficiency ranging from 84 to 90 percent, depending on the individual process and plant size. The new liquid-phase process, currently under development, can achieve a higher conversion efficiency, 93%.

Overall Biomass to Methanol Process

- The overall processing steps for an integrated biomass-to-methanol plant include biomass preparation, gasification, gas compression, acid gas removal, CO shift, CO_2 removal, and methanol synthesis.

- The overall process efficiency depends primarily on both the synthesis gas composition and the efficient utilization of waste heat for compression power requirements. The largest power requirements are associated with syngas compression and the oxygen plant.

- The high-temperature Koppers-Totzek gasification system gives a highest process efficiency of 60.4 percent, while the Purox process would have a lower efficiency of 43.6 percent (see Table 8).

Methanol Production Economics

- Estimated production costs of methanol from biomass range from 75 to 95 cents per gallon for a 1000 TPD biomass to methanol plant, depending on the conversion processes selected (see Table 9). These production cost are approximately 30 percent higher than the methanol production costs using other feedstocks. (see Table 10).

Table 5

POTENTIAL METHANOL SUPPLY FROM BIOMASS BY REGION: 1985-2000 (billion gallons per year)

	25 Percent Biomass Used				100 Percent Biomass Used			
Region	1985	1990	1995	2000	1985	1990	1995	2000
NE	0.55	0.64	0.75	0.85	2.20	2.56	3.00	3.40
MATL	0.82	0.97	1.18	1.39	3.27	3.88	4.72	5.57
SATL	3.00	4.06	4.92	5.78	12.00	16.24	19.68	23.12
ENC	2.30	2.82	3.34	3.85	9.21	11.26	13.36	15.38
ESC	1.97	2.54	3.06	3.57	7.87	10.17	12.24	14.29
WNC	3.63	4.24	4.70	5.15	14.52	16.96	18.80	20.59
WSC	2.51	3.63	4.40	5.18	10.04	14.52	17.60	20.72
MT	1.06	1.09	1.11	1.12	4.24	4.36	4.42	4.48
PAC	0.85	0.97	1.09	1.21	3.39	3.88	4.36	4.85
All Regions	16.69	20.96	24.55	28.10	66.74	83.83	98.18	112.40

Table 6

COST COMPARISON OF POTENTIAL BIOMASS GASIFICATION PROCESSES FOR PRODUCING SYNTHESIS GAS(1) (Cost in Million Dollars 1980)

	Koppers-Totzek	Purox	Wellman-Galusha
Feed capacity, TPD (dry)	1000	1000	1000
Synthesis gas output, MMBtu PD	13140	9880	11510
Capital costs			
Front end	17.0	4.3	4.3
Gasifier	41.0	35	20
Oxygen plant	14.2	9.0	13.5
Gas cleaning and pumping	-(2)	4.6	5.4
Gas separation	-	7.5	4.3
Contingency and engineering	3.3(3)	9.8	7.1
Interest, start-up and working capital	2.1(3)	6.0	4.8
Total	77.6	76.2	59.4
Operation and maintenance costs	11.5	11.2	8.80
Synthesis gas production costs, $/MMBtu			
Biomass feedstock cost, $/MMBtu			
1.0	7.52	9.83	6.95
2.0	8.81	11.6	8.43
3.0	10.1	13.3	9.91

---

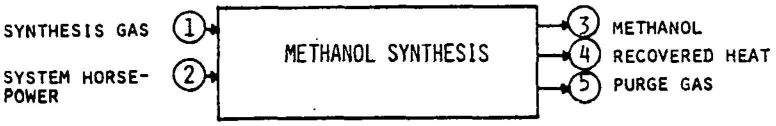
(1) Synthesis gas production considers CO + H_2 only for methanol synthesis

(2) Included in gasifier cost

(3) Gasifier cost includes some contingency and engineering cost.

Table 7

COMPARISON OF METHANOL SYNTHESIS TECHNOLOGY



	ICI Multi-Bed Adiabatic Quench System		Lurgi Tube-Wall Isothermal Reactor System		Chem System Liquid Phase Reactor System	
Operating pressure, psi	750	1500	500	1500	500	1100
Energy balance, MMBtu/ Ton MeOH						
① synthesis gas	23.52	23.52	23.52	23.52	23.52	23.52
② System horsepower requirements						
Feed compression(1)	0.80	2.48	-	2.48	-	1.70
Gas recycle	0.92	0.25	1.39	0.27	0.55	0.14
Oil circulation	-	-	-	-	0.27	0.12
Total energy input	25.24	26.25	24.91	26.27	24.34	25.48
③ Methanol	19.28	19.28	19.28	19.28	19.28	19.28
④ Recovery heat	1.15	1.17	2.35	2.40	2.57	2.58
⑤ Purge gas	0.99	1.00	0.99	1.01	0.99	1.19
Total energy output	21.42	21.45	22.62	22.69	22.84	23.05
Process Efficiency, %	84.9	81.7	90.8	86.4	93.8	90.5
Relative capital cost	1.00	0.71	1.20	0.86	0.77	0.54

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(1) Assume synthesis gas available at 500 psig.

Table 8

COMPARISON OF BIOMASS TO METHANOL PROCESSES(1)

	ICI + Kuppers-Totzek	ICI + Purox	ICI + Wellman-Galusha
Biomass Input to Gasifier, TPD (dry)	1000	1000	1000
Oxygen Requirement, TPD	560	300	520
Methanol Product, TPD	535	386	450
Process Efficiency(2), %	60.4	43.6	50.8
Heat and Steam Generated, MMBtu, PD			
Gasification	2060	0	0
Fuel gas by-product	410	4350	2030
Large Steam Turbine Requirement, HP			
Syngas compressor	15340	12380	13600
Oxygen plant compressor	10160	5440	9440
Methanol recycle compressor	320	220	250
Total horsepower	25820	18040	23290
Net Process Power Produced, MMBtu, PD			
Steam	200	-	-
Fuel gas	410	2780	300

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(1) Oxygen blown biomass gasification processes

(2) Process efficiency = $\frac{\text{HEATING VALUE OF METHANOL PRODUCT}}{\text{TOTAL HEATING VALUE OF BIOMASS AND OXYGEN PLANT ENERGY REQUIREMENT}}$

Table 9

COMPARISON OF INVESTMENT AND OPERATING COSTS OF CONCEPTUAL BIOMASS TO METHANOL PROCESSES (in 1980 dollars)

	KOPPERS-TOTZEK	PUROX	WELLMAN-GALUSHA
BIOMASS FEED, TPD (DRY)	1000	1000	1000
METHANOL PRODUCTION, TPD	535	386	450
CAPITAL INVESTMENT, MM$			
FRONT END	17.0	4.3	4.3
OXYGEN PLANT	14.2	9.0	13.5
GASIFIER	41.0	35.0	20.0
GAS CLEANING AND HC SEPARATION		12.1	9.7
SHIFT AND ACID GAS REMOVAL	6.5	5.3	5.9
METHANOL SYNTHESIS AND PURIFICATION	12.1	9.3	10.5
GENERAL FACILITY AND UTILITY	22.4	22.5	19.2
SUBTOTAL	113.2	97.5	83.1
CONTINGENCIES AND ENGINEERING @15%	16.9	14.6	12.5
INTEREST, START-UP AND WORKING CAPITAL, @10%	11.3	9.8	8.3
TOTAL INVESTMENT	141.4	121.9	103.9
ANNUAL OPERATING REQUIREMENTS, MM$			
WOOD AT $34/ODT ($2/MMBTU)	11.2	11.2	11.2
CATALYSTS AND CHEMICALS	0.12	0.09	0.10
LABOR	0.73	0.52	0.61
UTILITY	1.13	0.82	0.95
FIXED CHARGE @20% TOTAL INVESTMENT	28.3	24.4	20.8
ANNUAL TOTAL	41.5	37.0	33.7
METHANOL PRODUCTION COSTS, $/GALLON	0.85	0.96	0.75
METHANOL PRODUCTION COSTS, $/TON	258	290	226

Table 10

COMPARISON OF METHANOL PRODUCTION COST USING VARIOUS FEEDSTOCKS

PROCESS	METHANE REFORMING	RESID PARTIAL OXIDATION	K-T COAL GASIFICATION	IMPROVED COAL GASIFICATION	GASIFICATION OF BIOMASS[1]
	GULF COAST	GULF COAST	MID-CONTINENT	MID-CONTINENT	K-T
PLANT CAPACITY ST CH_3OH/DAY	2000	2000	2000	2000	535
FEEDSTOCK COST AT PLANT	-	$15.00/BBL	$21.80/TON	$21.80/TON	$34/TON
FEEDSTOCK COST AT PLANT, $/MBTU	3.15	2.35	0.96	0.96	1.76
INVESTMENT, $M					
ONSITE	87.9	137.9	206.1	175.2	100.2
OFFSITE	47.4	93.3	144.3	134.9	29.9
TOTAL PLANT	135.3	231.1	350.4	310.1	130.1
WORKING CAPITAL	13.6	11.4	5.2	4.9	11.3
OPERATING COSTS AND ANNUALIZED CAPITAL CHARGES, $MM					
FEEDSTOCK	73.7	61.5	24.7	23.1	11.2
OPERATION, MAINTENANCE COST AND CAPITAL CHARGES	55.1	85.2	126.1	109.3	30.3
TOTAL ANNUAL COST	128.8	146.7	150.8	132.4	41.5
SULFUR CREDIT ($50/TON)	-	(1.0)	(1.9)	(1.8)	-
NET PRODUCTION COST	128.8	145.7	148.9	130.6	41.5
COST PER GALLON (1980$)	0.65	0.74	0.75	0.66	0.85

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[1] USING WOOD CHIPS AS THE FEEDSTOCK.

- The capital costs of a plant constitute nearly 70 percent of annual production cost of methanol. It is anticipated that an improved biomass-to-methanol process would be able to reduce the overall capital cost by as much as 30 percent (Table 11). This translates to a reduction in the production cost of methanol to within the competative, 40-50 cents-per-gallon range.

- A realistic methanol plant size utilizing existing technology lies between 500 to 1000 TPD of methanol. Larger plants currently are not feasible because of the dispersion and limited availability of biomass resources as illustrated in Figure 2.

Summary Status of Methanol Production from Biomass

- Methanol production cost from biomass will be significantly affected by:
  - Feedstock availability and cost
  - Conversion plant size
  - Selection of a specific biomass production and conversion route
  - Future technological and process improvements in reducing capital cost requirements.

- In the near term (1985), methanol production from biomass using existing technology can not compete with methanol produced from other feedstocks unless economic incentives are introduced or innovative process schemes are developed.
  - Federal policy and economic assistance such as federal sale tax exemption (40 cents per gallon for a 10 percent methanol-gasoline blend), and capital investment cost amoritization credit.
  - Large scale methanol production based on dispersed conversion plant concepts such as pipeline transport of synthesis gas and barge-mounted conversion facilities.
  - Use of hybrid feedstocks in existing methanol production facilities such as methane with biomass or hydrogen with biomass.

- In the longer term (1990 and on), methanol production from biomass could become economically competitive contingent upon the following developments.
  - Large scale and low cost biomass production from energy farms.
  - Technological improvements in gasification and methanol technology such as a single stage catalytic gasification process for methyl fuel production.
  - Multiple product conversion routes such as methanol and ethanol, methanol and acetic acid, methanol and ammonia, or to methanol and gasoline.

Table 11

CAPITAL COST REDUCTION POTENTIAL FOR BIOMASS BASED METHANOL PROCESSES

Current Biomass to Methanol Process		
Processing Step	Percent of Capital Cost	Advantages of Advanced Biomass to Methanol Process
Front end	8.7	Reduce feedstock drying and size reduction requirements
Oxygen plant	12.5	Elimination of oxygen plant
Gasifier	32.7	Use pressurized steam gasification or catalytic fluidized bed gasification
Gas cleaning and separation	7.4	Reduce oil and tar formation, and hydrocarbon separation
Shift and acid gas removal	6.0	Produce synthesis gas with $H_2/CO=2$ and eliminate shift conversion step
Methanol synthesis	10.9	Use Chem System methanol process to improve methanol conversion efficiency and production cost
General facility and utility	21.8	Improve waste heat recovery efficiency and elimination of waste disposal facility. High pressure gasification to reduce need for compressor
Total	100	30 percent capital cost reduction

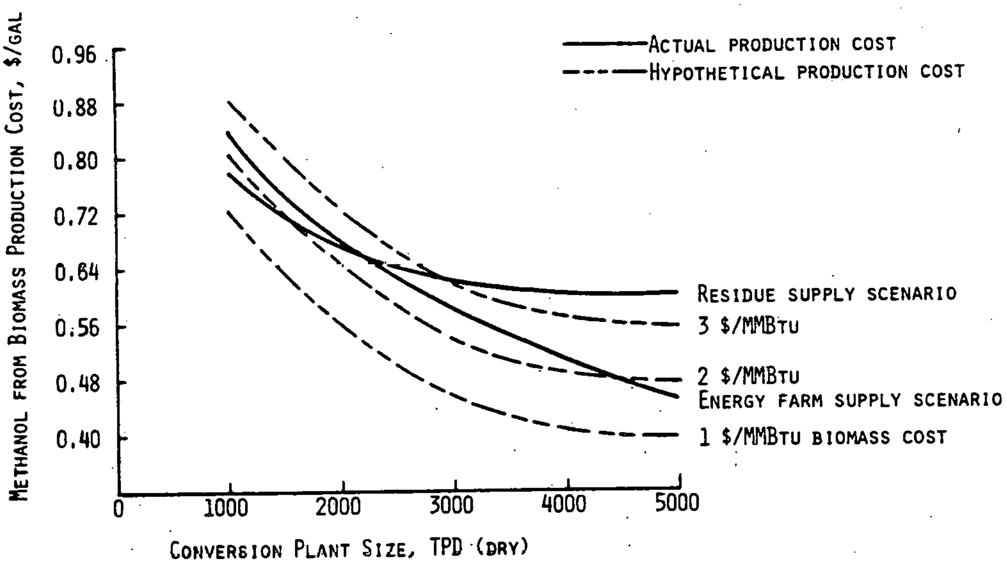

Figure 2

SENSITIVITY OF METHANOL FROM BIOMASS PRODUCTION COST

NOTES

POTENTIAL OF WOOD GASIFICATION FOR INDUSTRIAL APPLICATION

Abu Talib — The MITRE Corporation/Metrek Division
Jack Fritz — Formerly with MITRE, now with Office of Energy, AID, Department of State
Judy Gordon — The MITRE Corporation/Metrek Division

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ABSTRACT

Gasification of wood and wood residues using air and steam as the gasifying medium to produce a low-Btu (100-200 Btu/scf) gas is a technology which has been commercially available since the turn of the century. However, the advent of inexpensive fossil fuels phased this technology out of the market. Today, there is a renewed interest in revitalizing this old technology. Approximately two dozen private organizations are attempting to enter the commercial marketplace. MITRE conducted a study to assess the commercial readiness of wood gasification in the near-term. Several organizations involved in the manufacture and development of gasification technology were queried. Current status and systems costs were found to vary considerably.

This paper summarizes the MITRE findings, giving both a state-of-the-art review and outlining policy related issues. Economic potential is assessed by considering a close-coupled wood gasifier as a source of gaseous fuel for an oil- and gas-fired boiler. The cost of wood-based fuel in three regions* is compared with the cost of conventional fossil fuels for 1985 and 1990 under different economic scenarios. Results indicate that wood gasification is competitive in the New England and Mid-west regions by 1985. In the Northwest region, high wood prices prevent wood-based low-Btu gas from being cost competitive even in a long-term (1990) comparison. Some government participation in the form of further R&D and financial incentives will be required to accelerate commercialization.

GASIFICATION PROCESS DESCRIPTION

The gasification process converts a solid carbonaceous feed to a gaseous fuel that may be burned directly or upgraded to higher quality fuels or used as a feedstock for manufacturing chemicals, e.g., ammonia or methanol. However, the gas produced from an air and steam blown gasifier contains large amounts of nitrogen (nearly 50 percent by volume), and it is primarily suitable for on-site fuel application.

Reactors used for the gasification process are generally characterized by the method of contacting solids and gases. The principal reactor configurations are:

- Fixed Bed
- Stirred Moving Bed
- Fluidized Bed
- Entrained Flow

Of these configurations, only fixed-bed gasifiers have been commercially used for wood gasification. The most common design is an updraught fixed-bed reactor in which three distinct reaction zones, drying, pyrolysis and combustion, can be identified. Representative generic reactions are outlined below:

<u>Drying Zone</u>

212-400°F

Moist wood + Heat → Dry Wood + Water Vapor

<u>Pyrolysis Zone</u>

400-900°F

$$\text{Dry Wood + Heat} \rightarrow \text{Char} + CO + CO_2 + H_2O + CH_4 + C_2H_4 + \text{other HC + pyroligneous acids + tars}$$

<u>Gasification and Oxidation Zone</u>

$$\underset{\text{(Steam)}}{\text{Char} + O_2 + H_2O} \rightarrow CO + H_2 + CO_2 + \text{Heat}$$

These reactions follow in sequential fashion as the wood descends by gravity through the gasifier.

A commercial gasifier using the above design is presented in Figure 1. This single stage fixed bed gasifier is available from Davy Power Gas Co.,

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[illegible]ederal planning regions

Houston, Texas. The gasifier is a steel cylinder; the upper part is refractory lined, and the bottom part is surrounded by an annular boiler which generates steam. The by-product steam is normally introduced with the air or oxygen to aid in the gasification reactions and to control the fire zone temperatures.

Wood chips (typically 1 1/2" x 1/2") or other feed material (coal) is delivered to a bunker designed to contain an 8 to 12-hour supply at maximum rates. From the bunker the wood is automatically delivered by gravity through a fuel hopper into the gasifier. The gasifier operates at near atmospheric pressure. The gasifier contains an internal chute to distribute the wood chips to minimize gas channeling up through the fixed bed.

As the wood descends down through the gasifier it passes through a drying-devolatization zone, a gasification zone, fire zone and ash zone. Steam and air are used as the gasifying media, and are introduced through a rotating eccentric grate at the bottom of the gasifier to effect good distribution up through the fixed bed.

The fuel gas produced from the devolatization and gasification of wood, exits the gasifier at 250°F to 1200°F depending on the moisture content of the feedstock. The exit gas contains tars and oils as well as the combustible CO, H_2 and CH_4 gaseous components. Typical gas analyses for air blown gasification of wood are given in Table 1.

The product gas has a heating value of 120 to 175 Btu/scf depending upon the feedstock, moisture content and operational variables. The gas is most likely to be used in a close coupled boiler, kiln or dryer after particulate removal in a simple cyclone. It is desirable to keep the produce gas hot, usually 500°F to 800°F, to minimize condensation of the tars and to maximize sensible heat transfer.

WOOD GASIFICATION STATUS REVIEW

A number of commercial processes are available, with no single one currently possessing an obvious competitive advantage. Several firms have developed systems for specialized application such as agricultural residue conversion, wood residue conversion and municipal solid waste disposal. Most of these systems have capacities of less than 50 MM Btu/h of product gas, i.e., 80-90 oven dried tons per day (ODT/D) of feedstock. Larger systems gasifying more than 100 ODT/D of wood are similar in design to gasifiers developed prior to 1950.

Of systems currently in operation, few have demonstrated long term successful performance. Technical problems related with the disposal of organic liquids formed during gasification and ash removal are being resolved. Process development efforts also are underway at various research centers and universities. These processes are principally in the development stage and focus on the different aspects of fluidized-bed and catalytic gasification. Some of this work appears ready to be transferred to the private sector for commercialization. Increased cooperation between private developers and R&D institutions would aid in reducing remaining technical problems. A summary of several gasification processes along with the current status is presented in Table 2.

COMMERCIAL APPLICATION

Existing oil and gas-fired package boilers represent the most viable potential application for hot wood-based gas. The retrofit of these boilers to burn this gas does not present any unusual technical problems. It will generally include:

- Replacement of the existing burners
- Expanding the duct work for increased flue gas volume
- Adjusting the capacity of the inducted draft fan or replacing it with a larger unit.[1,2]

In most cases the retrofit of oil- and gas-fired package boilers to burn low-Btu gas will involve minimum boiler derating. This is particularly true if the original boiler has a relatively large furnace. However, the present day oil- and gas-fired package boilers are designed to have very close tube spacing which results in a compact furnace. This will limit the extent of retrofit operation and will introduce some boiler derating when low-Btu gas is used as fuel. The magnitude of boiler derating, depending on the original furnace design (relatively spacious, or very compact) may range between 5 and 25 percent of the initial design capacity.

ECONOMIC EVALUATION

Based on information supplied by manufacturers, capital costs of wood gasification systems vary from approximately $5,000 to $20,000/ODT/D of capacity.[3] This wide variation is the result of differences in system capacity, inclusion of specific components, varying cost assumptions, and the requirement for land and facilities. The average costs are $10,000 to $12,000/ODT/D for a system which includes materials handling equipment, gasification equipment and ancillary systems such as controls but excludes a boiler, land, building, contingencies and fees.

The low- and medium-Btu gas production costs are highly sensitive to feedstock cost and to a lesser extent to the system capacity. Some examples of published gas costs are given in Table 3. It is important to note that costing procedures differed with each case and that results are not directly comparable.

summary, both LBG and MBG can be produced at a cost of $2.00 to $3.00/MMBtu with a $1.00MMBtu feedstock cost, or $3.00 to $5.00/ MMBtu at a feedstock cost of $1.5MMBtu.

In order to assess the relative economic competitiveness of wood-based gas for industrial application, an oil/gas-fired boiler retrofitted with two gasifiers was considered. The capital investment includes the installed cost of wood handling/feed preparation, the wood gasification units, and all the auxiliary and support facilities. An estimate is made of the cost required to retrofit the existing boiler in such a way as to avoid derating.* This incremental investment is treated as part of the total capital investment. Plant costs are extracted from information supplied by Davy Power Gas Co., and from similar coal gasification systems.[4]

Total investment and annual operating costs were developed for a wood gasification fuel supply system capable of gasifying 220 ODT/D (Table 4). Two gasifiers supply fuel to an existing 100,000 pound per hour steam producing plant. Total investment is approximately $5 million, and the annual operating cost, excluding the wood feedstock cost, is about $730,000.

The cost of gas from the industrial boiler retrofit is compared to the delivered prices to the industrial sector of distillate and residual oil and natural gas. Comparisons are presented for the Midwest, New England and Northwest regions for 1985 and 1990 under three economic scenarios. The scenarios are based on assumptions regarding escalations in the price of oil and the resultant impacts on the prices of other conventional fuels. Scenarios are classified as low, medium, and high, corresponding to oil prices of $14.50, $15.15, and $19.72 per barrel, respectively, for 1985 and $15.60, $19.40, and $29.00 per barrel, respectively, in 1990. The scenarios are part of DOE's current policy and evaluation program.[5]

The major assumptions used in the gas cost estimates include:

- a mature technology plant
- inflation rate 6 percent
- effective income tax rate 50 percent
- investment tax credit 10 percent
- depreciable life 80 percent of project life
- double declining balance depreciation method for tax purposes
- debt/equity ratio for private competitive firm 30 percent/70 percent
- nominal after tax rate of return on equity of 16.1 percent
- nominal before tax ccost of debt of 9.2 percent

Costs for the wood biomass feedstock were also estimated by applying the above assumptions to the harvesting equipment. In addition, different assumptions regarding operation productivity--variations caused principally by differences in types of forest stand--yielded a range of costs. This range is the basis of the differences between the low, medium, and high price scenarios for the wood feedstock. The wood biomass cost for each region represents a weighted average of the proportion of the various types of wood residues available in 1985 and 1990[6] and their respective collection cost using present day wood collection technology.[7,8] The costs are escalated by average timber price increases projected by the Forest Service for each region considered.[9] Additional assumptions used in calculating wood feedstock costs are as follows:

- harvesting strategies: clearcutting to produce chips and residue collection
- technical and cost data for equipment from references 5 and 6
- stumpage cost: $1.00/dry ton
- transportation cost: $.14/ton-mile Northwest and $.12/ton-mile for Midwest and New England
- transportation distance: 50 miles
- operation schedule: 1600 hours/year
- annual cost escalation factors:
  .8% Midwest
  .5% New England
  4.3% Northwest

Results of the energy cost comparison are presented in Table 5. In the Midwest, low-Btu gas is estimated to cost $3.28 and $3.43 per million Btu in 1985 and 1990, respectively, under the medium price scenario. In 1985, low-Btu gas is cheaper than distillate oil and somewhat more expensive than residual oil and natural gas but essentially competitive with all three fuels. In 1990, low-Btu gas is competitive with residual oil and natural gas and possesses a substantial economic advantage over distillate oil under the medium cost scenario. Results from the high price scenario in both years show low-Btu gas at a disadvantage compared to natural gas, competitive with residual oil, and again, substantially cheaper than distillate oil.

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*It is assumed that the existing boiler has a relatively large furnace so that a retrofit operation will not significantly derate the boiler.

In New England, the economics of low-Btu gas are extremely attractive. Under the medium price scenario for 1985 and 1990, respectively, low-Btu gas is projected to cost $3.28 and $3.43 per MM Btu. Low-Btu gas is substantially cheaper than distillate oil and competitive with residual oil in both years under all scenarios. The same result holds in comparison to natural gas, with the exception of the high escalation scenario in 1990. Here the cost of low-Btu gas exceeds that of natural gas by about $.30 per MM Btu.

Low-Btu gas does not fare as well in the Northwest where wood-based gas remains an expensive option through the year 1990. Low-Btu gas at $4.60 and $5.26 per MM Btu in the medium price scenario in 1985 and 1990, respectively, is not competitive with any of the conventional reference fuels; this result does not change for any of the other scenarios considered.

CONCLUSIONS

The economic competitiveness of low-Btu gas is highly sensitive to the availability of low-cost wood feedstock and therefore regionally dependent. The New England Region where wood feedstock costs are relatively low, the economics of wood-based gas are extremely attractive. In contrast, in the Northwest, the higher wood feedstock cost makes low-Btu gas much less competitive. The Midwest enjoys relatively low wood feedstock and conventional fuel costs. On balance, wood-based low-Btu gas appears to be a potentially attractive option in this region. Except in the Northwest region, the comparatively high escalation in conventional fuel prices helps enhance the competitiveness of wood-based gas in the near-term.

The results of our preliminary study are based on an evaluation of a wood gasification retrofit system for industrial oil- and gas-fired boiler application. Some simplifying assumptions were made regarding the ease of retrofit and amount of boiler derating involved when switching to low-Btu gas. These assumptions are justified for taking a first cut at the relative competitiveness of wood-based low-Btu gas. However, an indepth technical and economic feasibility study must be conducted to confirm the viability of wood gasification for supplying industrial boiler fuel.

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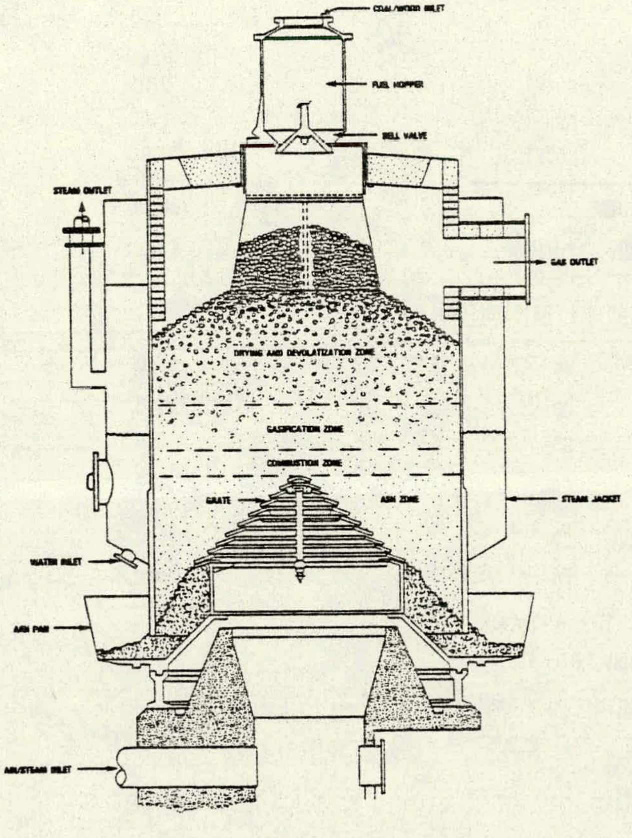

FIGURE 1
SINGLE STAGE FIXED BED WOOD GASIFIER
Reference Source

TABLE 1
FIXED-BED WOOD GASIFIER
GAS ANALYSES

FEEDSTOCK ULTIMATE ANALYSIS (WT%):	VENTEAK WOOD
C	30.90
H	3.60
O	25.35
N	0.7
S	0.05
ASH	1.00
MOISTURE	38.40
TOTAL	100.00
HHV, BTU/LB	5,410
GAS COMPOSITION (VOL. %, DRY BASIS, TAR FREE)	
H_2	13.0
CO	29.0
CO_2	6.6
CH_4	4.0
N_2 + A	47.4
H_2S + COS	----
TOTAL	100.0
HHV, BTU/SCF	176
SCF OF DRY GAS/LB FEED	17.2
TAR-OIL PRODUCT/LB FEED	0.081*

*INCLUDES METHYL ALCOHOL, DOES NOT INCLUDE ACETIC ACID.

TABLE 2

STATUS SUMMARY OF DIFFERENT GASIFICATION PROCESSES

PROCESS DEVELOPER	HOME OFFICE	DEVELOPMENT STATUS	CAPACITY (ODT/D)	ENERGY PRODUCT(S)	REACTOR TYPE	CURRENT APPLICATION
American Fyr-Feeder	DesPlaines, IL	C	1-15*	LBG	SMB	Steam
Andco Inc.	Buffalo, NY	C	75-200	LBG	MPB	Steam
Applied Engineering Co.	Orangeburg, SC	C	5	LBG	MPB	Boiler Retrofit
Battelle Columbus Lab	Columbus, OH	D	N.A.	MBG	EF	---------
Battelle Northwest Lab	Richland, WA	D	2	LBG	SMB	---------
Biomass Corp.	Yuba City, CA	D	3-4	LBG	FXB	---------
Bio-solar R&D Corp.	Eugene, OR	D	10	LBG	FLB	Drying
British Columbia Research	Vancouver, B.C., Canada	D	10	LBG	FLB	Boiler Retrofit
California, Univ., Davis	Davis, CA	D	18	LBG	MPB	---------
Canadian Ind. Ltd.	Kingston, Ont., Canada	D	24	LBG	N.A.	---------
Century Research Ind.	Gardenia, CA	D	N.A.	MBG/LBG	N.A.	---------
Davy Power Gas Inc.	Houston, TX	C	84	LBG	SMB	Various
DeKalb Ag. Research	DeKalb, IL	D	N.A.	LBG	N.A.	Drying
Eugene Water & Electric Board	Eugene, OR	D	N.A.	LBG	N.A.	Boiler Retrofit
Forest Fuels Inc.	Antrim, NH	C	2.3-9	LBG	SMB	Boiler Retrofit
Garrett E.R.E. Inc.	Ojai, CA	D	3	MBG	SMB	---------
Halcyon Inc.	East Andover, NH	C	15-80	LBG	N.A.	---------
Jamex Inc.	St, Peter, MN	D	8	LBG	SMB	Farm Utilization
Nova Scotia Tech. College	Halifax, N.S., Canada	D	5	LBG	MPB	Dual Fuel Engine
Pioneer Hi Bred Int. Inc.	Johnston, IO	C	20	LBG	MPB	Dual Fuel Engine
Quebec Electro Marine Diesel Inc.	Beaconsfield, Que., Canada	C	15	LBG	MPB	Dual Fuel Engine
Texas Tech University	Lubbock, TX	B	0.25	MGB	FLB	Synthesis Gas
Union Carbide Corp.	New York, NY	D	200*	MBG	MPB	Synthesis Gas
West Virginia University	Morgantown, WVA	B	N.A.	MBG	FLB	---------
Westwood Polygas Ltd.	Vancouver, B.C., Canada	D	60	LBG	MPB	Drying, Steam
Weyerhaeuser Co.	Tacoma, WA	B	N.A.	LBG	N.A.	---------
Wright Malta Corp.	Ballston Spa, NY	D	3	MBG	RK	Synthesis Gas

GLOSSARY

B - BENCH SCALE UNIT
C - COMMERCIAL SCALE SYSTEM, with units sold
D - DEMONSTRATION UNIT
LBG - LOW BTU GAS 100-200 BTU/SCF
MGB - MEDIUM BTU GAS 300 BTU/SCF

EF - ENTRAINED FLOW
SMB - STIRRED MOVING BED
FXB - FIXED BED
FLB - FLUIDIZED BED
MPB - MOVING PACKED BED

TABLE 3

LOW AND MEDIUM BTU GAS COST AS REPORTED IN LITERATURE

PROCESS AND SYSTEM CAPACITY (ODT/D)	FEEDSTOCK COST ($/MMBtu)		
	1.00	1.50	2.00
LBG Air Gasification (850)[1]	2.72	3.49	4.26
MBG Oxygen Gasification (850)[2]	2.55	3.32	4.00
MBG Oxygen Gasification (850)[3]	4.69	5.36	6.03
MBG Catalytic Gasification (850)[3]	1.83	2.35	2.87
MBG Entrained Flow Gasification (850)[3]	4.87	5.44	6.01
LBG Air Gasification (84)[4]	3.49	—	—
LBG Air Gasification (65)[4]	2.35	—	—

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TABLE 4

CAPITAL AND OPERATING COSTS FOR
A WOOD GASIFICATION/BOILER RETROFIT SYSTEM
(In Millions of Dollars, 1978)[1]

BASIS: Two gasifiers linked to a single boiler

Feed Capacity (ODT/D)		220
Product Gas Output (MBtu/hr)		125
Steam Production (lbs/hr)		100,000
System Stream Factor		0.90
Boiler Efficiency (%)		80
Plant Investment		
Installed Plant[2] Cost	4.14	
Interest During Construction	.21	
Total Installed Plant Construction	4.35	
Non-Plant Investment		
Land Cost	—	
Start-Up Costs	.51	
Working Capital	.16	
Non-Plant Total	.67	
		5.02
Total Plant Operating Costs		
Labor	.23	
Material	—	
Utilities	.05	
Supplies	.13	
Administrative and Overhead	.15	
Insurance and Property Taxes	.17	
Wood Feedstock Cost[3]	1.80	
Total Plant Operating Costs	2.53	

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[1] Davy Power Gas, private communication, January 1979.

[2] Includes equipment and boiler retrofit work.

[3] An average cost of $25/dry ton is assumed.

TABLE 5

COMPARISON OF PRICE OF LOW-BTU GAS FROM ON-SITE WOOD GASIFICATION PLANT AND DELIVERED PRICE OF CONVENTIONAL FUELS FOR YEARS 1985 AND 1990 ($\$/10^6$ BTU)

MIDWEST REGION	1985			1990		
	LOW	MEDIUM	HIGH	LOW	MEDIUM	HIGH
COST OF WOOD($/ODT)[a]	18.00	22.00	30.00	20.00	24.00	33.00
REFERENCE TECHNOLOGY						
• Distillate Oil	3.51	3.60	4.75	3.65	4.15	5.07
• Residual Oil	3.01	3.02	3.93	3.00	3.39	4.25
• Natural Gas	3.11	3.11	3.39	3.30	3.18	3.31
LOW-BTU GAS BOILER RETROFIT						
• Wood-Fired	2.99	3.28	3.87	3.14	3.43	4.09
NEW ENGLAND REGION						
COST OF WOOD ($/ODT)[a]	18.00	22.00	29.00	29.00	24.00	32.00
REFERENCE TECHNOLOGY						
• Distillate Oil	3.66	3.70	4.90	3.75	4.25	5.22
• Residual Oil	3.05	3.42	4.06	2.98	3.40	4.30
• Natural Gas	3.46	3.46	3.78	3.69	3.56	3.70
LOW-BTU GAS BOILER RETROFIT						
• Wood-Fired	2.99	3.28	3.80	3.14	3.43	4.02
NORTHWEST REGION						
COST OF WOOD ($/ODT)[a]	33.00	40.00	53.00	41.00	49.00	66.00
REFERENCE TECHNOLOGY						
• Distillate Oil	3.89	3.70	4.87	3.75	4.22	4.90
• Residual Oil	3.06	3.03	4.02	3.01	3.36	4.00
• Natural Gas	3.34	3.18	3.84	3.49	3.37	3.50
PROPOSED TECHNOLOGY						
LOW-BTU GAS BOILER RETROFIT						
• Wood-Fired	4.08	4.60	5.55	4.67	5.26	6.51

[a] An oven dry ton (ODT) is approximately 17 x 10^6 Btu.

NOTES

Session VI

BIOMASS ENERGY SYSTEMS: A CRITICAL ECONOMIC ANALYSIS

Anthony Y. Kam, Ronald L. Dickenson, Jerry L. Jones,
Anil K. Chatterjee, and Donald J. Wilhelm

SRI International
333 Ravenswood Avenue
Menlo Park, California 94025

ABSTRACT

This paper presents selected results from an economic and technical analysis of a broad spectrum of technology options (thermochemical and biochemical) for processing biomass materials into energy products. For each technology option, the study focuses on the projected timing of its readiness for commercial application, an assessment of its current status of process development, an analysis of the process/product economics, and, most significant of all, a critical and quantitative evaluation of the cost uncertainties associated with the option. The uncertainty analysis will provide valuable guidance to research programs and permit a more meaningful comparison of technologies in different development stages.

A large number of biomass energy systems are currently being developed. This study shows that the major factors contributing to cost uncertainties for biomass energy products are quite diverse in nature, including feedstock cost, state of process development, equipment estimation, design conservatism, and by-product credit.

INTRODUCTION

Biomass is a renewable resource that may play a significant role in providing for future U.S. energy needs. In this regard, a number of biomass energy systems (BES) are currently being developed. Technical and economic analyses of developing technologies are the principal means of determining their merits and commercialization potentials. The final product cost will, in most instances, be the major determining factor of the degree of market penetration by the product. Because cost estimates for any process/product are tentative, identification and quantification of the major uncertainties will provide valuable guidance to research programs and at the same time permit a more meaningful comparison of technologies in different stages of development.

This paper presents selected results from a critical economic and technical analysis of a broad spectrum of BES technology options. The work was performed under DOE Contract EY-76-C-03-0115/PA141 entitled "Preparation of a Cost Data Bank for DOE/Biomass Energy Systems Branch." This current project was initiated in October 1978. When completed, the data bank will provide a comprehensive and consistent data base for workers in the biomass energy field. For instance, it may be used with an appropriate methodology to analyze the complex issues of research program planning and analysis; project the likely future market penetration of biomass energy products; and examine the influence of government actions on the commercialization of biomass energy systems.

OBJECTIVES

The objectives of this study are to:

- Develop a technology and cost data bank for a broad spectrum of advanced thermochemical and biochemical conversion technology options.
- Estimate the likely regional prices, availability, and market share of biomass energy products (in quadrillions of Btu per year) from 1985 to 2025 using analytical market penetration models.
- Develop a set of cost goals for BES technologies and examine the aids and impediments to achieving these cost goals with particular emphasis on the federal role.
- Conduct a predesigned sensitivity study on the national market penetration of BES technologies based on interfuel competition analyses and projected end-use energy demands.

PROJECT STATUS

Figure 1 shows the performance schedule planned for this project. The data bank work is well underway and will be completed shortly. The methodology for meeting remaining project objectives is also being developed.

TECHNOLOGY AND COST DATA BANK

Eighteen BES technology options have been included in the data bank (see Table 1). These options cover the production of electric power, steam, gaseous, liquid, and solid fuels from biomass.

Most of the technology options are judged to be ready now or should be ready in the near term for commercialization. Actual commercialization of BES technologies will depend on additional market, institutional, comparative economic, and available resource considerations, and will be addressed during the upcoming phases of this project.

For each technology option selected, the data bank provides information on the projected timing of its readiness for commercial application, its current status of process development, an assessment of the process technology, an analysis of the process/product economics, and, most significant of all, a critical and quantitative evaluation of the cost uncertainties associated with the option.

COST UNCERTAINTY ANALYSIS

Technical and economic analyses of developing technologies are the principal means of determining their merits and commercialization potential. The final product cost will, in most instances, be the major determining factor of the product's market penetration. Until a plant is constructed and put into operation, all cost estimates for any process/product are subject to uncertainties; and the probability of a single-point cost estimate being correct is rather low. Identifying and quantifying the major areas of uncertainties will permit the establishment of the more meaningful upper and lower bound cost estimates to provide valuable guidance to research programs and also permit more meaningful comparisons of technologies in different development stages.

The required product revenue from a process plant can be expressed generally as:

Revenue required = capital-related charges + O&M* cost − by-product credit.

Each component on the right side of the equation is subject to uncertainties. In addition, the relative importance of these components varies from process to process.

Capital-Related Charges

Figure 2 illustrates the major sources of uncertainties for plant investment estimates. The state of technology development, equipment cost estimating, and the level of the cost analysis are included.

State of Technology Development. Plant investment estimates are generally developed from an estimated process design basis that may consist of feed and product specifications, process flow sheets, and heat and material balance data. The reliability of the design basis, which is strongly dependent on the process' development status, will affect the extent to which the actual plant cost deviates from the original estimate. Experience has shown that for processes in early development stages, deviations of 50 percent or much more are not uncommon.

Equipment Costing. The uncertainties associated with equipment costing often derive from the cost data available. Historical cost data, even when updated by means of cost indices, are often significantly in error. Installed equipment cost comprises design, material, fabrication, installation, and testing costs--all of which change almost continually over time. Until the equipment is installed and operational, a certain amount of cost uncertainty will persist.

Level of Analysis. The level of detail of a cost estimate is a significant determining factor of its degree of uncertainty. Major process units for a given process are usually readily identifiable. However, a process plant requires utility, energy conservation, piping and instrumentation, waste disposable systems, and many other ancillary facilities to support its operations. In many cases, these supporting systems represent a significant fraction of the total plant investment. Thus, the uncertainty of a cost estimate will decrease as the level of detail of the plant design on which it is based is increased.

For this study, a systematic approach was followed in quantifying the cost uncertainties of a selected technology option. The first step was the establishment of a conceptual plant design base case economic analysis. The plant design was then examined in sections (e.g., feed preparation, oxygen plant). For each plant section, technical uncertainty limits were assigned, based on current state of the art and outlook for further innovation. Equipment cost uncertainties were then defined based on the nature of the equipment, quality of the cost data, and consultation with suppliers. A set of locality cost uncertainty factors was also determined by considering the likely locations of such a plant.

Operation and Maintenance Costs. Operation and maintenance costs consist of feedstock, labor, supplies, and other costs. Biomass feedstock costs tend to be highly uncertain because of regional and seasonal availability, supply/price elasticity, alternative end-use market competition, transportation costs, and other factors. Labor and supply material costs tend to be highly site-specific. Within the United States, a variation of plus 15 percent to minus 5 percent (U.S. Gulf Coast as reference) can be expected.

By-Product Credit. For several BES technology options, particularly biochemical conversion routes, the revenue from by-products may be a significant fraction of the total plant revenues. Since

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* Operation and maintenance

product credit is a direct offset against the enue required from the energy product, the economic viability of such processes may hinge on the by-product credit. Examples of important by-products from BES technology options include fibers (obtained in the conversion of sugarcane to ethanol), pentosans, lignin chemicals (conversion of wood to ethanol), and cattle refeed (conversion of manure to substitute natural gas). In examining by-product values, one needs to consider their current market value, price/demand elasticity, alternative end uses, and potential alternative markets.

RESULTS AND DISCUSSIONS

Summary results of cost uncertainty analysis are presented and discussed below for:

1. Electric power generation
2. Oil and char production by pyrolysis
3. Catalytic liquefaction.

Technology for the first option is available now. Technology for production of pyrolytic oil from wood is estimated to be available by 1985. The availability of catalytic liquefaction technology appears to be more distant, most likely some time beyond 1985.

Power Generation

The estimated product revenue requirements for electric power generation by wood combustion is shown in Figure 3. The required selling price of the power ranges from about 5 to 9 cents per kWh. The most likely estimate is 6.5 cents per kWh. The estimates are for a grass roots plant consuming 1,000 dry tons per day of wood to produce 44 MW of salable power. Utility type of financing (65/35 debt/equity) and a 30-year project life with straight-line depreciation were assumed. Wood cost was assumed to be $1.25 per million Btu (base case) delivered to the plant gate.

Figure 3 indicates the major contributors to cost uncertainties are feed cost, capital-related charges, and labor costs. An increase in wood cost from $1.25 (base case) to $1.70 (high case) per million Btu would increase the power cost by 0.76 cent per kWh. Reducing the wood cost to $1.00 per million Btu (low case) results in a 0.42 cent per kWh decrease in the power cost. The capital-related charges were calculated based on the estimated range of plant investment shown in Figure 4. The major investment items are the boiler, turbine-generator, and utilities (including cooling towers, water treatment, etc.). Since these items are commercially proven, the cost uncertainties are principally due to equipment cost estimating and local construction cost. Labor cost uncertainties stem from the degree of plant automation and variation in local wage rates.

Oil and Char from Wood Pyrolysis

Destructive distillation is the basic principle of this technology option. A discussion of the process variants and their current development status has been reported by Kohan et al. [1] The basic process design presented in [1] provided the starting point for this evaluation.

Figure 5 shows the estimated product revenues required for the total product mix (oil and char), which range from about $7 to $12.5 per million Btu. The major contributor (almost 50 percent) to the products' cost is feedstock cost. Capital-related charges (about 25 percent) are comparatively small because of the relatively simple process scheme. The uncertainties shown in Figure 4 for feedstock charges reflect changes in wood cost from $1.25 per million Btu (base case) to $1.00 and $1.70 per million Btu for the low and high limits, respectively.

The estimated plant investment range is shown in Figure 6. The pyrolytic reactor and product recovery system are still under development, and are judged to be the plant sections with the highest degree of technical and equipment cost uncertainty. However, the investment for these sections is estimated to be only about 20 percent of the total, and thus the overall impact is somewhat moderated. As shown in Figure 5, capital-related charges uncertainty is considerably smaller than that of feedstock charges.

Figure 7 shows the impact of the char credit on the product oil revenue requirements. The overall thermal efficiency* of the process is about 74 percent. From processing 1,000 dry tons per day of wood, the plant's oil production at 1,210 barrels per day represents only about 40 percent of the product energy output; the by-product char (290 tons per day) constitutes the remaining 60 percent. Thus, the revenues generated by the char product will significantly affect the required product oil revenue, which may range from about $25 to about $32 per barrel. Thus, the char product value will likely be the key to the economic viability of this process.

Catalytic Liquefaction Process

The catalytic liquefaction of biomass is currently being developed in a DOE-sponsored pilot plant in Albany, Oregon. The history and technical development of the process have been reported previously [1 through 8]. A schematic diagram of the process is shown in Figure 7. The plant consists of two major process trains--the liquefaction train and an oxygen-blown gasification train to provide the carbon monoxide needed for the liquefaction

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* Defined as: HHV of products x 100 ÷ (HHV of feed + purchased energy for plant operation).

reaction. A catalyst (e.g., sodium carbonate) is used to promote the reaction of biomass with CO to produce an oil product. The reaction is carried out under rather severe operating conditions (2,000 to 2,500 psia at 500 to 700°F) and with excess CO. Conceptually, the catalyst, unconverted solids, and carbon monoxide are recovered from the reactor effluent stream and recycled. Engineering development work to implement the concept is currently ongoing. Various methods to introduce solid feed into the high-pressure reactor system are also being examined. The technology for oxygen-blown gasification of biomass is under development as well.

The estimated product revenue requirements for catalytic liquefaction of wood (based on the current state of process know-how) are shown in Figure 9 and range from about \$31 to \$58 per barrel. Capital-related charges represent about 35 percent of the product cost and exhibit a high degree of technical and equipment cost uncertainties because of the complexity of the process and its early stage of development. Feedstock cost is again a major contributor to product cost. The estimated plant investment for a 1,000 dry-ton-per-day wood plant ranges from \$50.1 million to \$94.2 million, as shown in Figure 10. A significant fraction (30 percent or more) of the plant investment is associated with the generation, recovery, and recycle of the carbon monoxide, suggesting that this may be an area deserving research and optimization. Utilities also require major investments. The high operating pressure and the large mechanical work requirements for compression, recycle, and wood grinding operations result in heavy steam and power loads. Optimization of process energy conservation may be quite cost effective in this case.

CONCLUSIONS

The uncertainties in the economic analysis of a process can be quantified by considering its state of development, the source of equipment cost data, the likely plant locations, and other factors. For complex processes in early stages of development, technical uncertainties tend to predominate. For proven processes, equipment and construction labor cost uncertainties are major considerations. For processes producing significant amounts of by-products, the by-product credit may determine the economic viability of the process.

FUTURE ACTIVITIES

Technology and cost analyses for the selected options (see Table 1) will be completed for the data bank. Market penetration studies and interfuel competition analysis will then draw on the developed data bank.

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Table 1

TECHNOLOGY OPTIONS SELECTED FOR DATA BANK

Thermochemical Options

- Electric power generation
- Steam generation
- Cogeneration
- Densification
- LBG (low pressure)
- IBG (low pressure)
- SNG
- Pyrolytic liquid products
- Methanol
- Catalytic liquefaction
- Hydrocarbons from plants

Biochemical Options

- Ethanol from sugar crops
- Ethanol from corn
- SNG and IBG from manure
- Ethanol from wood
- Ethanol from crop residues
- Ethanol from aquatic biomass
- SNG from kelp

FIGURE 1

PROJECT PERFORMANCE SCHEDULE

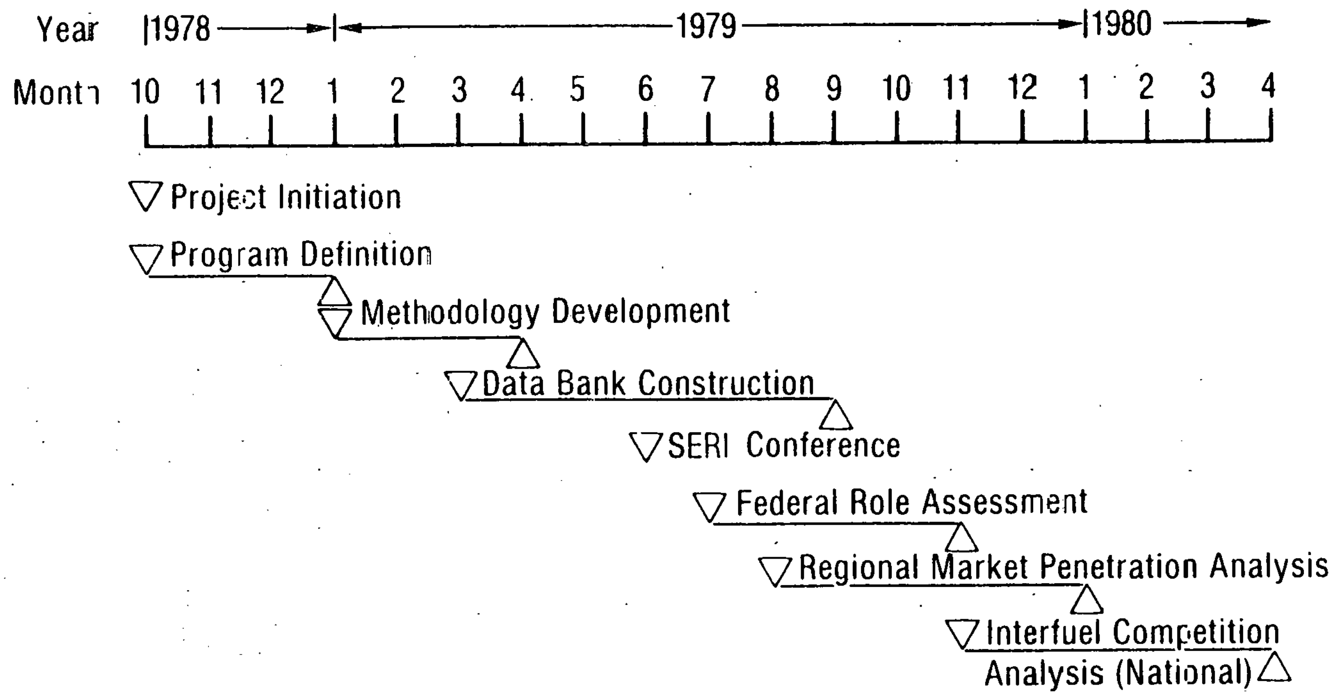


FIGURE 2

SOURCES OF UNCERTAINTIES FOR PLANT INVESTMENT ESTIMATES

State of Technology:	Basic and Applied Research	Technology and Engineering Development	Demonstration or Prototype Plant	Commercial Application
Equipment Costing:	Historical Cost Data	Vendor Quotes (General)	Firm Bids (Specific)	Installation and Testing
Level of Cost Analysis:	Scoping	Conceptual Design	Detailed Design and Engineering	Post Audit

FIGURE 3

PRODUCT REVENUE REQUIREMENT FOR POWER GENERATION BY WOOD COMBUSTION (MID-1979 DOLLARS)

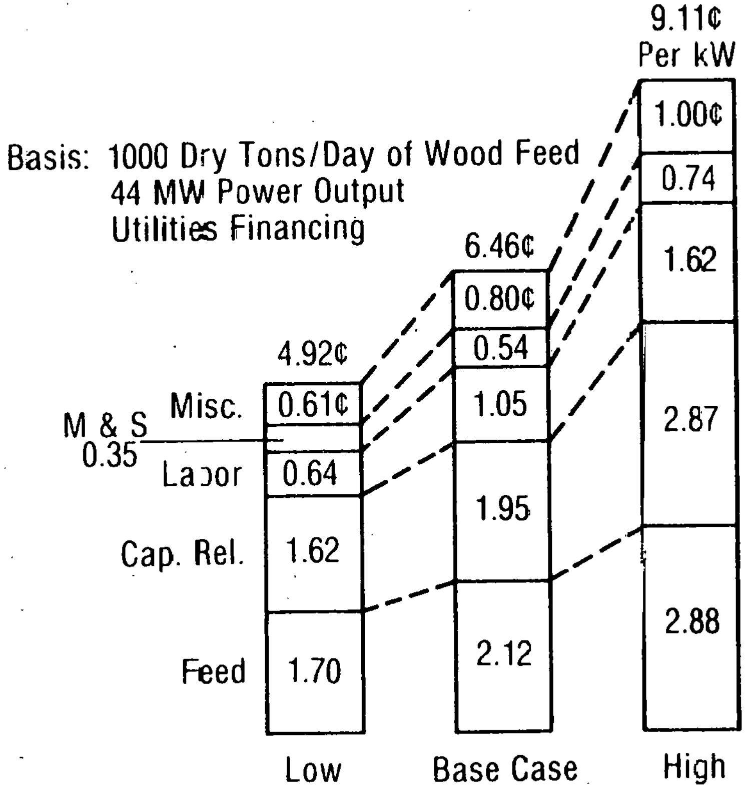
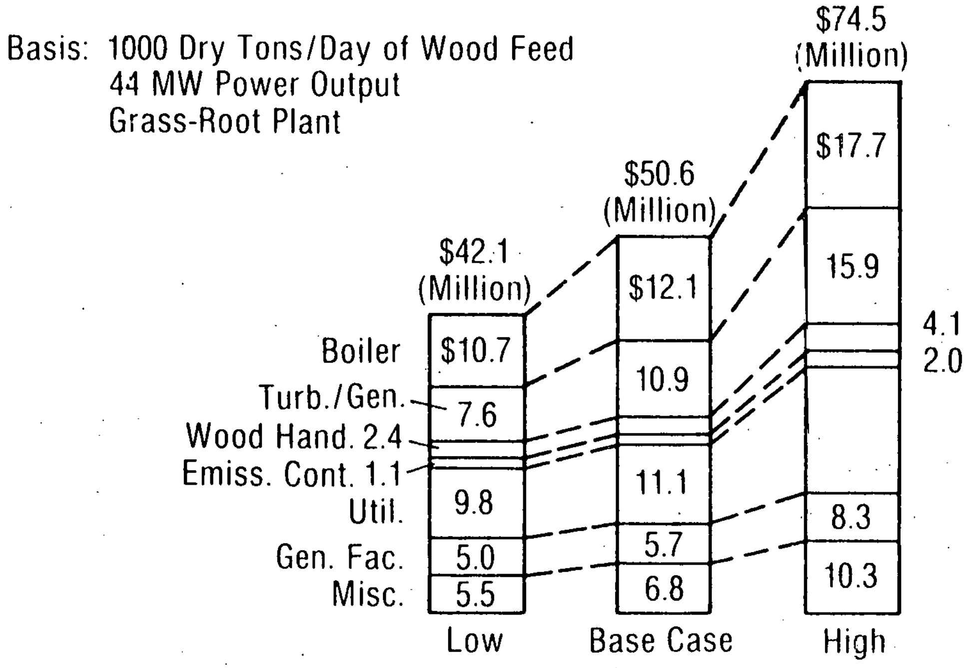


FIGURE 4

PLANT INVESTMENT ESTIMATES FOR ELECTRIC POWER GENERATION BY WOOD COMBUSTION (MID-1979 DOLLARS)



Basis: 1000 Dry Tons/Day of Wood Feed
44 MW Power Output
Grass-Root Plant

FIGURE 5

TOTAL PRODUCTS' REVENUE REQUIREMENT FROM WOOD PYROLYSIS (MID-1979 DOLLARS)

Basis: 1000 Dry Tons/Day Wood Feed
Yielding 1210 BBL/Day Oil and 290 Tons/Day Oil Char
Utilities Financing

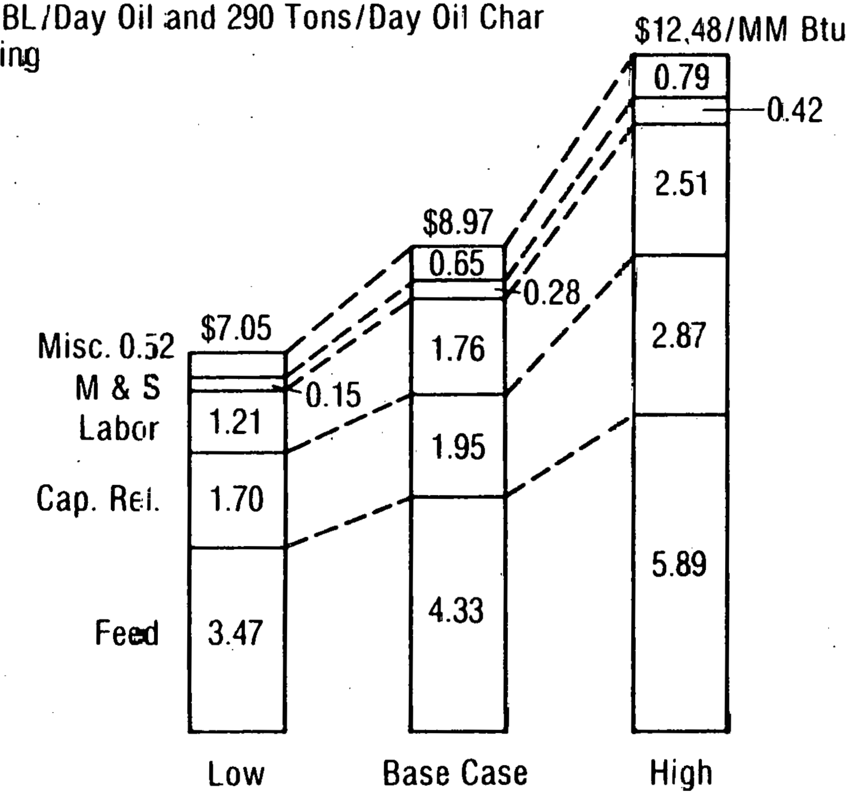
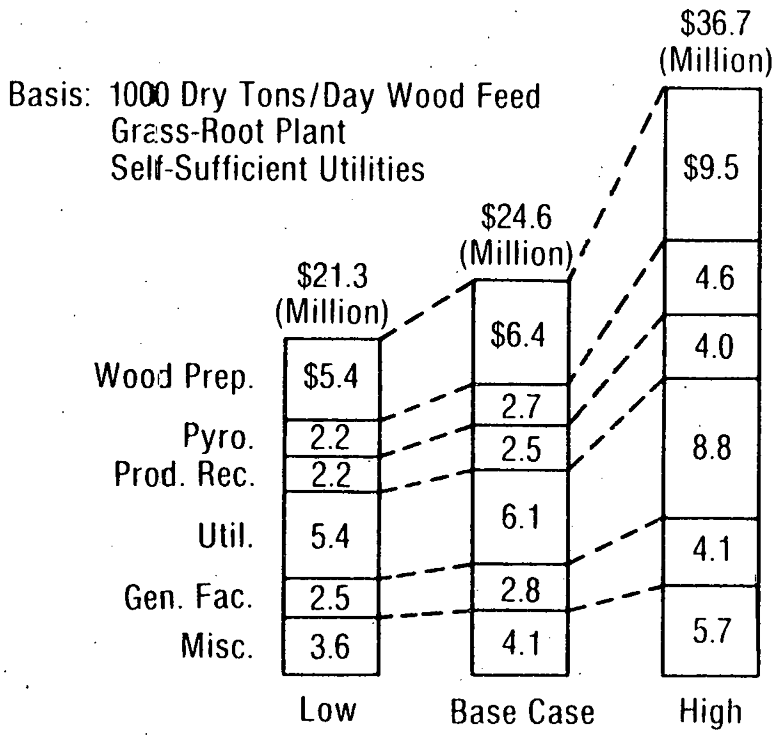


FIGURE 6

PLANT INVESTMENT ESTIMATE FOR PRODUCTION OF OIL AND CHAR FROM WOOD BY PYROLYSIS (MID-1979 DOLLARS)



Basis: 1000 Dry Tons/Day Wood Feed
Grass-Root Plant
Self-Sufficient Utilities

FIGURE 7

EFFECT OF BYPRODUCT CHAR CREDIT ON REQUIRED PRODUCT OIL REVENUES (WOOD PYROLYSIS/MID-1979 DOLLARS)

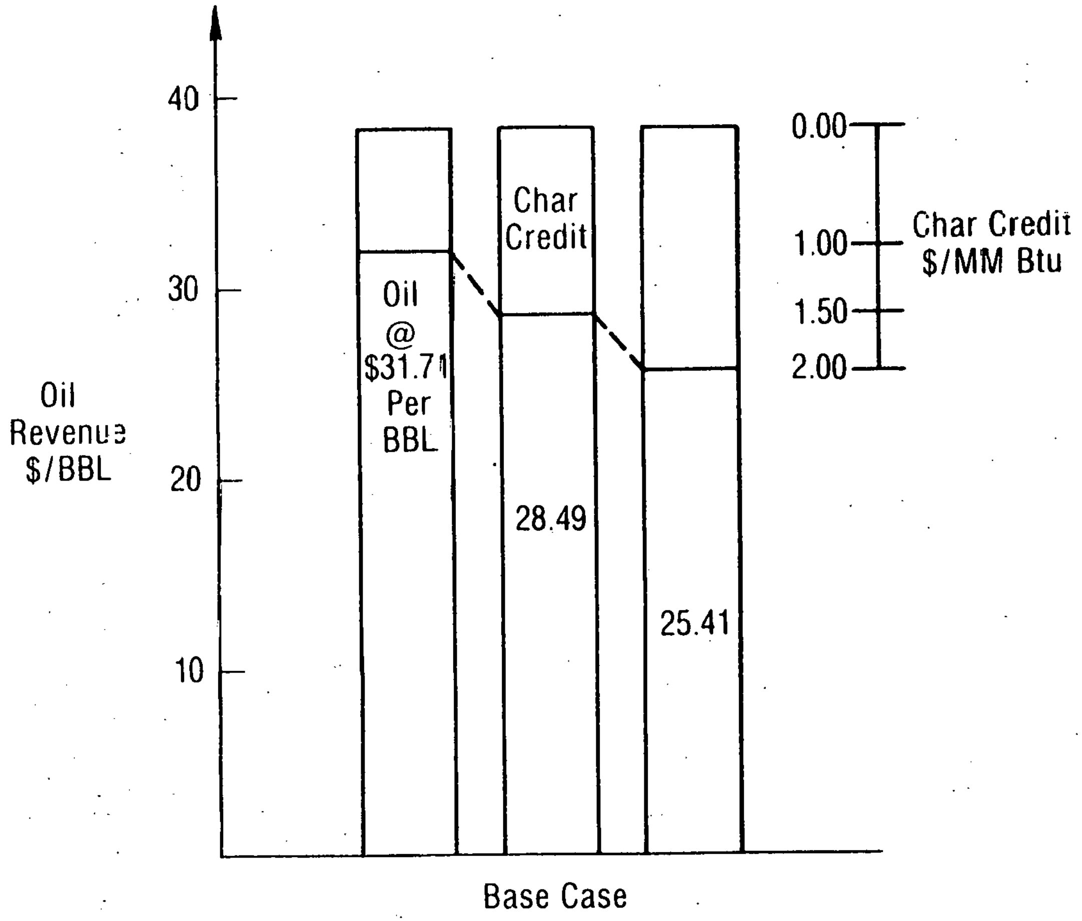


FIGURE 8

CATALYTIC LIQUEFACTION OF WOOD (PERC/ALBANY PROCESS)

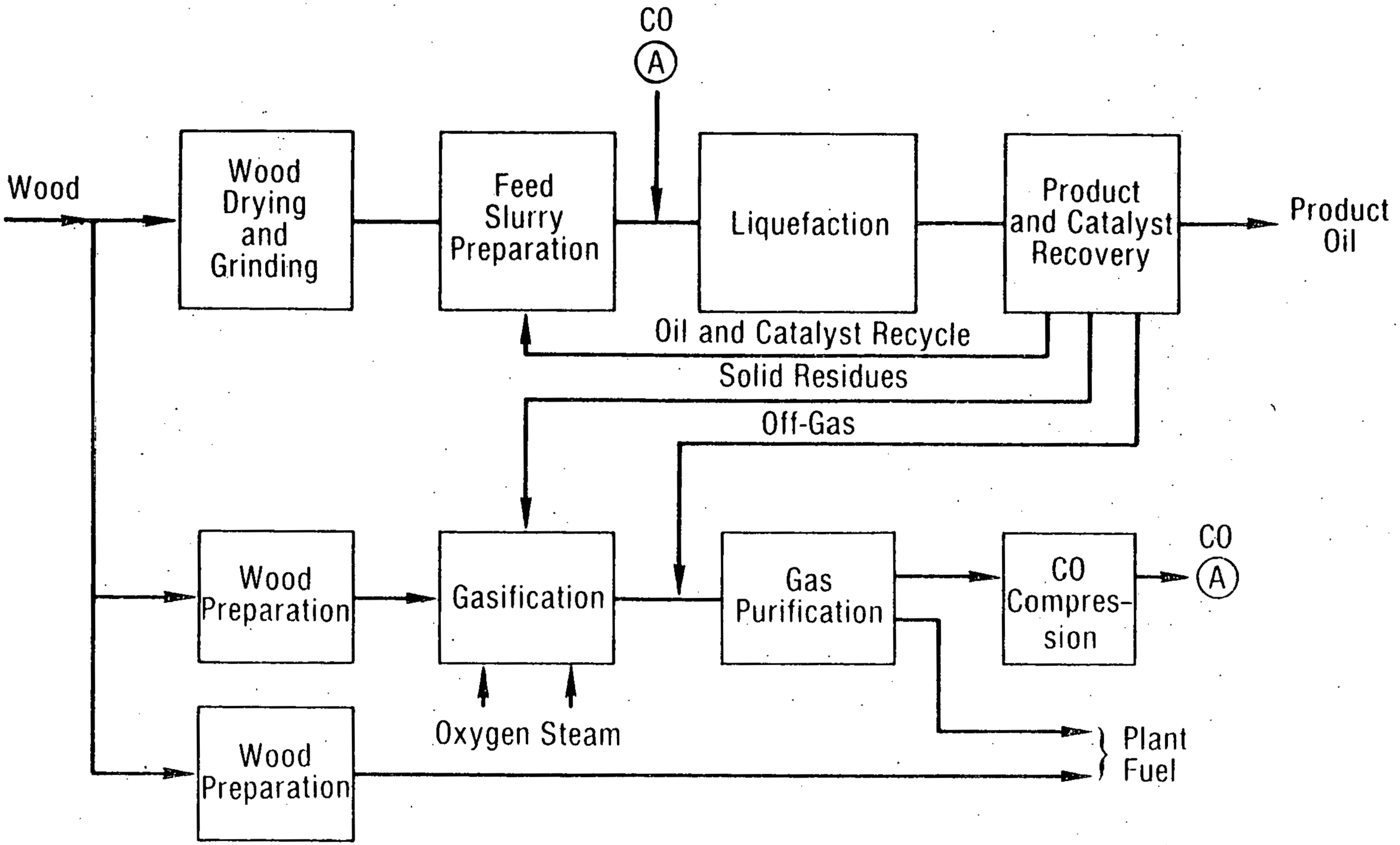


FIGURE 9

PRODUCT OIL REVENUE REQUIREMENT ESTIMATES
CATALYTIC LIQUEFACTION OF WOOD
(MID-1979 DOLLARS)

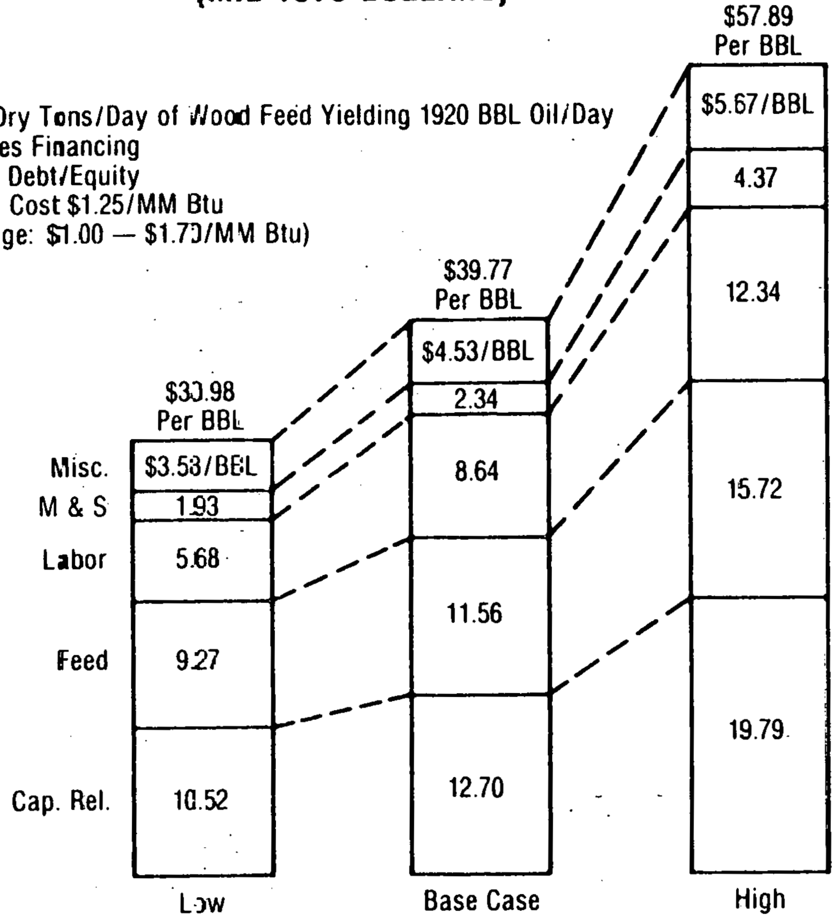
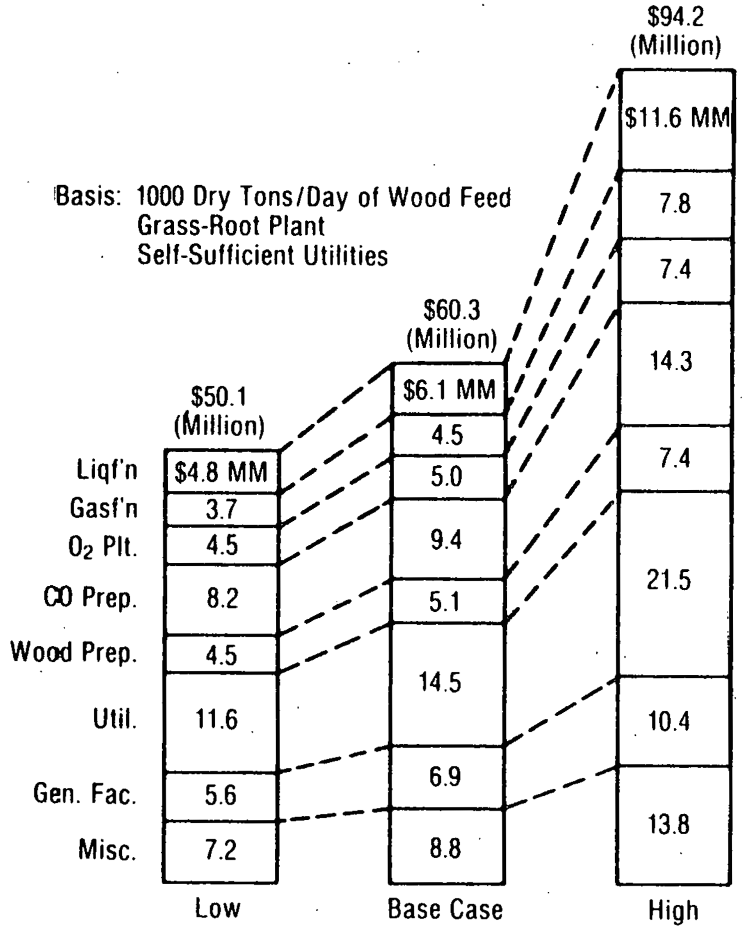

Basis: 1000 Dry Tons/Day of Wood Feed Yielding 1920 BBL Oil/Day
Utilities Financing
65/35 Debt/Equity
Wood Cost $1.25/MM Btu
(Range: $1.00 — $1.70/MM Btu)

FIGURE 10

PLANT INVESTMENT ESTIMATES FOR CATALYTIC LIQUEFACTION OF WOOD
(MID-1979 DOLLARS)



NOTES

Late Arrivals

CONCEPTUAL DESIGN OF A BIOMASS FERMENTATION FACILITY

by

D. J. O'Neil
M. K. Bery
A. R. Colcord
R. S. Roberts
D. Sondhi

Engineering Experiment Station
Georgia Institute of Technology
Atlanta, Georgia 30332

ABSTRACT

This report summarizes the initial results of Phase 1 of a planned multi-year program sponsored by the U.S. Department of Energy. The overall objective is the design, construction, and operation of a three oven-dry ton per day process development unit (PDU) to demonstrate the economic and technical feasibility of producing anhydrous ethanol from lignocellulosic biomass residues (wood, corn stover, wheat straw, principally).

The report specifically focuses on a proposed conceptual design for the PDU. It discusses biomass cost and availability in addition to unit operations: (1) pretreatment, (2) hydrolysis, (3) fermentation, (4) alcohol recovery, (5) by-product utilization, and (6) environmental monitoring. Results of a process optimization and sensitivity analysis for dilute acid hydrolysis are included.

1.0 INTRODUCTION

The conceptual design for a three oven-dry ton per day biomass fermentation facility is presented. Based on a detailed evaluation of emerging technologies and improved modifications of current technology, a mainstream process and optional unit operations and sub-systems have been selected which offer the greatest probability of success for an economic and technically feasible process for production of ethanol from lignocellulosic biomass (hardwoods, wheat straw, corn stover, etc.)

The design is intentionally flexible as stipulated in the contractual objectives. Recommendation of equipment is premised on its versatility for multi-functional application, thus enabling investigators to assess a number of process configurations while adhering to a cost-effective capital investment in the process development unit. A specific criterion in selection has been to facilitate the generation of engineering data based on the application of the results of research contractors of the U. S. Department of Energy.

The design for a total system includes the facility for evaluation of pretreatment options, for isolation of by-product streams, for evaluation of acid and enzymatic hydrolysis, sugar concentration, alternative fermentation technologies and alcohol recovery for production of absolute ethanol. In order to maintain capital costs for the PDU within reasonable limits monitoring of by-product streams will be undertaken and, for unit operations with medium-term potential (such as membrane concentration of ethanol), but which are in early stages of development, smaller-scale equipment and/or plumbing taps for later addition of full-scale (i.e., PDU-scale) equipment is recommended.

The rationale, and process economics, upon which the recommendations are based is detailed, as is a study of biomass feedstock availability.

The presentation is necessarily brief and further details may be found in the first two quarterly reports of this contract (1,2).

2.0 PROCESS DEVELOPMENT UNIT

The preliminary design of the process development unit is presented in Figure 1 as a process block diagram. The preferred mainstream process is denoted by solid lines.

3.0 BIOMASS FEEDSTOCKS AND AVAILABILITY

Georgia Tech estimates indicate that the total national biomass potential, estimated as "annually recoverable", is over 305 million dry tons, of which agricultural residues represent 34.5% and woody biomass 65.5% of the total. Assuming 10-33% usage of this total, 30.5-101.7 million tons will be available as feedstock. This volume corresponds to 87-290 plants of 1,000 ODT/day input which, exclusive of economic considerations, could be sustained in prepetuity on a nationwide basis. Wood could sustain 57-190 fermentation plants of this size.

A 1000-ODT/day wood fermentation plant, operating 340 days per year, might produce about 17 million gallons of anhydrous ethanol. Consequently, a nationwide annual production level of from 1.1 to 3.2 billion gallons of anhydrous ethanol may be possible assuming between 10-33% usage of available feedstocks of wood.

A survey of agricultural and forest experts and organizations has indicated that wood feedstocks are available, on a delivered basis, for less than $30/ODT, with costs averaging about $26/ODT in Georgia. There is no evidence to suggest that corn stover or wheat straw is available at costs approaching that of wood. Estimates for wheat straw delivered to Atlanta from within Georgia ranged from $40-80 per ton, with $60/ton being a norm. The cost of corn stubble, undelivered, was $45/ton (wood, undelivered, ranged from $7.50 - $10.00/ton). The USDA, North Central Region (Peona, Illinois) quoted wheat and oat straw at $79.75/ton and "Hay, alfalfa, heavy grass mixed, U. S. Grade No. 2" at $61.70/ton for February of 1979. (See Table 1)

The information stongly suggests that agricultural residues of a cellulose nature are not now economically attractive as a feedstock for alcohol production on a large-scale.

Figure 1

Biomass Fermentation Facility

Process Block Diagram

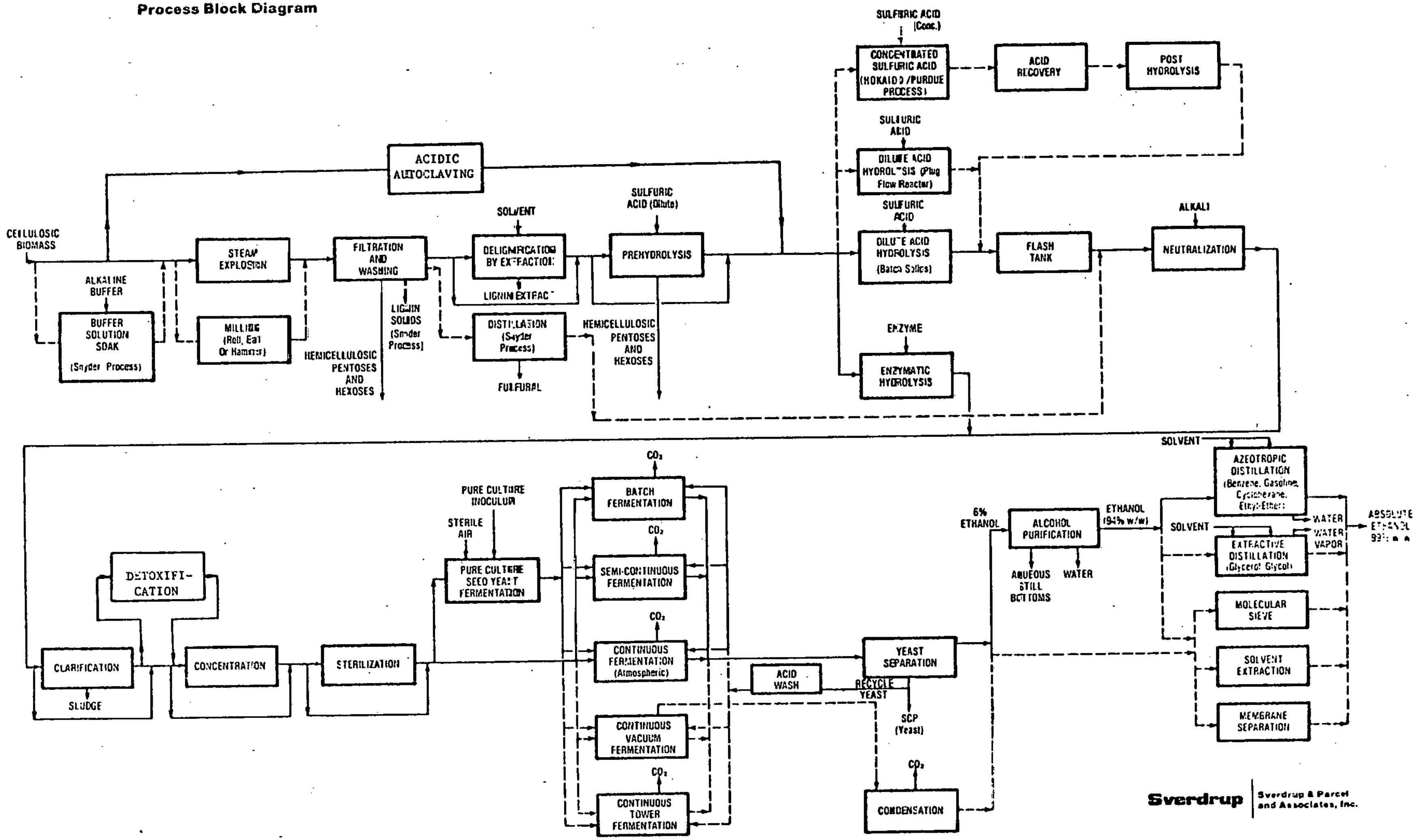


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TABLE 1

SELECTION OF BIOMASS FEEDSTOCK

WOOD	–	SUFFICIENT FOR PDU AND COMMERCIAL PLANTS	
		SAWDUST	$12-13/ODT
		CHIPS	$26-28/ODT ($15-20/ODT)
		SHAVINGS (25% MOISTURE)	$6-12/GREEN TON
WHEAT STRAW	–	SUFFICIENT FOR PDU	
		DELIVERED COST:	$40-60
CORN STOVER	–	SUFFICIENT FOR PDU	
		ESTD. DELIVERED COST:	$45/TON

4.0 PRETREATMENT

Lignocellulosic materials are essentially immune to cellulolytic enzymes unless they are pretreated. The resistance is apparently due to the close physical and/or chemical association between lignin and the cellulosic and hemicellulosic fractions and to the crystallinity of the cellulose. In order to utilize the cellulose effectively some method of pretreatment to disrupt the physical association with lignin and which reduces the molecular weight of the cellulose or renders its structure amorphous is desirable.

An investigation of various pretreatment options strongly indicates that not only disruption of the lignin-cellulosic "complexes" was preferred but also the physical isolation of the lignin from the raw material is recommended prior to hydrolysis and fermentation. The principal advantages of this approach are as follows:

(a) Increased surface area accessibility of cellulose

(b) Reduced reactor volume for hydrolysis and fermentation (Higher throughput or lower capital costs)

(c) Enhanced degree of enzymatic conversion

(d) Removal of toxic lignin degradation products

(e) Possibility for recycling during acid hydrolysis leading to effective conversion levels of 80-85% sugars from cellulose (See Process Economics).

(f) Isolation of a "pristine" lignin by-product with high value-added commercial potential.

A number of chemical pretreatments and physical pretreatments (Table 2) which focus principally on the removal of lignin from the raw material were evaluated. Similarly, pretreatments which emphasize cellulose dissolution were assessed. The regenerated cellulose (rayon) fiber industry, and more recently investigators at Purdue, have studied the latter technique for some years. Typical solvents are given in Table 3. With regard to cellulose solvents their cost and problems of toxicity render them economically and environmentally unattractive with the possible exception of concen

TABLE 2

PRETREATMENTS

I. CHEMICAL

SWELLING WITH ALKALINE AGENTS

1) SODIUM HYDROXIDE
2) AMMONIA

DELIGNIFICATION

1) PULPING
2) ClO_2
3) AMMONIUM BISULFATE
4) SO_2
5) SODIUM CHLORITE
6) ORGANIC SOLVENTS

CELLULOSE DISSOLUTION

1) TSAO/PURDUE PROCESS
2) CELLULOSE "SOLVENTS"

II. PHYSICAL

1) STEAM
    - A) IOTECH
    - B) STAKE TECHNOLOGY
    - C) GE
    - D) AUTOCLAVING
2) GRINDING
    - A) BALL MILLING
    - B) ROLL MILLING
3) IRRADIATION
    - A) GAMMA
    - B) UV
    - C) HEAT
4) FREEZING
5) PRESSURE

TABLE 3

CELLULOSE "SOLVENTS"

METAL COMPLEX SOLUTIONS

- CUOXAM OR CUAM
- CUPRIETHYLENEDIAMINE, CUENE OR CED
- CU:BUIRET:ALKALI
- COOXENE
- NIOXAM

NEUTRAL SALTS IN CONCENTRATED AQUEOUS SOLUTION

- $ZnCl_2$
- $Ca(SNC)_2$
- LiSCN, NaSNC, LiI, NaI, KI
- $K_2(HgI_4)$

STRONG BASES IN AQUEOUS SOLUTION

- LiOH
- NaOH
- NaOH + ZnO
- NaOH + BeO
- TETRAETHYLAMMONIUM-HYDROXIDE AND SOME HIGHER HOMOLOGS
- TRIMETHYLBENZYL-AMMONIUMHYDROXIDE ("TRITON B"), DIMETHYLDIBENZYL-AMMONIUMHYDROXIDE ("TRITON F")

PURDUE/TSAO SOLVENTS

- "CADOXEN" (ETHYLENEDIAMINE, CADMIUM OXIDE, H_2O)
- CMCS
- ETHYLENE DIAMINE
- SULFURIC ACID

sulfuric acid. However, this treatment type which was extensively studied as the Hokkaido Process, and which more recently is being explored at Purdue, appears to be seriously impaired by the lack of an economic recovery method for the concentrated acid.

Based on a technical and economic assessment of the various pretreatments (see Table 4), the proposed design for the process development unit might incorporate:

(a) Steam treatment with or without solvent delignification
(b) Autoclaving with dilute acid and high shear mixing
(c) Dilute acid prehydrolysis (hemicellulose removal)
(d) Cellulose dissolution.

Final selection of these pretreatments rests on the final capital costs for incorporation.

This design allows the isolation of the hemicellulosic fraction prior to cellulose hydrolysis or, alternatively, allows hemicelluloses to be carried through with cellulose for later hydrolytic and fermentation operations.

Batch extraction studies on poplar wood chips and on poplar chips which have been steam-exploded suggest solvent delignification efficiencies of 73-100% may be achievable after a one-two hour extraction of steam-exploded chips with 95% ethanol. This represents up to a five-fold increase in degree of delignification.

5.0 HYDROLYSIS

In arriving at recommended hydrolysis operations for inclusion in the conceptual design of the PDU a detailed investigation of acid and enzymatic processes was initiated. The principal processes which were investigated are given in Tables 5 and 6.

5.1 ACID HYDROLYSIS

A process optimization study, which will be detailed elsewhere (3), was undertaken to predict the cost of hexose production by dilute acid hydrolysis using a continuously stirred reactor (CSTR) and a plug flow reactor (PFR).

TABLE 4

ESTD. COST OF SELECTED PRETREATMENTS

TREATMENT	POWER	POWER COST	
		¢/LB	$/TON
BALL MILLING**	1.4 KW-HR.	4.9	$ 98
2-ROLL MILLING**	0.25 KW-HR.	0.90	$ 18
	0.66 KW-HR.	2.3	$ 46
STEAM EXPLOSION	0.7-1.3 KW-HR. STEAM***	0.28-0.52	$5.60-10.40
PULPING			224-242 (PRODN
			320-345 (PRICE

* ELECTRIC POWER @ 3.5¢/KW-HR.

** MESH SIZE: 53 MICRONS

*** $4.00/1000 LB. STEAM

TABLE 5

ACID HYDROLYSIS

CONCENTRATED SULFURIC

- HOKKAIDO
- THROUGH DRYING PROCESS (KOBAYASHI)

CONCENTRATED HYDROCHLORIC

- BERGIUS-RHEINAU
- UDIC-RHEINAU
- PRODOR PROCESS (VAPOR)
- DARBOVEN
- HERENG
- NOGUCHI-CHISSO
- BATTELLE (GENEVA)

DILUTE SULFURIC

- SCHOLLER
- MADISON
- TVA
- RUSSIAN
- GRETHLEIN
- PURDUE

WEAK ACID (AUTOHYDROLYSIS)

- SNYDER

TABLE 6

ENZYMATIC HYDROLYSIS

- NATICK
- BERKELEY
- GE - UNIV. OF PENNSYLVANIA
- MIT
- GULF
- RUTGERS
- RPI
- SWEDISH FOREST PRODUCT LAB
- INDIAN INSTITUTE OF TECHNOLOGY
- MIYAZAKI UNIVERSITY

Previously published reports have estimated the cost of hexose produced by the dilute acid hydrolysis of cellulose (4,5). Both studies employed a design which was based on operating conditions which maximized hexose production exiting the reactor. Operating conditions which maximize hexose concentration would also represent the minimum cost for hexose produced, if the cost of the cellulose raw materials were negligible. In fact, the contributions of the cost of cellulose raw materials to the cost of hexose production is substantial (1). This analysis has considered the effect of raw material costs in order to project minimized hexose costs.

The analysis was modelled on a process which assumes a wholly cellulose feedstock entering the hydrolytic reactor where it is reacted with dilute sulfuric acid, the sugar solution and unreacted cellulose slurry exiting the reactor being neutralized by lime, filtered to recycle the unreacted cellulose, with the dilute hexose solution being concentrated by multi-effect evaporation (Figure 2).

The major cost elements of producing hexose by this process is given as:

$$\frac{\text{cost}}{\text{kg hexose}} = \frac{\text{Cost of cellulose consumed}}{\text{kg hexose}} + \frac{\text{cost of acid-lime}}{\text{kg hexose}} + \frac{\text{cost of reactor}}{\text{kg hexose}}$$

$$+ \frac{\text{concentrator cost}}{\text{kg hexose}} + \frac{\text{cellulose recycle cost}}{\text{kg hexose}} \qquad \text{(Eqn. 1)}$$

The cost function (Eqn. 1) can be expressed symbolically as follows:

$$\frac{\text{cost}}{\text{kg hexose}} = \frac{v_o(C'_{co} - \theta C'_c)\$_c}{v_o C_g} + \frac{v_o C_A \$_A}{C_g v_o} + \frac{V\$_R}{C_g v_o} + (/C_g - 1/C_g{}^*) \quad \$_{conc} +$$

$$\frac{\theta C'_c \; v_o \quad \$_{recycle}}{C_g \quad v_o} \qquad \text{(Eqn. 2)}$$

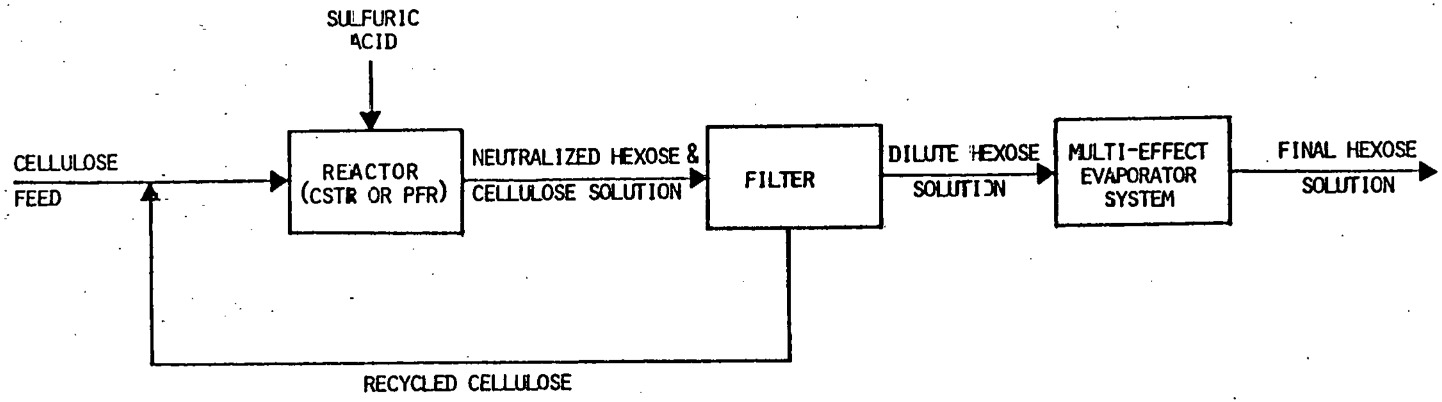


Figure 2. Process to Produce a Hexose Solution by the Dilute Acid Hydrolysis of Cellulose

where $\$_C$ = cost of cellulose, \$/kg equivalent hexose

$\$_A$ = cost of acid-lime, \$/kg acid

$\$_R$ = cost of reactor (fixed + operating charges), $\$/m^3$ reactor-hour

$\$_{conc}$ = concentrator charges, $\$/m^3$ evaporated

$\$_{recycle}$ = recycle charges, \$/kg cellulose recycled

where v_o = volumetric flowrate, m^3/hr

θ = fraction of cellulose exiting the reactor which can be recovered and recycled

C_A = concentration of acid, k/gm^3

V = volume of reactor, m^3

$C_g{}^*$ = desired concentration of hexose, exit concentrator, kg/m^3

The assumptions made in constructing the model process are shown in Table 7. The parameters which were investigated are given in Table 8.

The kinetics for the saccharification of cellulose and the degradation of glucose reported by Saeman were used to model the reactions (6). This includes rate expressions both for cellulose disappearance and glucose degradation.

A separate optimization was developed for both types of reactor, CSTR and PFR.

The cost optimization equations for a CSTR and PFR were derived. Both minimized cost functions were derived from differentation with respect to the residence time, (hours). The optimum residence times for both reactor types were obtained as functions of temperature, acid concentration, feedstock concentration, etc. to obtain the minimum cost of production of hexose solutions. A computer program was written to evaluate both optimum residence time and minimized hexose cost.

Under standardized "baseline" operating conditions the minimized hexose cost and the contributions of the primary cost elements of the model process are given in Table 9. It was determined that the use of a plug flow reactor is far superior to a continuously stirred reactor in the dilute acid hydrolysis process. An expanded sensitivity analysis for PFR operation indicated that sugar production costs could be lowered to 4.7¢/lb for a 20% feedstock slurry input and to 4.4¢/lb for a 30% slurry,

TABLE 7

PROCESS ASSUMPTIONS

1. Cellulose feed free of hemicellulose and lignin.
2. Douglas fir kinetics (Saeman).
3. Cost of cellulose based on wood cost at 45% cellulose content.
4. No credit for lignin or hemicellulose.
5. 90% recovery of unreacted cellulose from reactor product stream.
6. Lime neutralization of hydrolysis acid solution.
7. Multiple-effect evaporation.
8. 13% final hexose concentration.

TABLE 8

MODEL PARAMETERS

1. Reactor types (CSTR, PFR, and Fixed Bed).
2. Cost of cellulose (wood).
3. Concentration of cellulose feed.
4. Concentration of sulfuric acid.
5. Temperature
6. Cost of Reactor.
7. Cost of sulfuric acid and the equivalent lime for neutralization.
8. Cost of recycling unreacted cellulose.
9. Cost of concentrating hexose solution.
10. Concentration of final hexose solution.

TABLE 9. Minimized Hexose Costs Under Base Line Operating Conditions*

	CSTR		PFR	
Unit Costs	$/kg Hexose	Percent of Total Cost	$/kg Hexose	Percent of Total Cost
Cellulose Raw Material Cost	.1040	64.8	.0906	72.7
Acid and Equivalent Lime Cost	.0217	13.6	.0150	12.0
Reactor Cost	.0002	.1	.0002	.1
Hexose Concentration Cost	.0264	16.5	.0150	12.0
Cellulose Recycle Cost	.0080	5.0	.0040	3.2
TOTAL ($/kg)	.160	100.0	.125	100.0
TOTAL ($/lb)	.073		.057	

*10% Slurry; 1% Sulfuric Acid; 210^{o}C; $30/ODT Wood; 13% Sugar Solution

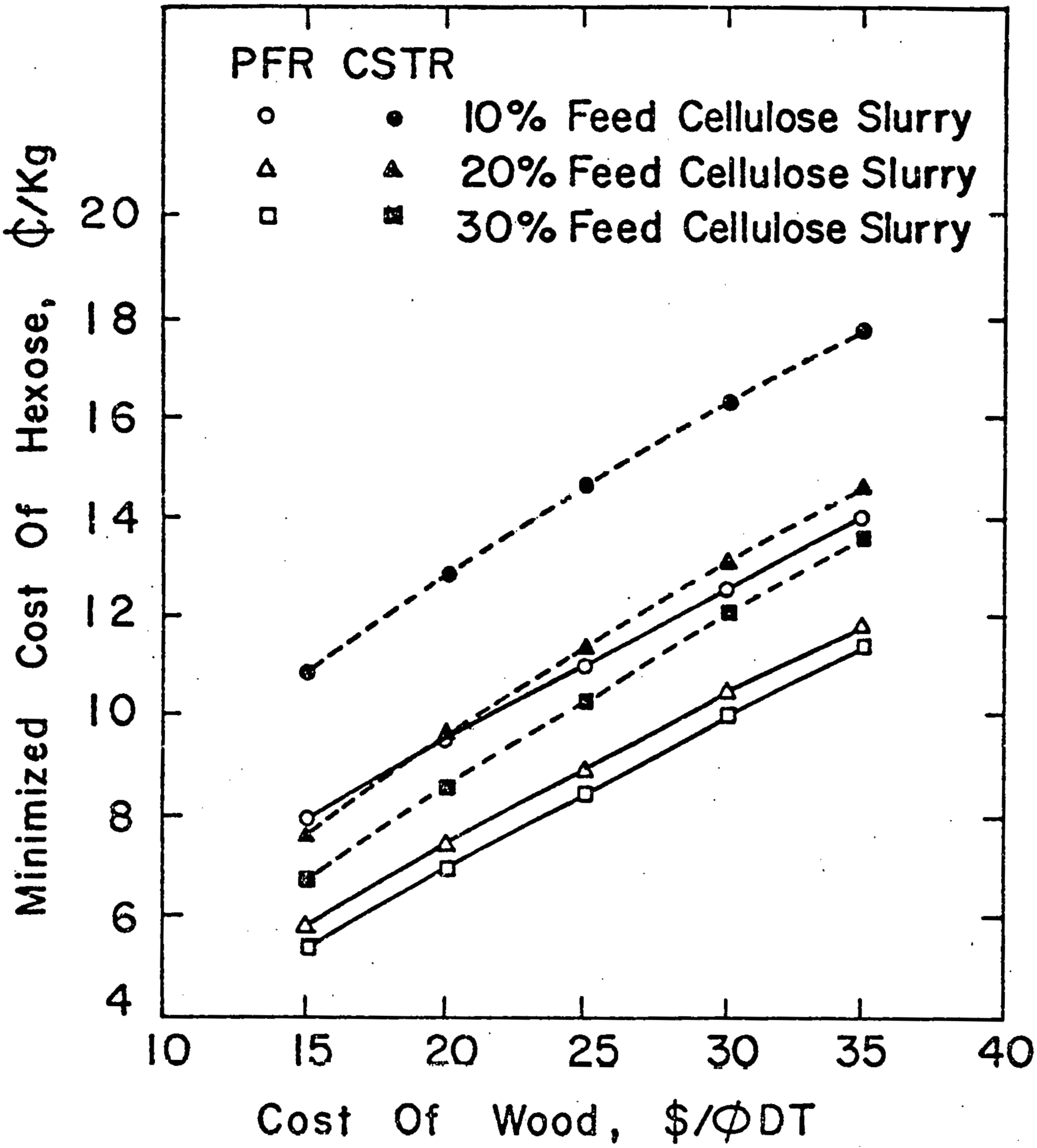


Figure 3 . Effect of Wood-Cost on the Minimized Cost of Producing a 13% Hexose Solution in a CSTR and a PFR System (see Table 4-3 for other operating conditions).

(Table 10). Whether a plug flow reactor can actually operate at the slurry concentration must be determined experimentally. The data in Table 10, nevertheless, do point out the significant cost contribution made by the raw material.

Ethanol at \$1.20/gallon of 95% alcohol (\$15/MMBtu) is the equivalent of \$0.169/kg. The estimated cost of converting hexose to 95% ethanol is \$0.044/kg of hexose. Thus, it appears that the cost of ethanol (95%) can be held to \$1.20/gal. if hexose can be produced from cellulose for \$0.125/kg hexose. The results of this study (Table 10 and Figure 3) indicate that an optimized PFR recycle system will meet this cost criterion.

A similar optimization study is underway for a fixed bed reactor (FBR) system and current results indicate that the FBR economics possibly will be superior to those projected for the PFR. Based on the proven technological feasibility of FBR systems, and the modern advances in materials of construction and fluid dynamics, our conceptual design favors the FBR, with a strong recommendation for incorporation of an optional PFR unit.

5.2 ENZYMATIC HYDROLYSIS

The benefit of sensitivity analyses, as was undertaken with dilute acid hydrolysis, is to predict the advantages and relative merits of production with different reactor designs. Central to the success of such analyses is the reliability of an accurate chemical model and reaction kinetics. Unfortunately, in the case of enzymatic hydrolysis, there is a real lack of data to effect such a study. Our studies have confirmed that a number of organisms have a significant and increasing level of cellulolytic activity and that major improvements in cellulolytic activity might be expected in the next few years by progressive strain selection. However, the dearth of kinetic information and engineering design makes it extremely difficult to specify hardware which could be incorporated into the PDU design, although its utility in future economic production processes based on enzymtic hydrolysis is anticipated.

TABLE 10

Effect of Feed Concentration on Hexose Production Cost
For PFR Operation

Production Cost Components	Slurry Concentration		
	10%	20%	30%
Cellulose Cost	72.7%	86.5	90.6
Cost of Acid and Equivalent Lime	12.0	6.7	4.2
Reactor Cost	0.1	0.1	0.1
Hexose Concentration	12.0	2.3	0.0
Cellulose Recycle	3.2	4.4	5.1
TOTAL	100.0	100.0	100.0
Hexose Production Cost			
($/kg)	0.125	0.103	0.097
($/lb)	0.057	0.047	0.044
Hexose Soln. Conc. Exiting Reactor	5.5%	11.0%	16.6%

As an initial determination our design has excluded the incorporation of an ancillary enzyme production sub-system. Future studies in the PDU will undoubtedly explore more than one enzyme system and sizing and specification of enzyme production units are influenced by the nature and rate of production of an enzyme. It has been further judged that a recommended enzyme hydrolysis reactor should emphasize versatility in design.

At this stage, simultaneous saccharification and fermentation (SSF), as represented by the Gulf process, appears to have significant advantages, particularly on an economic basis, relative to multistage enzymatic hydrolysis and fermentation processes such as employed in the Natick process. According to W. S. Fong (7), in the multi-stage processes, the enzyme production and hydrolysis plant sections represent the most costly plant sections, accounting for 31% and 24% respectively, of the fixed capital investment. This major capital investment is reduced in SSF operation. According to G. H. Emert,(8) in an SSF process the ethanol yield is increased by as much as 40% by immediate conversion of glucose on formation, thus eliminating glucose inhibition of hydrolysis. SSF eliminates the need for separate hydrolysis and fermentation vessels and separate stages for recovery and concentration of enzyme.

An investigation of bioreactor designs included those identified in Table II. Our choice of design is based on a CSTR. It can be specified in a variety of sizes and configurations and purchased conveniently.

Its flexibility and versatility is apparent in the ability to operate in batch or continuous modes. With a large height-to-diameter ratio, within limits, it may be adapted for operation as a fluidized bed reactor. With appropriate ancillary piping it may be used as a fixed bed-type reactor. The PDU design allows the preferred SSF mode of operation also.

TABLE II

REACTOR DESIGNS FOR ENZYMATIC HYDROLYSIS

CSTR	
	Natick
	Indian Istitute of Technology
	University of California - Berkeley
	GE - University of Pennsylvania (Counter-Current)
Fixed Bed	
	Tokyo Institute of Technology
	Massachusetts Institute of Technology
	Dynatech R and D Company
Fluidized Bed	
	Tokyo Institute of Technology
	Oakridge National Laboratory

6.0 FERMENTATION

A number of designs were evaluated for the fermentation operation (Table 12).

Table 12

Fermentation Processes

Open-Vat Fermentation
Deep-Vat Fermentation
Tower Fermentation
Vacuum Fermentation
Deep Jet Fermentation

It was decided to select the deep vat fermenter which may be used as an open vat fermenter, to evaluate if a strict asepsis is needed. It can also be used as a deep vat fermenter with or without asepsis, and to study the extent of power requirement depending on the degree of agitation. The fermenter can also be used in SSF type operation. Therefore, this universal design is quite flexible and can be operated in three different modes. The same fermenter may also be used for continuous vacuum fermentation as well as simultaneous saccharification and fermentation type operations. The flow diagram (Fig. 4) of a conventional continuous fermentation type operation indicates that carbon dioxide produced during the course of fermentation causes an extra load on the vacuum system. This is undesirable. The proposed PDU unit

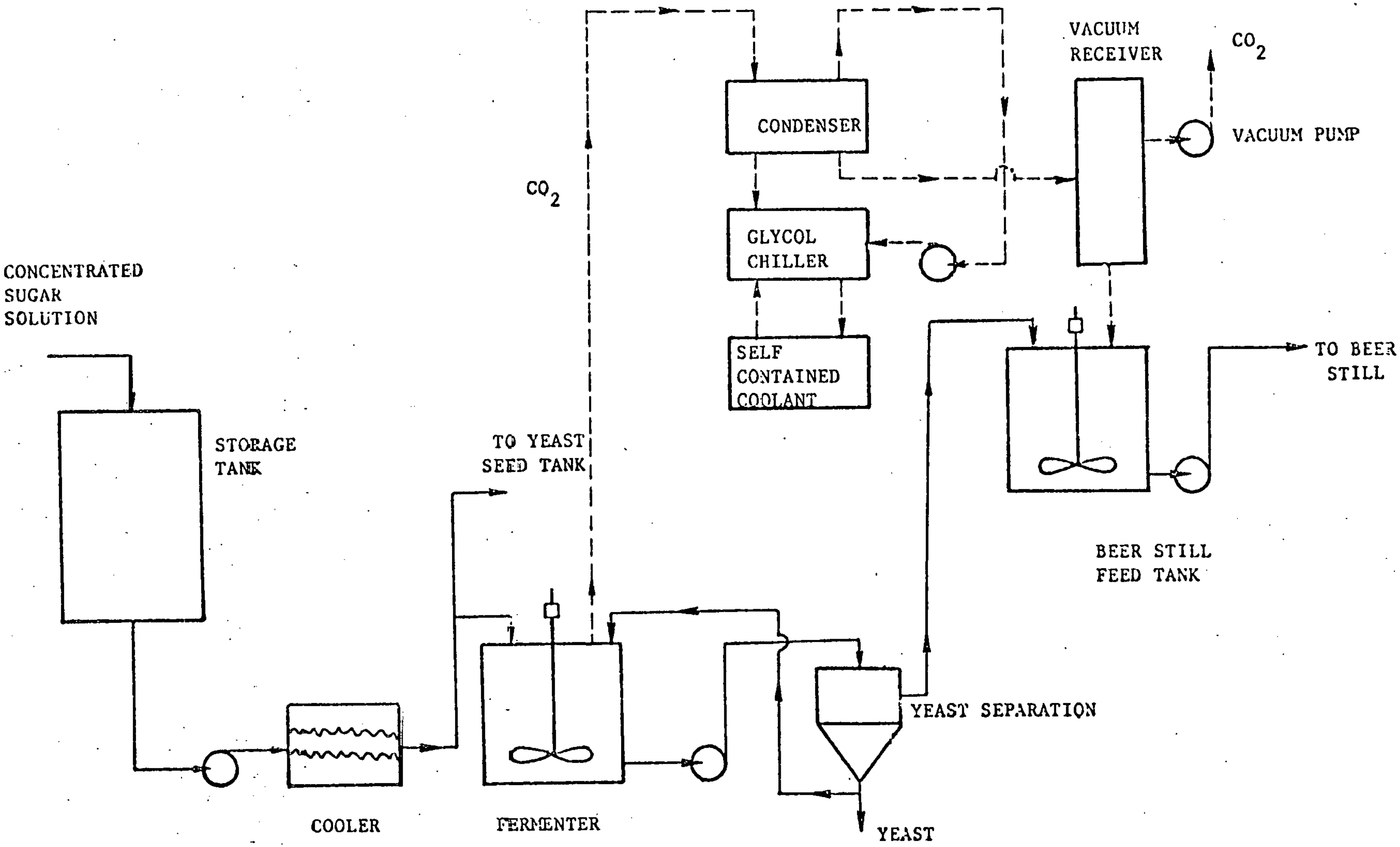


Figure 4 CONTINUOUS VACUUM FERMENTATION (CONVENTIONAL)

will include the provision for SSF and continuous fermentation with indirect vacuum (Fig. 5). This is superior to the classical vacuum fermentation; first, carbon dioxide is recovered separately without causing any extra load on the vacuum system; second, since the flashed liquor is recycled it does reduce the requirement of cooling water for the fermenter. Finally, since the fermentation itself is carried out under normal conditions, its control is much simpler as compared with the conventional vacuum fermentation.

The tower fermenter is a desirable option. A decision for its inclusion to handle full scale throughput rate will be controlled by its cost but a small scale unit to process a side stream may be desirable for the following reasons:

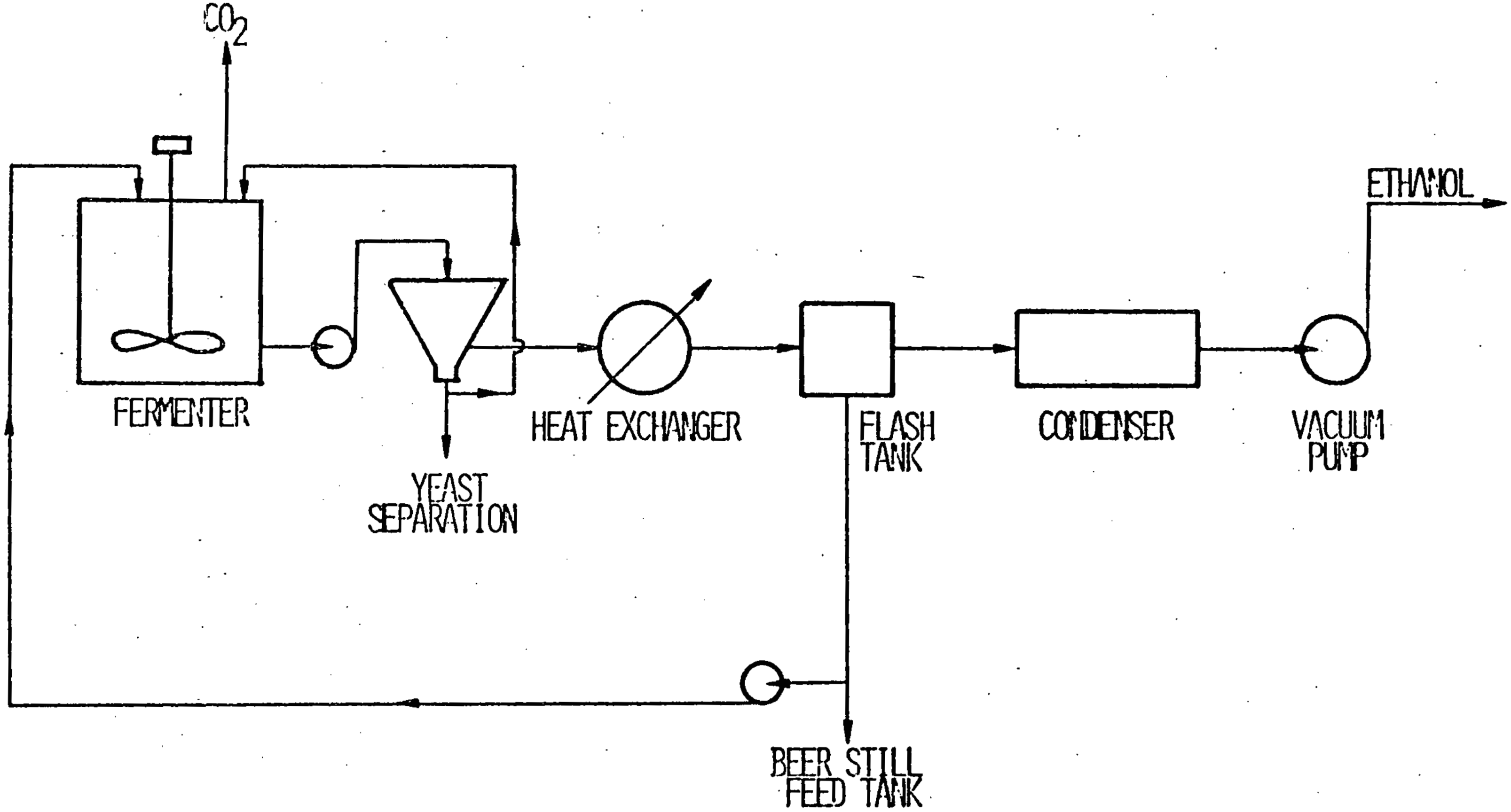
1. Considerable increase in the efficiency is reported by the European brewing industry. Fermentation times as low as 2 hours, as compared with the batch fermentation times of 3-5 days, are reported.
2. The unit is very simple to maintain and operate.
3. Because of its shape the tower fermenter occupies a much smaller floor space as compared with conventional CSTR type system.

7.0 ETHANOL RECOVERY

It is well-recognized that the production of 95% ethanol is relatively easy by conventional distillation but that a costly energy-intensive azeotropic distillation is required to produce anhydrous ethanol. Removal of water is necessary for blending with gasoline.

The filtered liquor after fermentation contains not only ethanol and water, but also small quantities of many other relatively volatile components, including low boiling-point acetaldehyde and esters and a complex mixture generally known as fusel oil. This has a higher boiling point than ethanol or its aqueous azeotrope

FIGURE 5. CONTINUOUS FERMENTATION WITH INDIRECT VACUUM (PDU)



and contains many higher alcohols including amyl, isopropyl and butyl, together with their esters. It is impossible to obtain pure ethanol in a single simple fractionation, but it can be obtained as the azeotrope containing 4.4% water. The complete operation is usually split into several stages:

1. The beer is stirpped with steam to yield a dilute ethanol which still contains part of the unwanted volatiles.
2. This is fed through a heat exchanger to a distillation column, from which separate fractions can be taken off including:
    (a) Volatile esters and aldehydes
    (b) fusel oil
    (c) ethanol-water mixture containing about 25% ethanol
    (d) low fraction, mainly surplus water.
3. The dilute ethanol is again taken through a heat exchanger into another column from which the azeotrope can be taken off near the top and water from the bottom. This gives an extra opportunity for removal of last traces of other volatiles.

A number of methods have been reviewed for production of absolute ethanol from the azeotrope (Table 13).

The pros and cons associated with each process are detailed elsewhere (1,2). It appears that azeotropic distillation, while still energy-intensive, provides the most economic means of production of absolute alcohol. While, classical commercial schemes utilize benzene as the entrainer in azeotropic distillation, other azeotrope formers have been used effectively with significant reductions in energy consumption. Cyclohexane has been substituted for benzene because of environmental control pressures. Gasoline has been suggested. The use of diethyl ether leads to a reduction of net dehydration energy.

TABLE 13

ALCOHOL RECOVERY

Azeotropic Distillation

Extractive Distillation

Proprietary Distillation Processes

Vulcan Cincinnati

Vogelbusch

A.C.R. Process Corporation

Absorption of Water on Salts

Vacuum Distillation (77 mm Hg)

Freeze Concentration

Solvent Extraction

Membrane Separation

Chemical and Physical Adsorption

Research at the University of California at Berkeley indicates that a reduced pressure system will lead to further reductions in energy consumption during the distillation process. In general, 40-60% reduction in the energy consumption associated with benzene-based azeotropic distillation are claimed by a number of proprietary processes.

Anhydrous ethanol can also be produced by the use of molecular sieves, solvent extraction, membrane separation, etc. but all appear to be economically unattractive at this time. The PDU design has the capability to incorporate new developments by these systems, nevertheless. Incorporation and testing of these alternate processes is visualized at a later date and at smaller-than-PDU scale.

8.0 BY-PRODUCT UTILIZATION

During the operation of the proposed Process Development Unit (PDU) all the streams producing any product other than ethanol would be monitored and evaluated for their commercial value. It is not intended to propose a separate pilot plant at this stage to process each of these streams to produce derivative products with the exception of pentoses. An assessment of the utility of by-products in terms of their fuel or chemical value is anticipated as part of any manufacturing economics analysis.

The various streams are described as follows:

After the delignification stage the attempt would be made to produce lignin in the dry form. This "pure" form of lignin will be evaluated for a variety of high-value added products as chemical feedstock, and for its process fuel value.

During the prehydrolysis stage where hemicellulose is converted into C_5 and C_6 sugars, the quantities of C_5 and C_6 will be dependent on the type or the source of hemicellulose. It may be very profitable to produce a single cell protein (SCP), furfural or xylitol from C_5 sugars. C_5 can also be converted into 2,3-butanediols (Purdue, Auburn). Therefore, any future breakthrough can easily be assessed in the proposed PDU for the total conversion of hemicellulose and cellulose into liquid fuels or chemicals.

The fermentation stage produces two by-products; first, carbon dioxide which can only be monitored in terms of its value as dry ice, or as a gas for the food or beverage industry. Liquified CO_2 is also a very valuable solvent to extract various organic components from beans (decaffeination of coffee beans). It has been proven to be relatively cheaper to regenerate the carbon by liquified CO_2 as compared with the high temperature conventional furnace system. Secondly, the excess amount of yeast produced might be marketable as SCP or a source of vitamin B complex. At this time we intend to evaluate these by-product streams by monitoring only.

9.0 WASTE TREATMENT MONITORING

The PDU system will produce three principal kinds of wastes:

1. Solid waste or sludge produced by the neutralization of acid stream. This will be separated and disposed of in a conventional manner.
2. Very dilute liquid waste streams generated by wash water during the various stages of operation.
3. Organic liquid waste streams, generated during enzymatic hydrolysis and fermentation operations.

The second and third items described above would be of main concern to us. In PDU operation, it is planned to collect samples of every effluent stream on a 24 hour basis, homogenize these samples and then analyze the samples for BOD, COD, N, P, K, SO_4, NO_3, carbohydrates, etc. The above analyses will help in generating the data for scaled up operations (e.g., 2000 ODT/day plant). We will also be able to evaluate the possibility of recycling these streams and their effects on the overall operation of the PDU system. It is also desirable to have two small tanks where one can be used as an effluent holding tank to homogenize the waste streams and the other tank could be used as aeration tank to reduce the BOD values of the effluent. The above information regarding agitation, aeration and the power requirement to treat the organic waste is essential for a commercial scale design.

10.0 SUMMARY

This report has summarized a recommended conceptual design for a three oven-dry ton per day cellulose biomass-to-ethanol process development unit (PDU). The ultimate goal is to construct and operate the PDU in such a way that real and reliable engineering and economic data can be generated so that a commercial process can be designed and subsequently tested.

Immediate follow-on work is addressing detailed engineering design and detailed cost estimates for the PDU and a one-thousand ton per day plant.

ACKNOWLEDGEMENTS

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LOW COST METHANE GENERATION ON SMALL FARMS

by

W. J. Jewell, S. Dell'Orto, K. J. Fanfoni, T. D. Hayes
A. P. Leuschner and D. F. Sherman
Department of Agricultural Engineering
Cornell University
Ithaca, New York 14853

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ON SMALL FARMS

W. J. Jewell, S. Dell'Orto, K. J. Fanfoni, T. D. Hayes,
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ABSTRACT

This paper examines the effect of farm size on the technical and economic feasibility of methane generation. After the rationale is developed for considering small farm installation of anaerobic fermentors, the technical requirements and information needs to implement low cost biogas generation systems are identified. This report summarizes the small farm fermentor development program that has been on-going at Cornell University since 1974.

The large majority of agricultural residues are generated on relatively small agricultural operations. More than three-quarters of all organic residues are generated at continuous rates of less than 2700 kg per day (3 tons per day). The average size of the U.S. farm was 364 ha in 1974, and 80% less than 204 ha. For this reason, on-site generation and utilization of energy is assumed to be required in many cases.

The potential of small farm methane generation is significant but it is essential that low-cost systems be developed. Laboratory, pilot scale, and full scale demonstration studies have identified a low-cost, easily constructed approach that may be cost-effective for operations as small as the 40-cow dairy. This system uses an unmixed plug flow design with the methane generation reactor constructed with soil supported rubber-like pond liner material. Full-scale operation for over a year, including the coldest New York winter, has shown that this unit is a feasible alternative.

Finally, comparison of the total investment required for widespread adoption of methane generation technology to nuclear, fossil fuel, and massive installation of solar power indicates that it could be highly competitive. Installation of units on all dairies in New York State alone would result in the production of replacement energy equivalents about 500 megawatts per day of energy generation capacity. Equivalent installation costs for small methane generation systems are estimated to be $200 to $600 per kilowatt of energy generation capacity, whereas other sources vary from $300 to $6,000 per kilowatt of energy generation.

LOW COST METHANE GENERATION
ON SMALL FARMS

INTRODUCTION

Organic residues contributed from all farm land in the U.S. could be used to produce annually more than $30 billion worth of energy.[1] Increased plant production and utilization of presently idle land has a potential of increasing this amount to as much as one-third of the total U.S. energy needs exceeding the amount of all imported fuels. However, major questions relate to the availability and economics of conversion of organics to energy. Animal manures represent a significant biomass that is available and already collected, in many areas. Unless properly handled these organics represent a pollution hazard. The feasibility of using these materials as a source of methane via anaerobic fermentation is the topic of this paper.

Application of anaerobic fermentation on individual farms may be difficult to justify because of the small size and availability of capital and operational skills. But it may also be the area where this method of clean and renewable fuel generation holds greatest potential for contributing to the energy problem. This paper will summarize some of the on-going work in this area by a multidisciplinary team of researchers at Cornell University since 1974 (Jewell, et al., 1976; Jewell, et al., 1978). The main emphasis on this team has been to develop a practical and economically feasible small farm methane generation system.

Biogas (a mixture of methane [60% by volume] and carbon dioxide) is perhaps the highest quality of energy that is easily derived from photosynthetic material at a reasonable cost. It can directly replace many forms of energy; i.e., liquid and gaseous forms. Because it has been known for over a hundred years, the basic concepts of methane generation from organics is well known. The agricultural community is open-minded and appears to be ready to adopt the technology as soon as it is available, and recommended for implementation with accompanying financial and technical support mechanisms. Unfortunately, relatively little technology and financial aid exist to implement methane production from agricultural residues, particularly in small

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[1] This assumes that one ton (dry) per acre is collected, the organics are 80 percent organic (20 percent ash) and 50 percent of the organics are converted to energy, leaving the refractory organics or humus and plant nutrients available for recycle.

scale operations. The objectives of this paper are to examine the effect of size of agricultural operation on the technical and economic feasibility of methane generation, to consider the technical requirements and information needs to construct and operate fermentors on smaller farming operations, and to present some of the initial data on the successful low cost system developed and operating at the Cornell University research dairy.

IMPLEMENTATION OF METHANE GENERATION AT SMALL SCALE - RATIONALE FOR CONSIDERATION

Background

Several reports have concluded that it is feasible and desirable to utilize energy generation technology that focuses on clean, renewable fuels as a mean component of the energy plan for the U.S. (Carter, 1978; Office of Technology Assessment, 1978). Obviously, with more than 45 billion dollars of foreign energy imported annually to the U.S., any significant renewable energy supply should be developed as rapidly as possible. The total quantity of biomass energy that could be generated from all animal and crop residues has been estimated to be between 1 and 10 percent of the total U.S. needs. As will be seen, it appears that methane generation from organic residues can be implemented on a broad scale in many agricultural operations and result in the production of a significant amount of substitute natural gas (biogas) at competitive costs with existing fuel sources. This can be accomplished using a technology that conserves valuable plant nutrients and soil humus and thus encourages improvement of crop production capability.

Number and Sizes of U.S. Farms

One ton of wet animal manure has a maximum energy value of about 5 dollars, with the other constituents, such as nutrients, refeedable protein and fiber, and water having a combined value of 5 to 10 dollars. Because of these relative low values per unit mass it is assumed that transport of animal wastes with large amounts of water to central processing sites is not practical in most cases. Thus, even for larger farms it may be necessary to consider methane generation at or near the food production cite. Also, because of the high cost of compressing and moving both biogas and biogas that has been treated to commercial natural gas standards, on-site utilization or conversion will oft be the most appropriate use.

Although the vision of U.S. agriculture has become synonymous with large agri-business, the size of the average American farm is still small. The arithmetic average size of the 2.3 million farms was 364 ha in 1974, but this includes many large and relatively unproductive operations in Alaska, Arizona, Colorado, Montana, Nevada, South Dakota and Wyoming (Committee for the World Atlas of Agriculture, 1969; U.S. Dept. of Commerce, 1974). For example, although the total numbers are small, the average farm in Alaska and Arizona exceeds 2000 ha. If these states are deleted from the calculation the average U.S. farm in 1974 occupied a land area of 129 ha, or only slightly larger than the 1960 value. Only a little more than 5% of the farms have more than 400 ha (1000 acres).

The distribution of animal manure residues as determined from census values for animal numbers and average waste production rates is distributed among the major groups as follows: milk cows - 36%, beef cattle - 28%, swine - 23%, and poultry - 13%. In essence this indicates that all major animal groups produce residues with significant methane generation potential. Due to the economics of scale, the potential for each animal production to provide residues for energy production depends on the size of the individual operations. The two size extremes are represented by swine and beef. With a modest fraction of exceptions, swine-raising operations are relatively small (with 40 mature pigs a common size) (U.S. Department of Commerce, 1972). The quantity of residues generated by these facilities may not be large enough to warrant installation of a generation system solely for energy production. However, other waste processing needs, such as odor control, may warrant consideration of the anaerobic fermentation process. Conversely, the huge quantities of residues generated in beef feedlots of more than 100,000 cattle would indicate that these would be areas favored for consideration of this technology. However, there are only a few facilities of this size and 83% of all beef production installations have less than 1,000 head (Loehr, 1977).

A simplistic approach to determine the viability of the technology on small scale operations would be to estimate the total quantity of residues that would be generated from a given size of operation. The potential gross value of the energy generated could then be related to the capital investment required to install the methane generation system. This overview is developed in the following sections.

A summary of methane generation potential from the four major animal groups is shown in Table 1. The high gas yields for manures from swine and poultry represent a more biodegradable manure produced from these animals. Using the gas production potential listed earlier (Table 1), and assuming a conservative value for the methane of \$7.93 per 10^6 kcal (\$2 per 10^6 Btu) generated, the annual gross income from any operation can be estimated. If this is assumed to represent the maximum available capital to pay for the operation, and it must be completely paid for within 10 years, the maximum allowable cost of a system can be estimated. For ten animal units the maximum acceptable total cost would be about \$2400, whereas \$24,000 represents the maximum cost for a 100 animal uni operation. Although this is a highly simplistic analysis, it does indicate the order of magnitude of expenditures acceptable with these sizes of operations. It should be emphasized that in cases where the value of odor control or other characteristics of these systems are of benefit, the acceptable investment will rise.

The relationship between the size of the agricultural operation, numbers of different sizes in the U.S. and the annual energy value of the processed organics is shown in Figure 1. This data also includes the number of human population equivalents that could be potentially supplied with all energy needs by plant and animal organic residue production rates. This equivalence was calculated by dividing the energy generated by the average annual total per capita energy usage rate of 300 million Btu per capita. This data emphasizes that there are several million farming operations that have the potential of generating between 0.5 and 3 tons per day of agricultural residue which would be available for fermentor feed material that would generate, annually, energy valued at between \$1000 and \$8000 per year. On the other hand, there are several thousand animal raising operations that generate greater than 10 tons of dry organic per day; and no crop or animal raising operations under one management that produce 1000 tons per day of agricultural residue.

Providing an economically feasible small farm system is clearly a difficul problem if the energy value must pay for the entire system. At an energy value of \$10 per million kcal, the 65-cow dairy may generate energy with a gross value of \$3000. Thus a total capital investment of less than about \$30,000 is required, depending on factors such as fraction of energy that is useful, loan interests, amortization period, and operation and maintenance costs. If it is concluded that these small systems are too small to be made economically attractive, this decision would eliminate about 90% of all

Table 1. Estimated manure production rates and biogas generation potential from animal wastes (Estimated output in kg per 453.6 kg (1000 lbs) live weight).*

	Dairy Cattle	Beef Cattle	Swine	Poultry
Manure Production kg/day	39	26	23	27
Total Solids kg/day	4.8	3.4	3.3	7.9
Volatile Solids kg/day	3.9	2.7	2.7	5.8
Digester Efficiency	35	50	55	65
COD/VS Ratio	1.05	1.12	1.19	1.28
Biogas Production** liters/kg VS added	219	325	381	490
Liters cf/453.6 kg animal/day	860	870	1,020	2,860

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*Actual values may vary from these values due to differences in feed ration and management practices.

**Based on theoretical gas production rate of 831 liters per kg (13.3 cf/lb VS) destroyed (where 0.45 kg (1 lb) COD stabilized equals 159 liters (5.63 cf) CH_4 (McCarty, 1964) and assuming the CH_4:CO_2 ratio is 60:40 and the conversion factor for VS to COD is 1.42).

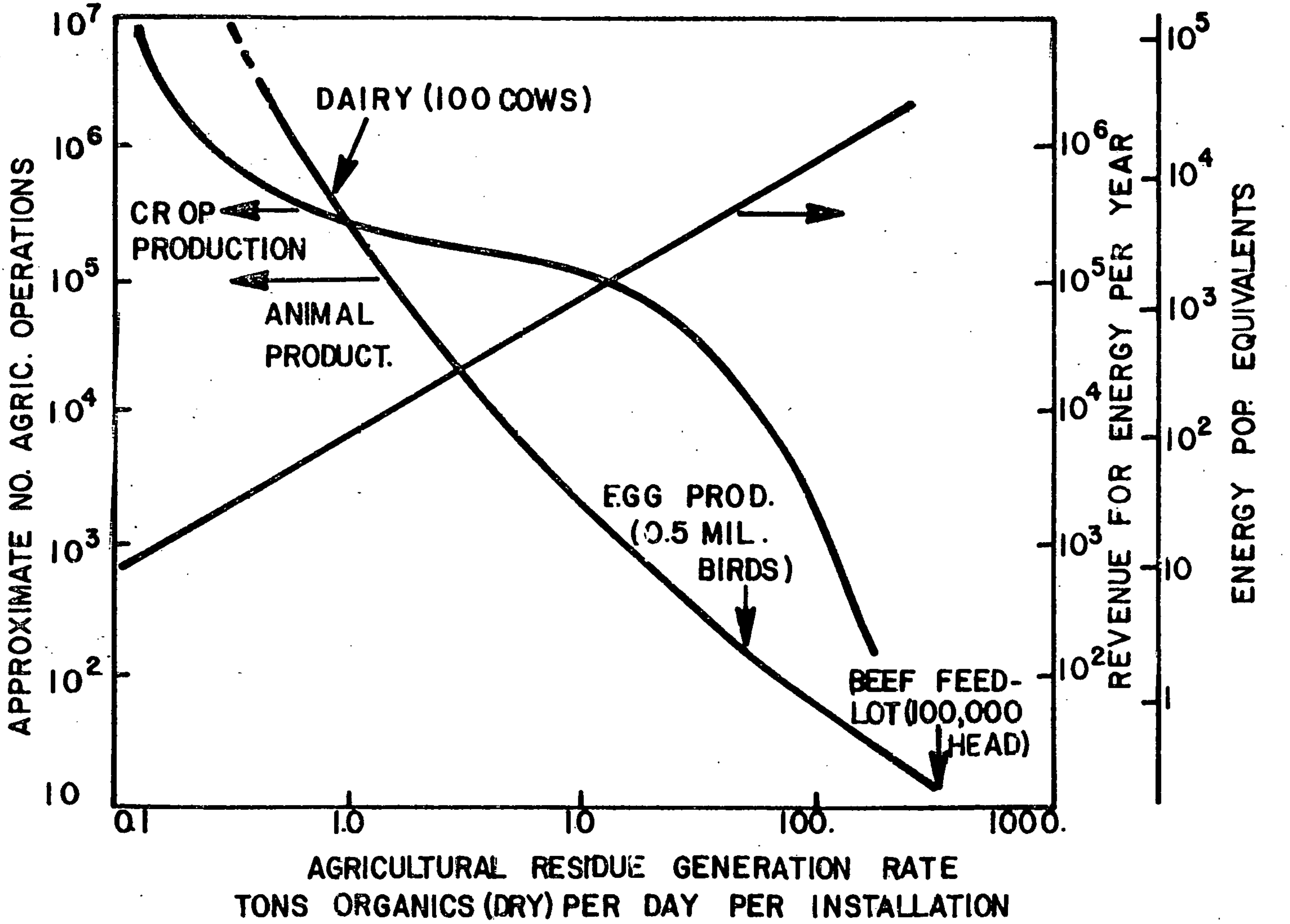

Figure 1. Relation of the number of farming operations, size of operation, and sustainable

agricultural residue generating operations from considering on-site renewable fuel generation.

Energy Balances on Small Farms

It is essential that the technology of methane generation be able to produce a significant quantity of energy and that it be in a form that can be easily used for existing purposes in order to displace existing critical fuels. To determine whether this could be possible, detailed energy balances were developed for dairies and medium-size beef feedlots in an earlier study (Jewell, et al., 1976). A summary of energy flows on a dairy is shown in Figure 2. The total energy use per cow on 40 and 100-cow dairies was found to be 4.1 x 10^6 and 3.1 x 10^6 kcal per year, respectively. Depending on the type of fermentor design utilized, the estimated total per cow net quantity of energy that was estimated to be potentially available from agricultural residues produced on these dairies (located in New York) was 3.4 x 10^6 kcal to 5.4 x 10^6 kcal for the 40-cow dairy and 2.5 to 4.7 x 10^6 kcal for the 100-cow dairy. These energy balances indicated that the quantity of energy utilized was often equal to or exceeded by the energy content of the biogas generated from the residues. These energy balances identified the interesting possibility of developing energy self-sufficient food production operations. They, however, do not address the difficult problems of utilization including storage cost and existing energy system conversion feasibility.

Estimated Costs of Methane Production Systems

A number of methane production systems were designed and costs for possible use on dairies and beef feedlots (Jewell, et al., 1976). This economic analysis included the valuation of capital investments, maintenance, and operational costs based on 1975 dollars. A summary of costs for three different designs is given in Table 2. Note that the estimates for the costs of the fermentation systems use existing components and only the net available energy (gross energy minus that used to maintain the process temperature). The added cost to use biogas is not shown in Table 2. The existing cost of natural gas is highly variable, depending on the situation but will cost \$10 per 10^6 kcal (\$2.50 per 10^6 Btu) or more. The cost of methane generation for a 40-cow dairy is about double present costs, but only 50% more than existing cost at the 100-head size, and only half the existing market value for the 1000-head beef feedlot. When only considering energy generation costs, it would appear that animal

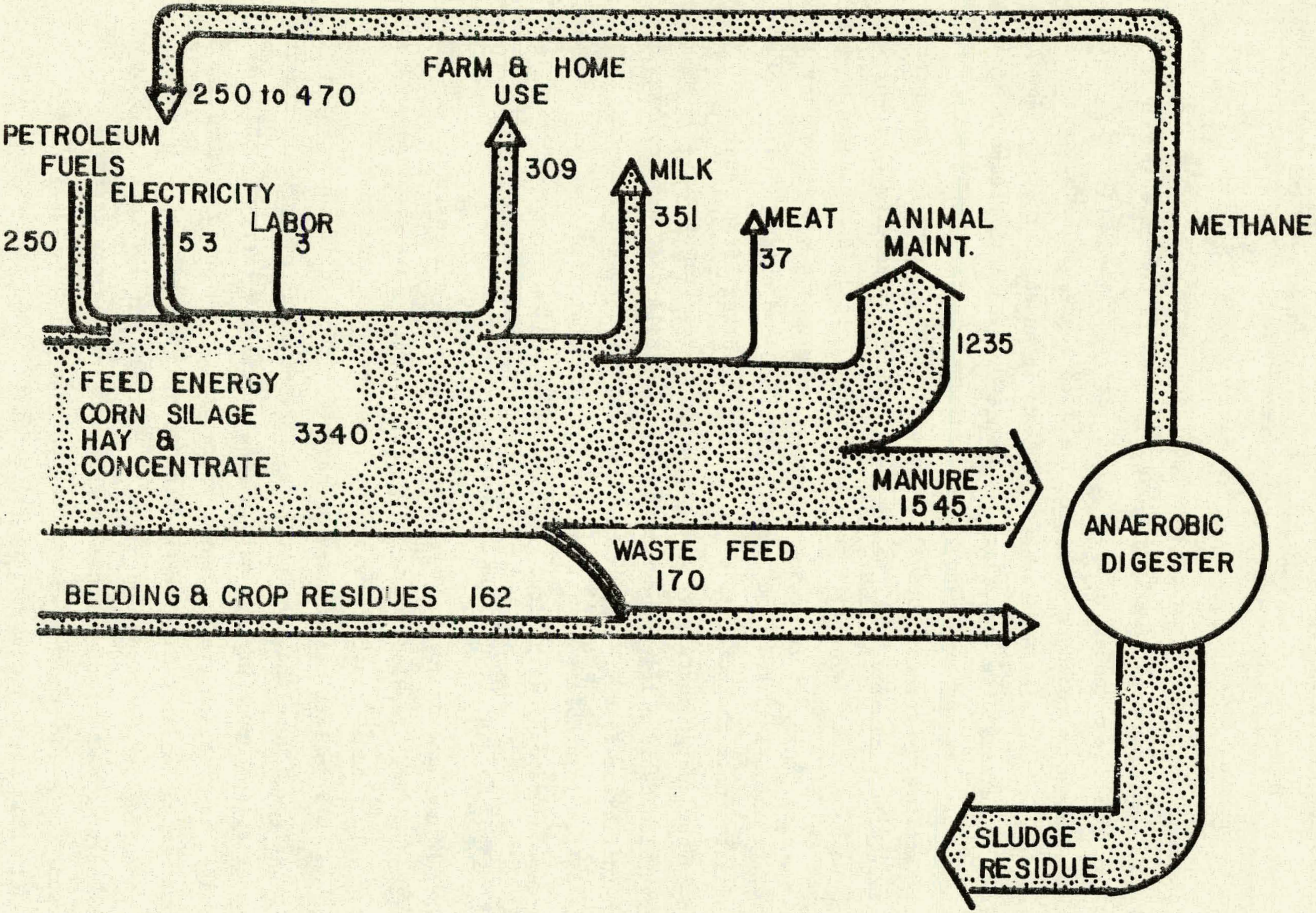

Figure 2. Energy flow on a 100 cow dairy farm (numbers are in millions of kilocalories per year).

Le 2. Estimate of the cost of generating the net energy (available energy) on dairies and beef feedlots (from Jewell, et al., 1976).

	40 Dairy			100 Dairy			1000 Beef		
	CON	B.L.	P.F.	CON	B.L.	P.F.	CON	B.L.	P.F.
Investment									
Total $	12,500	22,000	10,000	16,000	29,500	14,500	32,000	57,500	27,500
$ Per Cow	312.50	550.00	250.00	160.00	295.00	145.00	32.00	57.50	27.50
Annual Costs									
Total $	3,650	6,150	2,460	4,500	7,650	3,300	8,750	14,400	6,300
$ Per Cow	91.25	153.75	61.50	45.00	76.50	33.00	8.75	14.40	6.30
Gross Energy									
Production 10^6 kcal/yr	224	265	193	477	567	413	2,280	2,510	185
Net Energy									
Production (actual) 10^6 kcal/yr	136	157	108	296	358	246	1,720	1,890	1,395
Energy Production Costs									
$/$10^6$ kcal	26.80	39.20	22.80	15.20	21.40	13.40	5.10	7.60	4.50
($/$10^6$ Btu)	(6.80)	(9.85)	(5.75)	(3.85)	(5.40)	(3.40)	(1.30)	(1.95)	(1.15)

raising operations with more than 100 cows can provide energy at competitive costs.

If the costs of gas processing, biogas conversion and utilization, and the variability of demand are added to the costs of generation, the total costs of developing the useful energy can be determined. If all the biogas could be utilized, nearly all situations would generate useful energy at costs competitive with some alternatives. However, because only a fraction is usable, the actual cost of generation is higher, thus making only the larger scale operations appear to be attractive.

It should be emphasized here again the above approach relates the value of anaerobic fermentation of agricultural residues to the net usable energy produced. However, these systems provide benefits other than energy, such as labor reduction, odor control, runoff reduction, aesthetic value, general ease of operation, residue recovery, and nitrogen conservation. Even in the smaller operations these combined values may exceed the cost of installing the system, thus making the energy available at a low cost or a credit to the pollution control system.

The overview analysis emphasizes several important aspects of the feasibilit of adopting this technology. The methane generation reactor costs represent about 80% of the total gas usage for the 40-cow dairy, but drop to 30% for the 1000-head beef feedlot. The basic design concepts used in the previous discussion utilize current practices employed both for sewage sludge digestion and liquid manure storage. Aspects that have potential for cost reduction are in the development of low-cost methane generators composed of available materials and improved use of energy.

TECHNICAL BASIS FOR LOW COST METHANE GENERATOR DESIGNS

Data presented earlier indicated that the largest majority of agricultural operations are smaller than the 100-cow dairy. The estimated total investment in the methane reactors varied between $15,000 and $30,000 depending on the type of design chosen. Since this level of investment results in marginally acceptable energy costs, and the reactors represent the largest fraction of t total costs, one obvious alternative is to seek lower cost reactor designs.

The effect of high reactor cost on the economic feasibility of methane generation can be illustrated by comparing the annual value of energy produced per unit volume to the reactor capital investment expressed on a unit volume basis. Based on existing design criteria for methane reactors, between 1 and 3 cubic meters per cow would be the required reactor volume and this would be loaded at 8.3 kg VS/m^3-day (0.5 lb VS/ft^3/day). The gas production rate would be about 0.8 to 2 volumes per volume of reactor per day, or around 400 volumes per year. Minimum construction costs for tanks would be about $100 per m^3. Since the biogas has a maximum value of $12 per m^3 per year, it is clear why these systems are not economically attractive. In essence, this indicates that conventional rigid tank construction is difficult to justify for small scale installations because of the high capital cost and the large volume requirements. Instead lower cost alternative materials need to be considered. Also, the strict control systems known to be required for sewage sludge digestion need to be examined for a uniform agricultural residue reactor feed.

The Cornell University effort to develop low-cost alternatives has emphasized low-cost construction materials and techniques that would be compatible with existing farm skills of self construction and maintenance. Two full-scale 65 cow dairy systems were placed into operation in the late Spring of 1978 after laboratory and pilot plant operations had identified specific design needs. Due to the high cost of rigid tanks, the emphasis is on soil supported structures that are lined and covered with flexible rubber-type materials. It would appear that materials are available that would be useful and significantly reduce the cost of such systems. For example, several liner type materials are available costing less than $3.50 per square meter ($0.50 per square foot). When using a reactor with a 5 meter depth, the cost of the reactor vessel itself will be less than $2 per cubic meter of reactor volume ($0.10 per cubic foot) as compared to $100 per m^3 for rigid tank construction. Additional cost will be excavation, insulation and reactor temperature controls and pumping. Insulation and soil preparation may add another $3 per cubic meter ($0.14 per cubic foot). Temperature controls and pumping will be common to all systems. If these construction materials have the required characteristics for durability and handling it would appear that reactor costs could be less than $5 per cubic meter ($0.25 per cubic foot).

This low construction cost would also enable larger reactors to be built without significantly changing the economic feasibility of the system. For example, a change from a 30-day hydraulic retention time to a 60-day unit would increase the capital cost by $1500 for a 100-cow dairy. Thus these economics introduce interesting possibilities in intermittent feeding, lower temperature of operation and other factors that may be feasible in large low rate reactors.

Reactor Design Considerations

The main focus of the simplified reactor development has been on the use of low cost materials with simplified operation and control of the reactor. Additional details can be obtained from the final report by Jewell, et al., 1979 In essence, the approach has been to use the simplest design first and then to modify each paramater when and if necessary. Laboratory and pilot plant testing have been completed with unmixed reactors and temperatures as low as 22°C. The main design concept which has been tested is the "unmixed horizontal displacement reactor" or the plug flow design.

In general, most of the conditions tested should not lead to successful reactor operation according to the bulk of information in the literature. However, the data developed for highly reactive and variable substrates, such as sewage sludge, should not be expected to simulate the more viscous and homogeneous animal and crop residues. The fact that the information developed so far with simpler reactors has been positive indicates that some of the parameters found to be critical with sewage sludge are not as sensitive when treating agricultural residues. It is likely that the inconsistencies are more a reflection of varying reactor feed material characteristics than changes in the fundamental controlling parameters of anaerobic fermentation.

Description of Experimental Approach

Laboratory studies of the plug flow reactor feasibility applied to dairy manure were initiated in 1974 and expanded in 1976 (Jewell, et al., 1976, Jewell, et al., 1978). These were bench scale studies with about 20-liter reactor volumes. Attempts to simulate an actual elongated plug flow reactor using a 6-inch diameter tube reactor was not successful. However, completely mixed reactors connected in series are considered to be mathematical approximations of the plug flow concept. Three sets of four reactors in series (3-liter liquid volume each) were operated, one each at 22°C, 35°C, and 60°C.

In the fall of 1976 a 5000-liter plug flow pilot unit was placed into operation to process the manure produced from 3 or more cows (see Figure 3). Most recently a cube-shaped 5000-liter reactor was placed into operation to provide a comparison of reactor shape and mixing on the fermentation process. Based on positive data from these experiments, two full scale reactors were placed into operation for 65-cow dairies in May 1978. One unit is a conventional completely mixed concrete tank reactor, simulating the conventional sewage sludge digester design (see Figure 4), and the other is a plug flow soil supported flexible cover reactor (see Figure 5). These two units have been operating in parallel. Valuable data on cold climate operation of the simplified, plug flow reactor has been obtained. This is the first large scale parallel operation comparison of the major design variable of mixing and plug flow designed for anaerobic fermentors.

Plug Flow Fermentation of Dairy Manure Results

A summary of feed substrate characteristics of the dairy cow manure is given in Table 3. It should be noted that the biodegradable fraction as measured in long term batch studies varied from a low of 40 percent of the TVS to a high of 65 percent of a TVS over a 3-year period of testing. This should be kept in mind in comparing data and kinetic analysis. The primary reason for this variation would appear to be animal diet variations.

A summary of data from the bench scale laboratory series reactors is given in Table 4. The HRT expressed in this table is for the total in all four units (i.e., a 10-day HRT refers to a 2.5-day retention period in each of the series reactors). The laboratory plug flow simulation data indicate that this design results in stable operation at rather extreme conditions, including lower temperatures and at a 1-day HRT (with the aid of effluent recycle).

From initial start-up, the 5000-liter plug flow pilot unit operated well with few problems as indicated by the start-up in Figure 6. Steady state gas production at 35°C with 8% total solids feed about 5150 liters per day or slightly more than one volume per volume of reactor per day. A summary of conditions tested with the pilot plant and results are given in Figure 7. Since the maximum biodegradable fraction is about 45%, most conditions tested have achieved conversion of 60% or more of the available or biodegradable organic matter. This data also indicated good correlation between the simulated oratory plug flow data and the pilot scale analysis.

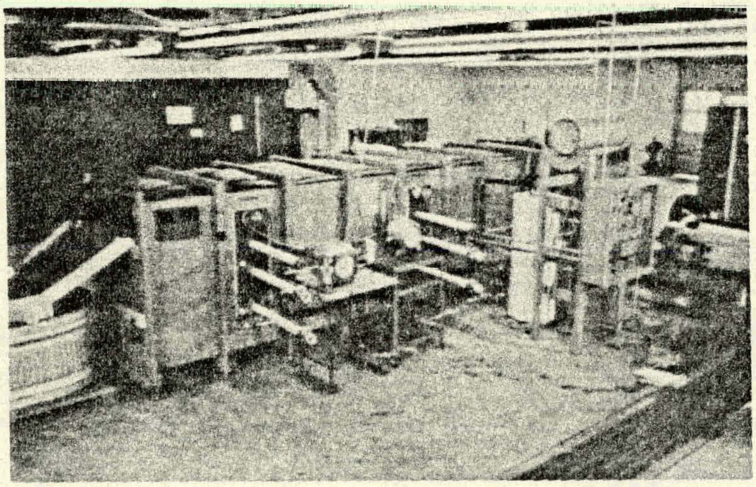

Figure 3. Photograph of the pilot scale plug flow fermentor located at the waste management laboratory of the Cornell University Teaching and Research Center Harford, New York.

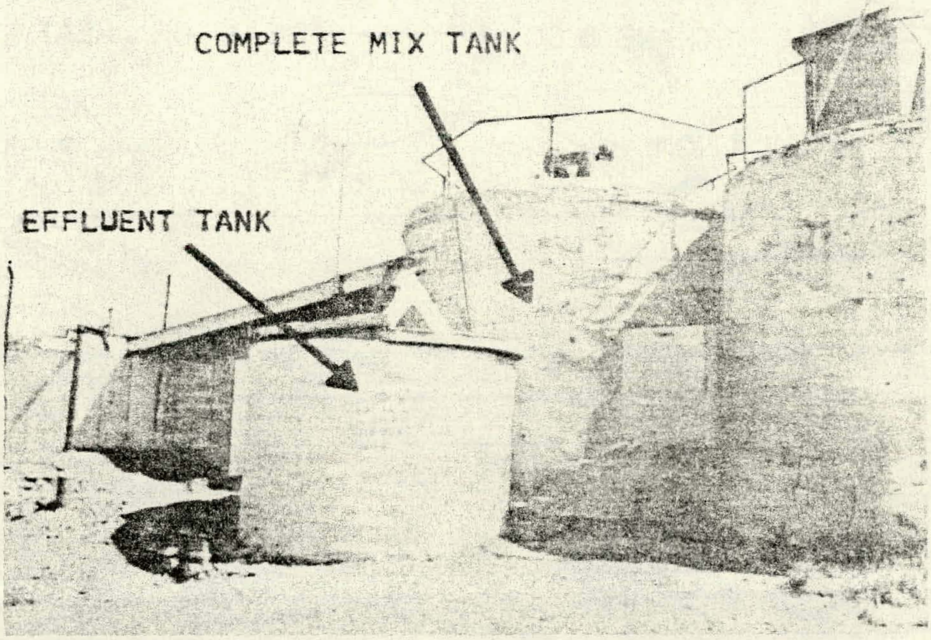

Figure 4. Photograph of 65-cow dairy full scale completely mixed anaerobic fermentor tested in parallel operation at Cornell University with a plug flow design of equal volume.

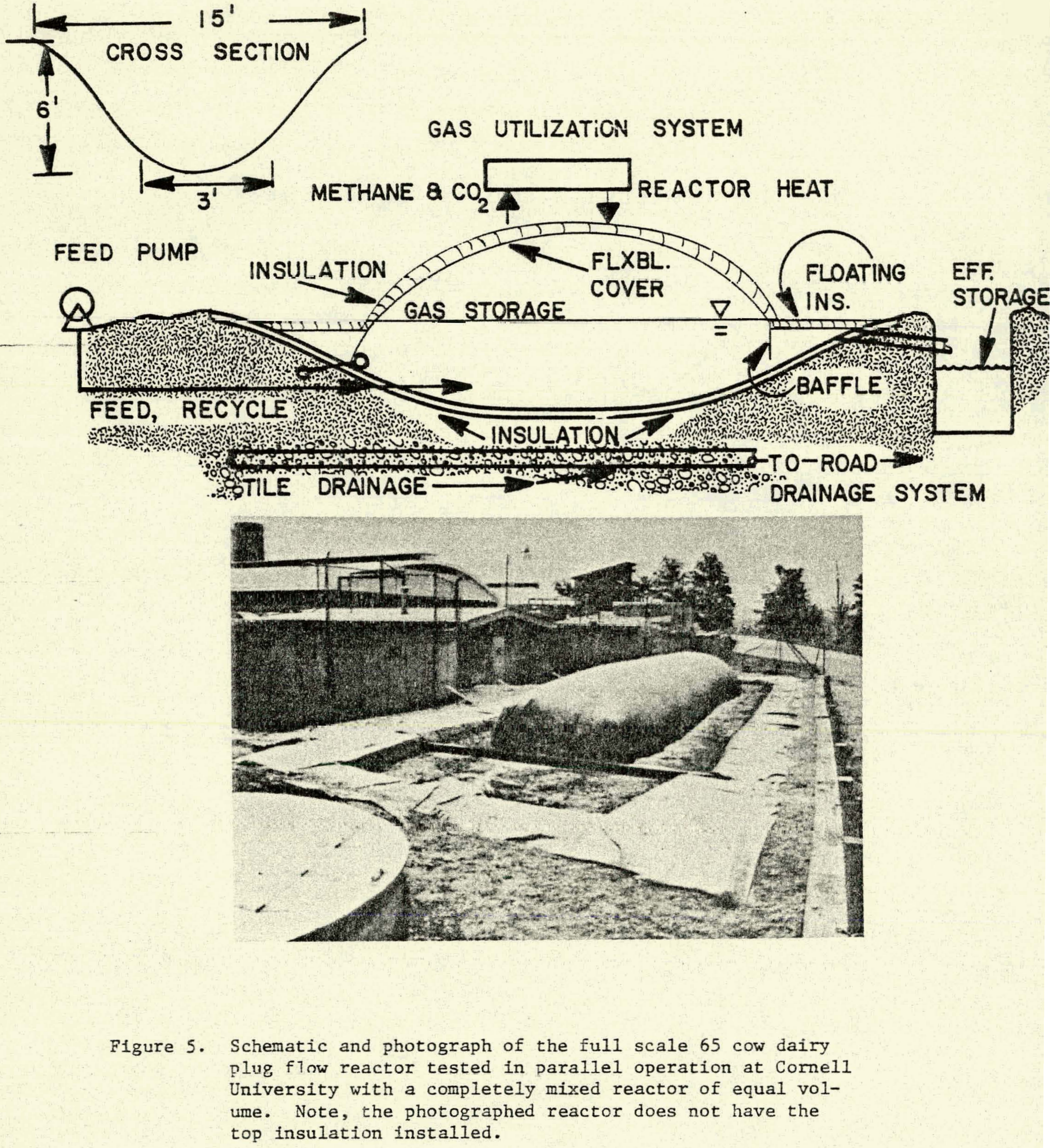

Figure 5. Schematic and photograph of the full scale 65 cow dairy plug flow reactor tested in parallel operation at Cornell University with a completely mixed reactor of equal volume. Note, the photographed reactor does not have the top insulation installed.

ble 3. A summary of analyses conducted on manure feedstocks and effluents from plug flow pilot reactor.

arameter	8% TS Feed Condition				10-12% TS Feed Condition			
	Influent		Effluent		Influent		Effluent	
	Range	Avg.	Range	Avg.	Range	Avg.	Range	Avg.
H	7.1-8.8	7.8	7.0-8.1	7.7	6.5-7.5	6.7	7.3-7.9	7.7
lkalinity, /ℓ $CaCO_3$	9.3-14.1		11.2-15.6		9.7-14.7		15.0-17.4	
KN, mg/ℓ	2958-4768		3240-5412		3600-5652		3926-5316	
H_4-N, g/ℓ	1451-2362		1947-2498		1698-2820		2460-2961	
ree Ammonia, g/ℓ	34-843		38-333		7-86		68-243	
otal Solids, /ℓ	63-91	82	47-77	59	103-120	112	75-96	88
Destroyed				28.0				21.4
otal Volatile olids, g/ℓ	54-79	71	36-65	48	90-104	97	62-80	73
Destroyed				32.4				24.7

Table 4. Summary of example performance data from reactor systems of four CSTRs in series.

	Temp 22°C		Temp 35°C		
	HRT = 12 days	HRT = 30 days	HRT = 10 days	HRT = 12 days	HRT = da
			--------(recycle)---------		
Influent Total Solids (gm/ℓ)	80.00	80.00	80.00	80.00	80.00
Influent Total Volatile Solids (gm/ℓ)	71.69	71.18	70.36	71.84	71.02
Effluent Total Solids gm/ℓ	69.87	59.09	59.60	59.11	59.34
% reduced	12.7	26.1	25.5	26.1	25.8
Effluent Total Volatile Solids (gm/ℓ)	61.86	51.01	50.73	50.92	50.51
% reduced	13.7	28.3	27.9	29.1	28.9
Total Gas Production ℓ/ℓ reactor-day	0.268	0.368	1.298	1.318	1.36
Methane Production Rates ℓ/ℓ reactor-day	---	---	0.543	0.681	0.71
ℓ gas/gm VS_A	0.044	0.155	0.185	0.184	0.21

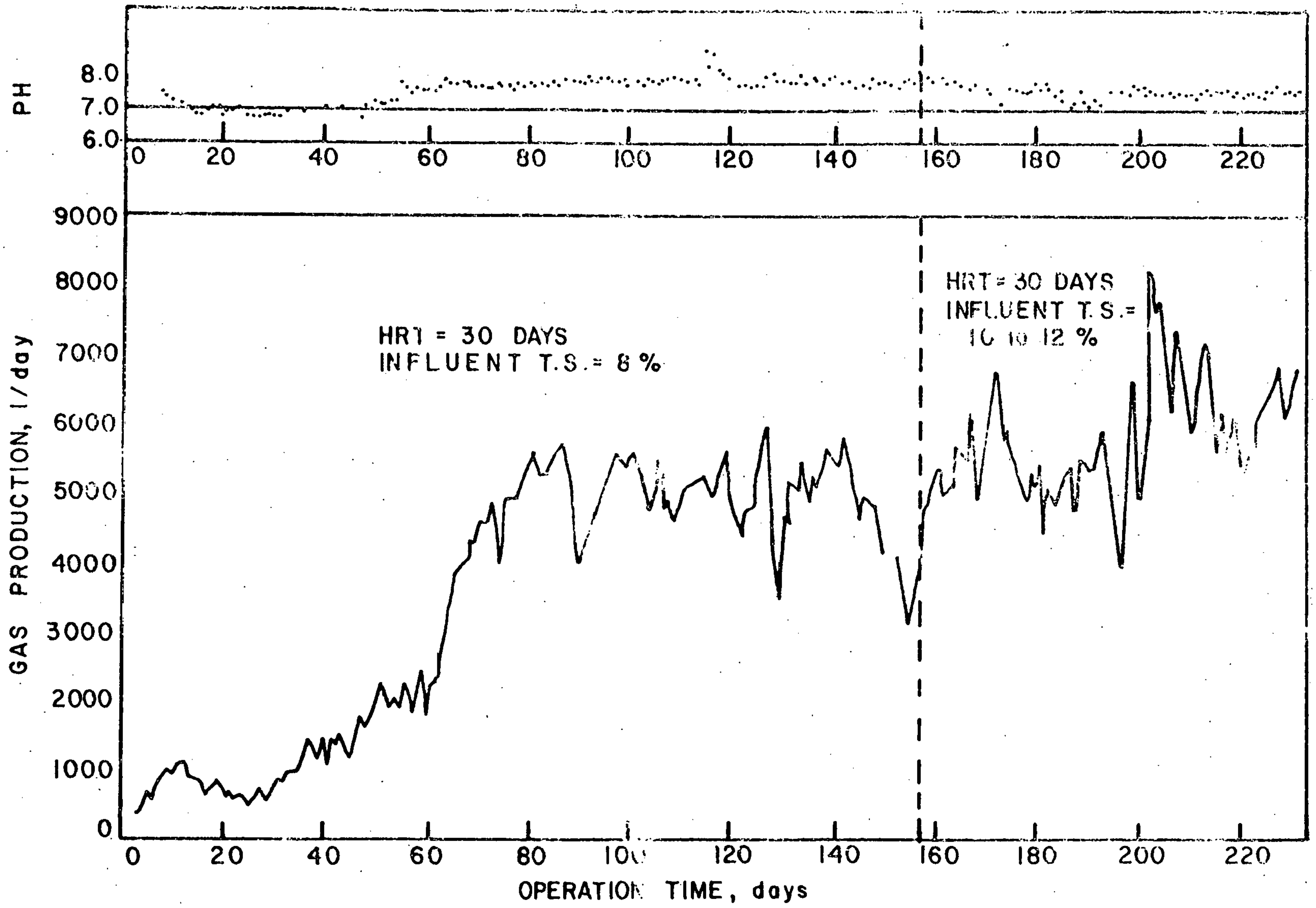

Figure 6. Chronological development of gas production and effluent pH during initial start up of the pilot-scale plug flow reactor.

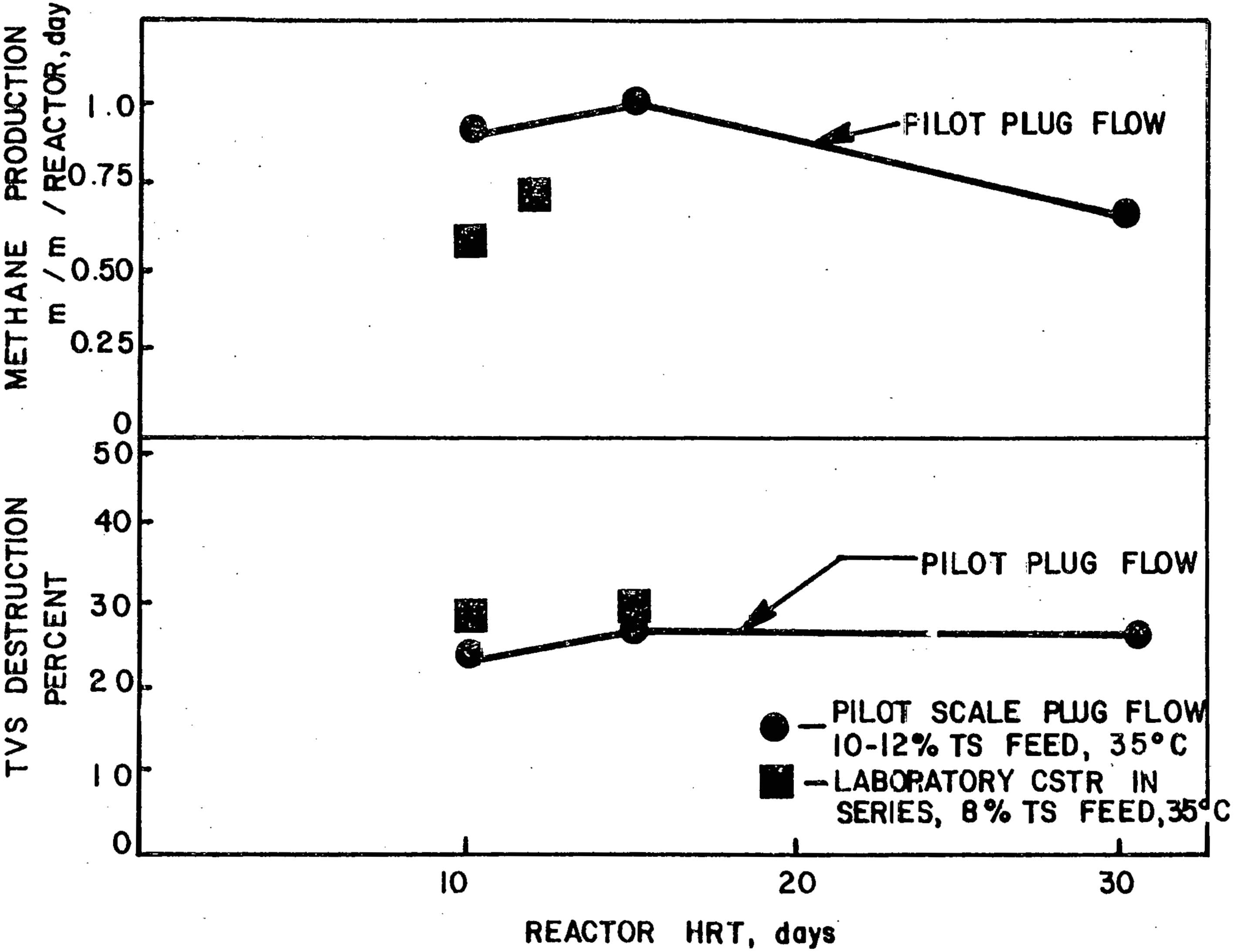

Figure 7. Comparison of laboratory and pilot scale plug flow anaerobic fermentation of dairy cow manure

The majority of more than a year's operational data for the two full scale units is summarized in Table 5 and illustrated in Figure 8. Under all conditions the plug flow reactor resulted in significantly more solids conversion to biogas. Although a specific design will most likely be constrained by site characteristics, the 15-day HRT plug flow reactor that produced greater than 2.3 volumes of biogas per volume of reactor would appear to be a good compromise between reactor size and solids conversion efficiency.

Effect of Cold Climate

The temperature of northern New York often drops below minus 30°C, and can average minus 20°C for several weeks. Since maximum space heating and other energy needs may coincide with this period it is essential that careful net energy analysis of the system be known. Uninsulated or poorly insulated systems will not result in positive energy balance in cold climates. The full scale units were characterized for energy losses under different conditions with the results as indicated in Table 6. Note that even with a well designed plug flow reactor operated without an insulated gas cover, resulted in a negative energy balance; i.e., it required 17 percent more energy to operate at 35°C than was produced. However, a well insulated top results in a net energy production of 44 percent of the total even when the air temperature averaged -18°C. The projected year-round variation of biogas production as affected by a cold climate is shown in Figure 9. Although it appears that operation of a unit at a lower fermentor temperature (such as 25°C) would result in a high net energy production, the increased reactor size and related costs may not justify this operating condition.

Table 5. Summary of full scale operation of dairy manure anaerobic fermentors when fed a TS feed of 129 gm/ℓ.

		15-day HRT	30-day HRT	
Gas Production	vol/vol	2.13	1.13	FULL SCALE COMPLETELY MIXED REACTOR
	ℓ/gm VS_A	0.281	0.310	
	ft^3/lb VS_A	4.5	5.0	
Gas Composition	% CH_4	55	58	
Solids Destruction	% TVS red	27.8	31.7	
	% TBVS red	61.9	70.4	
Gas Production	vol/vol	2.33	1.26	FULL SCALE PLUG FLOW REACTOR
	ℓ/gm VS_A	0.337	0.364	
	ft^3/lb VS_A	5.4	5.8	
Gas Composition	% CH_4	55	57	
Solids Destruction	% TVS red	34.1	40.6	
	% TBVS red	75.8	90.1	

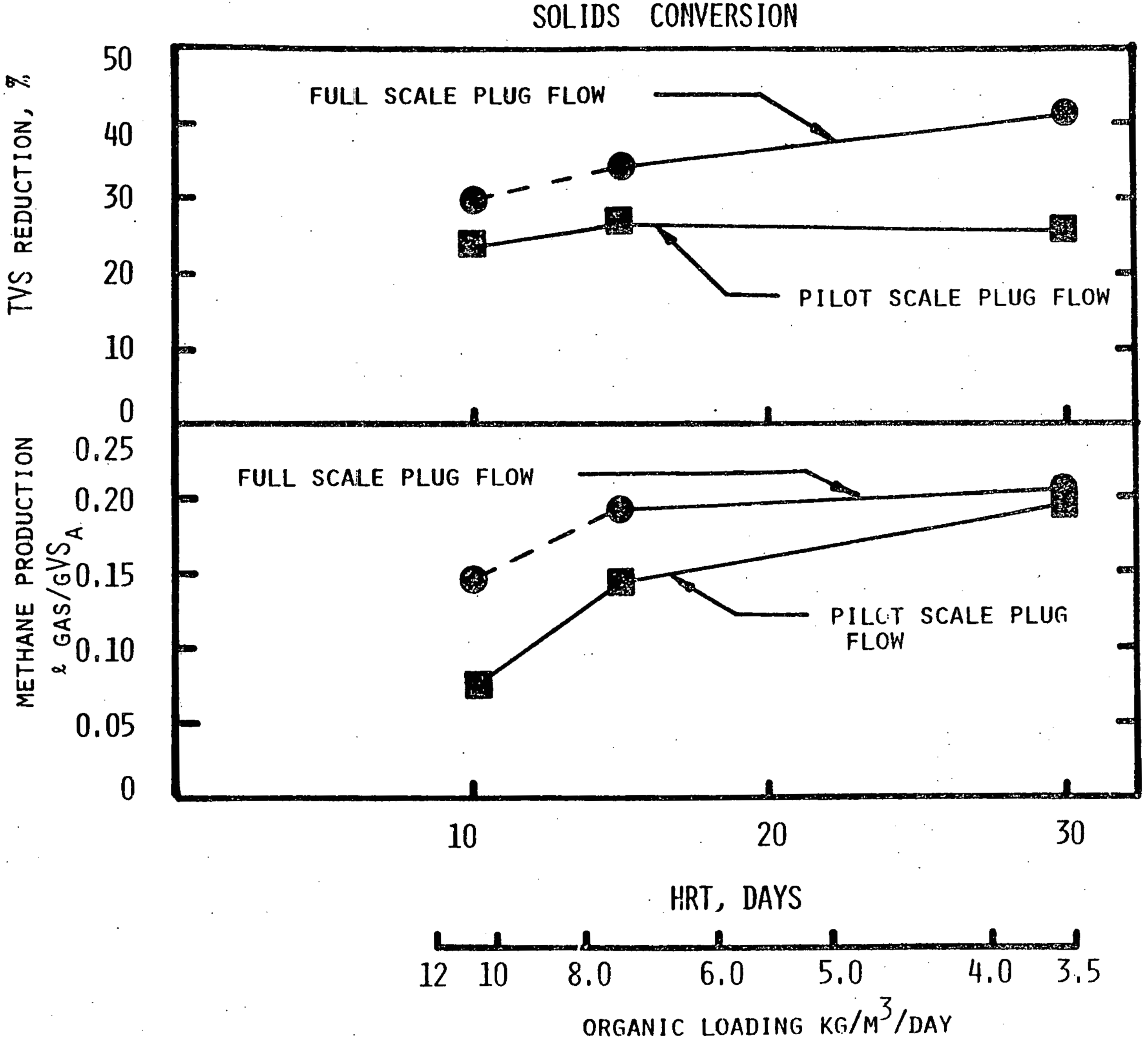

Figure 8. Comparison of solid conversion efficiency in full scale completely mixed and plug flow reactors fed about 13 percent solids dairy manure

Table 6. Comparison of energy generation with the two full scale systems operating under cold conditions.

Parameter	Plug Flow Reactor			Completely Mixed Reactor		
	30 Day HRT†	15 Day HRT*	15 Day HRT*	30 Day HRT	15 Day HRT	15 Day HRT*
Ambient Air Temperature, (°C)	1	1	-18	1	1	-18
Heat Loss to Surroundings, (Btu/day)	496,000	111,000	170,000	78,250	78,250	123,900
Feed Temperature, (°F)	50	50	40	50	50	40
Heat Loss to Heat Feed, (Btu/day)	143,000	246,000	303,000	130,340	242,210	302,760
Total Heat Required, (Btu/day)	639,000	357,000	473,000	208,490	320,460	426,860
Total Heat Supplied, (Btu/day)	1,132,000	638,000	845,000	496,000	736,000*	1,015,860
Heat System Efficiency (%)	56	56	56	42	42*	42
Total Biogas Produced (ft^3)	1717	2500	2500	1412	2500*	2500
Total Energy Produced, (Btu/day)	971,000	1,500,000	1,500,000	786,000	1,500,000*	1,500,000
Energy Used, (Btu/day)	1,132,000	638,000	845,000	496,000	736,000*	1,015,860
Net Energy, (Btu/day)	0	862,000	655,000	290,000	764,000*	484,140
Energy Used, (% of Total)	117	43	56	63	49*	68
Net Energy, (% of Total)	0	57	44	37	51*	32

*Calculated values from measured thermal characteristics

†The negative energy balance that occurred with the 30 day HRT plug flow reactor experiment resulted from operation in winter weather with no insulation applied to the gas collection cover. The remainder of the plug flow experiments were carried out after 4 inches of fiberglass insulation and a hypalon sheet were placed on the reactor top.

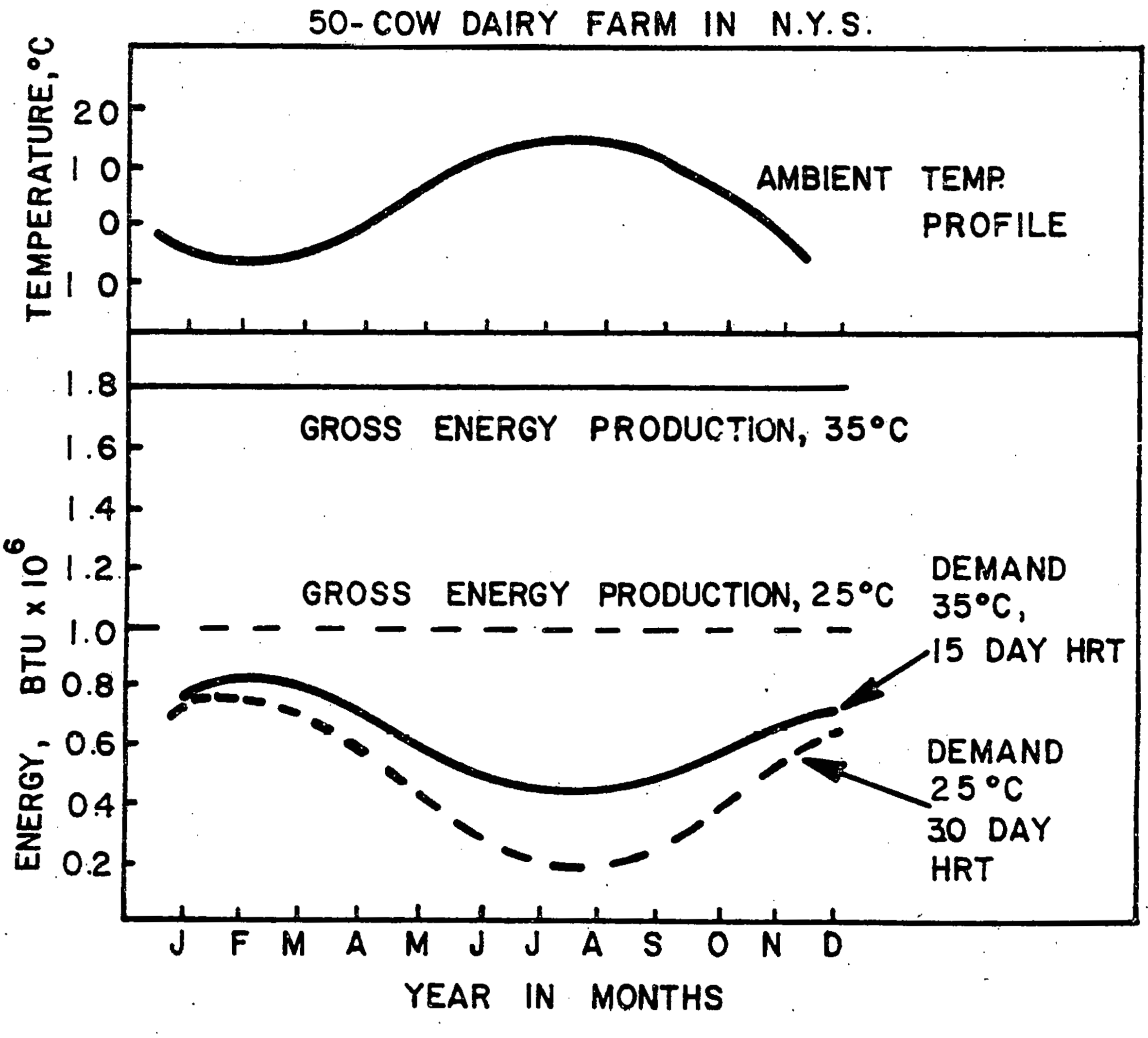

Figure 9. Annual variation of net energy production from a plug flow dairy anaerobic fermentor when operated at two temperatures.

DISCUSSION

The size distribution of U.S. farms and the quantities of organic residues indicate that as much as 75% of the total is generated by small scale agriculture. Most of these farms could supply a raw agricultural residue feed of less than 2700 kg per day (3 tons per day) on a year round basis. Although there are reasons other than energy production why anaerobic fermentation may be installed, it would be desirable for these systems to be capable of generating energy that would be cost competitive with existing supplies. This would enable direct replacement of common fuels such as gasoline, diesel oil, and natural gas, and indirect replacement of electricity. However, conventional designs do not appear to be economically feasible since the value of the resulting energy varies from 2 to 10 times as expensive as existing supplies.

A number of factors can be considered to change the economic feasibility of methane generation technology for small scale farms. First, more efficient use of the energy is required. This means that a better match of the supply with the varying energy demands is necessary. This may be difficult to achieve without significant capital investment in conversion and storage of the energy. Little work appears to be on-going in this area.

The second area where significant savings may be obtained is in the basic design and operation of the reactors. The simplest and lowest cost reactor would involve the use of low cost flexible materials and soil supported structures. A number of more practical full scale installations have built these types of systems, but little operating data is available (Fry, 1974; Meynell, 1976). The definition of operating requirements for mixing and temperature have been the major focus of Cornell University's recent activities For the first time it has been clearly demonstrated that the simplified, low-cost plug flow designs that have been tested on a laboratory, pilot plant scale and full scale are more efficient than the conventional completely mixed reactor design commonly used for sewage sludge treatment. The plug flow design when applied to dairy cow manure with feed solids up to 13% total solids has achieved 10 to 30 percent greater solids conversion than a completely mixed reactor of equal size operated under identical conditions in parallel tests. Although further information is necessary, it would appear that operation at lower temperature with dairy cow manure may be feasible. On-going demonstration of the simplified reactor design with low cost materials and construction techniques will continue to clarify specific design requirements.

Although several large scale plug flow reactors have been used for significant time periods, there remains a lack of data to interpret the basic mechanisms that enable active fermentation to proceed under these adverse reactor conditions. The effluent volatile acids from the plug flow pilot plant were usually less than 300 mg/ℓ as acetic acid and most of the activity occurred in the first third of the reactor. Apparently enough mixing was induced by diffusion, temperature gradients, and bubble movement to provide the minimum mixing requirements to support the complex microbial activities required to form methane. Additional studies on the fundamentals need to be completed before the influence of temperature and mixing on agricultural residues fermentation is clarified.

Finally, it is necessary to provide some overall perspective on the competitiveness of the methane generation technology in relation to competing energy sources. Since there are nearly 10,000 dairies with 100 head of cattle in New York State, it is interesting to consider what the total energy generation potential would be if it was possible to install this number of systems, and also to compare the total investments to those being considered for other technologies. The total potential biogas energy generation capacity would be about 483 million watts per day (energy equivalent to the methane). If the average investment per 100-cow dairy equalled $30,000 each, the cost of implementing this technology would be $620 per kilowatt of generation capacity. If the improvement presented in this paper prove to be successful, the resulting generation costs would decrease to as low as $200 per kilowatt. Comparison of this source of energy to costs for nuclear and fossil fuel electricity generation given in Table 6 shows that the cost of biogas is presently competitive, and could be one of the least expensive options available if on-going efforts are successful in developing lower cost systems. Thus it would appear that the smaller farms should play a major role in providing clean, renewable cost competitive energy in the near future.

SUMMARY AND CONCLUSIONS

The possibility of installing methane generation technology on small farming operations raises complex questions relating to the availability of organic residues, limited financial capability, available skills, and appropriate technology. The objectives of this paper were to examine the effect of size of agricultural operation on the technical and economic feasibility of methane generation and to consider the technical requirements and information needs to construct and operate fermentors on smaller farming operations.

Table 6. Comparison of the capital cost of energy generation using various technologies (in 1978 dollars).

Type of Generation	Unit Nominal Size, Megawatts per day	Cost $ per kW of Generation Capacity	Reference
Nuclear	1260	850	New York Power Pool 1978
Coal	850	740	"
Gas Turbines	---	230	"
Pumped Storage Hydro	1000	312	"
Methane Generation on New York dairies (10,000 units)	0.05	200 to 620	This Study
Mass Produced Solar Power Systems	---	1600 to 6000	Barber 1978

Number and Sizes of U.S. Farms

The arithmetic average size of the 2.3 million farms in the U.S. is 364 ha, and 80% of all farms are less than 204 ha (500 acres) in size. Only a little more than 5% of the farms have more than 400 ha (1000 acres).

The present total amount of manures is distributed among the major animal groups as follows: milk cows - 36%, beef cattle - 28%, swine - 23%, and poultry - 13%. The sizes of most existing agricultural operations and the rate of generation of organic residues that would be available as an energy feedstock indicates that the large majority produce less than 3 tons (dry) per day. Be relating the biodegradability of the organic matter to the potential energy generated, the approximate acceptable capital investment was estimated. The investment that could be offset by the methane generated by 100 animal units varies between $10,000 and $35,000 depending on the biodegradibility of the animal manures.

Energy Balances on Small Farms

Due to the high moisture content of most organic residues and the low generation rate, it is necessary to consider both on-site generation and utilization of the energy. Detailed energy flow studies indicated that the total amount of diesel, gasoline and electricity fuels use in milk and beef production was less than the energy theoretically contained in the biogas from anaerobic fermentation of the agricultural residues. These energy balances support the possibility of developing energy self-sufficient food production units. However, a large amount of development work remains to produce economically feasible utilization systems.

Estimated Costs of Anaerobic Fermentation Systems

Several different cattle manure anaerobic fermentation systems were designed and costed in order to estimate the total cost of the energy developed in these systems. Compared to the existing cost of natural gas of about $10 per 10^6 kcal, the losest cost of generation was about double this value for 40-cow dairies, 1.5 times the market price at the 100-cow dairy size, and about half this level for 1000 head beef feedlots. Addition of the costs for gas handling and conversion and adjusting the cost due to the wasting of energy at times when the demand and supply did not match resulted in making the gas more than twice as costly as the above values.

Thus these estimates indicate that significantly decreased costs of biogas generation would have the potential of producing significant quantities of usabl energy at cost competitive with existing fuels.

Technical Basis for Low Cost Methane Generation Systems

The definition of low cost simplified anaerobic fermentor that would be self-erectable by the farmer has been a main goal of the Cornell University program since 1974. In order to develop the basis for the technology the minimum needs for mixing, temperature control, reactor size, and new construction materials have been examined. These constraints have led to laboratory, pilot plant and presently full scale testing of unmixed, soil supported reactors constructed of low cost rubber-like pond liner material.

Because of the uniformity of the dairy manure feed its high viscosity, operation of laboratory and pilot scale plug flow reactors have been successful. Full scale demonstration at a 65-cow size has shown that this design is superior in operation to a completely mixed design for dairy manure substrate fed at 13 percent TS in almost every way. Based on the estimated cost of this simplified reactor design it is estimated that the total cost could decrease from about $30,000 using rigid tank construction similar to that used with sewage sludge to as little as $10,000. This reduction decreases the cost of energy sufficiently such that small dairies may be able to generate biogas energy at a cost competitive with other energy sources.

Finally, comparison of the total investment required for widespread adoption of methane generation technology in agriculture to nuclear, fossil fuel, and massive installation of direct solar power indicates that it would be highly competitive. Energy equivalent to nearly 500 megawatts per day could be replaced by biogas energy produced on New York dairies alone, at a construction cost varying between $200 and $620 per kW of generation capacity. This compared to values from $1000 to $6000 per kW generating capacity for all alternative energy sources.

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SUMMARY REPORT
ON THE
BIOLOGICAL PRODUCTION OF LIQUID FUELS FROM BIOMASS

by

Arthur E. Humphrey, University of Pennsylvania
Edward J. Nolan, General Electric Company

The purpose of this Summary is to describe past, present, and proposed technologies for the biological production of liquid fuels and chemical feedstocks from biomass raw materials and to assess their potential for contributing to the national energy supply. The biomass raw materials of interest can be grouped into four categories, i.e., 1) cereal grain starches, 2) sugar from cane and other similar agricultural crops, 3) cellulose from agricultural residues and municipal wastes and 4) wood. The fuel under consideration is alcohol derived by fermentation processes (synonymous with ethyl alcohol and ethanol).* Selected chemical feedstocks examined include butanol-acetone and acetic acid. However, none of these materials have the potential that alcohol does as a significant chemical feedstock alternative without major breakthroughs in higher fermentation concentrations and productivities of the product and significantly better separation techniques than distillation of dilute feed streams. For this reason, the major focus of this Summary is on the production of alcohol from biomass.

The Summary comprises three major considerations:

1. Economic analysis of existing process technologies and estimation of alcohol production costs of these processes translated into mid-1978 costs.

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*This should not be confused with methanol derived by pyrolysis of coal or wood.

2. Comparison of processes based on present and envisioned technological developments.
3. Technological challenges and breakthroughs needed to make the production of alcohol from biomass economically and energetically attractive.

The essential findings of these examinations are summarized below.

Economic Analysis

Utilizing the design of a 25,000,000 gallon per year alcohol production plant as the basis for making cost comparisons, the production costs for 95% alcohol (190° proof) as the base case product were estimated. If anhydrous alcohol is the desired end product, approximately 3¢/gallon should be added to the costs reported here. In the production cost analysis, no byproduct credits were given, mainly because these are truly unknown if significant quantities of byproducts, i.e., CO_2, furfural, xylose, fusel oil, animal feed, and waste solids such as lignins for burning fuel credit, etc., are produced. We believe that, at best, a 20¢ to 25¢ per gallon credit for animal feed and a 5¢ to 6¢ per gallon credit for waste solid fuel is the most credit that can be reasonably assigned to the process. However, we have left it to the readers and users of the Summary to select from the data presented those byproduct credits that they wish to assign in their consideration. Further, we have not included any return on investment (ROI) in the production cost because these can be so variable dependent upon the means of plant financing. For example, the interest charges for the capital investment raised from the sale of municipal bonds (perhaps for a plant to consume garbage) would be substantially less than the charges for capital raised as equity and from standard bonds. With these constraints and assumptions, the following production costs and capital investments were obtained for the various processes:

Basis: 25,000,000 gal/yr of 95% Alcohol Product
(No byproduct credits)

Raw Material	Assumed Raw Material Costs delivered	Capital Costs $/1,000,000	Production Costs $/gal
Sugar from sugar cane juice	$13.50/ton of sugar in cane	37.6	1.15
Molasses from sugar cane	$50/ton of molasses	17.7	1.08
Starch from corn grain	$2.50/bu	29.4	1.40
Starch from grain sorghum	$1.96/bu	29.4	1.24
Cellulose from wood chips	$20/oven dry ton	88.8	1.37
Cellulose waste paper	$24/oven dry ton	64.8	1.96
Cellulose from corn stover	$30/oven dry ton	67.4	1.72
Syrup sweet sorghum	$7.80/ton (yield 27 tons/acre)	37.4	0.95

Process Comparison

A number of processes were considered for producing alcohol. These included:

A. Cellulosic Substrates:
  - Natick Process
  - Gulf Oil Process
  - SRI/Wilke Process
  - SRI/Tsao Process
  - Purdue Tsao Process
  - Battelle Report
  - Katzen Process
  - Madison Process
  - Penn/GE Process

B. Molasses Substrates:
  - Schaffer/Battelle Report
  - ICAITI EX-FERM Process
  - SRI/Tilby Process

C. Grain Substrates:
  - Scheller Process
  - Seagram Process
  - Schaffer/Battelle Report
  - Schaffer/Battelle Process

D. Miscellaneous:
  - Intertech Process
  - Gregor Process
  - Dynatech/Kolbe Process

The cellulosic substrate processes utilize inexpensive substrates, available in large quantities, but currently relatively expensive methods for converting the cellulose to alcohol. In general, the yields are poor (less than 50%). The alcohol costs are in the range of $1.00 - $2.00 per gallon. Of these processes, only the Madison and Gulf Oil processes have been run at more than bench scale level. Hence, considerably greater study is needed before these processes can be fully evaluated.

The production of alcohol from molasses is existing technology which is being applied in many countries. Thus, the technical details are more reliable than for the other processes. It would appear that, if large quantities of molasses were available at $50/ton, alcohol could be produced for costs of around $1.08/gallon and sold competitively today with alcohol produced from ethylene. Unfortunately, molasses is a limited raw material in the U.S.

However, there is a strong possibility that syrup sweet sorghum can be used as a feedstock adjunct to molasses. In the Southwest, syrup sweet sorghum is harvested after the sugar cane season thereby effectively extending the use of sugar feedstock to a year-round basis. Moreover, it is known that syrup sweet sorghum can grow in many other areas of the U.S. in addition to Southern regions. In particular, sweet sorghum can grow wherever corn is grown. This raises the possibility that the sweet sorghum may be used also in place of or in conjunction with corn as a feedstock for making alcohol.

With respect to grain substrates, alcohol production from corn, milo, wheat or other grains is existing technology. The Scheller process appears to be based on such technology. Further, Scheller appears to be overly optimistic in terms of substrate costs, equipment costs, and by-product credits in arriving at a production cost of $0.94/gallon. However, the basic assumptions in his estimates are clearly spelled out and can be corrected to whatever case the person analyzing the process feasibility should wish. Other estimates based on corn, wheat and milo suggest an alcohol production price in the range of $1.27 - $2.05/gallon.

With respect to other processes, such as the Intertech process based on growth of algae in open ponds followed by production of alcohol from algal carbohydrates and the Gregor process based on the use of membranes for various separation steps, the alcohol cost estimates cannot be given serious consideration because many problems have been overlooked and neither process has been reduced to practice.

If one wished to produce alcohol in the next few years, it would appear to be feasible to do so at costs estimated to range from $1.00 to $2.00 per gallon from molasses, grain or cellulosic substrates using processes described in the Schaffer report for molasses, the Scheller process for grain, and the Madison process for sawdust.

Technological Challenges

This review of the alcohol process economy indicates that the major barrier to the conversion of biomass to alcohol fuel is in the price of the feedstock. Clearly, technologies must be evolved to utilize cheap feedstocks such as municipal and agricultural residues, etc. Utilization of these materials not only affords cheaper starting materials but their development may allow financing through municipal bond issues, thus reducing the risk and encouraging more private entrepreneurs to participate in construction of such processes. Cheap collection and biomass separation techniques are key to the solution of these problems. Also, the potential of sweet sorghum as a feedstock supplement needs to be verified and its broad base suitability for agricultural production tested.

With respect to the process technology, there are several important technologies that can use further development. These include improvement in recovery techniques and enhancement of the alcohol concentration in the fermentation in order to reduce the separation energy demands. Also, means of increasing productivities should be investigated in order to reduce fermentation equipment size.

With respect to low energy separations, alternatives to distillation should be studied. These include membrane separations, extractive fermentation, vacuum fermentation, etc. Development of thermo-tolerant alcohol producing organisms are essential to vacuum fermentation techniques. Also, alcohol-tolerant yeasts need to be developed that will give either higher rates of alcohol production for 8-10% broth concentrations or produce broths at 14-20% at similar rates.

In dealing with cheap biomass sources such as agricultural residues and municipal wastes, cheap collection costs in order to achieve economics of scale must be found as well as ways to improve the rates of cellulose saccharification either through acid or enzymatic hydrolysis of these materials. Also, ways must be developed to utilize the hemicellulose and lignin associated with the cellulose and/or starch in these materials.

Finally, byproduct credits are very important to the overall process economics. Burning the byproducts for fuel credit is not the solution. Rather, effective ways must be found of upgrading the byproducts into high value feed and chemicals. This latter area has received very little attention and must be developed if fuels from biomass are ever to be economical processes.

NOTES

3rd ANNUAL BIOMASS ENERGY SYSTEMS CONFERENCE

Colorado School of Mines
Golden, Colorado

June 4-7, 1979

ATTENDEES

A

Nick Andrianas
Rensselaer Polytechnic Institute
Dept. of Chem. & Env. Eng.
Troy, NY 12181
518/270-6376

Alex G. Alexander
University of Puerto Rico
Agr. Experiment Stateion
Rio Piedras, P.R. 00928
809/767-9705

James K. Alexander
Department of Biological Chemistry
Hahnemann Medical College
230 N. Broad Street
Philadelphia, PA 19102
215/448-1805

Billy R. Allen
Battelle - Columbus Labs
505 King Avenue
Columbus, OH 43201
614/424-4921

C. Tony Allen
EG & G
P.O. Box 1625
Idaho Falls, ID 83401
208/526-0250

John M. Allen
Battelle - Columbus Labs
505 King Avenue
Columbus, OH 43201
614/424-4848

Terence G. Alston
Argonne National Lab
9700 S. Cass Avenue
Argonne, IL 60439
312/972-3026

Kenneth R. Ames
Battelle Northwest
Box 999
Richland, WA 99352
509/946-2072

William F. Amon, Jr.
Cetus Corporation
600 Bancroft Way
Berkeley, CA 94710
415/549-3300

David A. Andrews
Woods Hole Oceanographic Inst.
Harbor Branch Foundation
R. R. 1 Box 196
Fort Pierce, FL 33450
305/465-2400

Antonios Antonopoulos
Energy & Environmental Systems Div.
Argonne National Lab
Argonne, IL 60439
312/972-3368

Edward Ashare
Dynatech R/D Company
99 Erie Street
Cambridge, MA 02139

J. M.C. Ashwanden
Prince Albert Pulp Company
1020 Place Du Canada
Montreal, Quebec H3B 2N2
514/866-9767

B

Suresh P. Babu
Inst. of Gas Technology
3424 S. State Street
Chicago, IL 60616
312/567-5758

Richard C. Bailie
Environmental Energy Engineering, Inc.
P.O. Box 4214
Morgantown, WV 26505

Karen L. Barclay
Montana Energy & MHD R&D Institute
P.O. Box 3809
Butte, MT 59701
406/494-6246; FTS 587-6246

Attendees - Biomass Energy Systems

Ernest A. Barcell
Seven Engineering Corporation
3760 Vance
Wheat Ridge, CO 80033
303/425-4239

John R. Barrett
USDA - SEA and Purdue Univ.
Agen Bldg.
West Lafayette, IN 47907
317/749-2971

Ivor Bateman
Morbark Industries
P.O. Box 1000
Winn, MI 48896
517/866-2381

Albert Beaufrere
Brookhaven National Labs
Bldg. 120
Upton, NY 11973
516/345-2094

Steven R. Beck
Dept. of Chemical Engineering
Texas Tech University
Box 4679
Lubbock, TX 79409
806/742-3553

Vicente C. Bendanillo, Jr.
Gas Research Institute
10 West 35th Street
Chicago, IL 60616
312/567-6642

Charles Bendersky
Pyros, Inc.
#2 Prof. Drive Suite 220
Gaithersburg, MD 20760
301/948-5686

John R. Benemann
University of California - Berkeley
Sanitary Engineering Research Lab
Berkeley, CA 94720
415/231-9516

Paul F. Bente, Jr.
The Bio-Energy Council
1625 Eye Street, NW - Suite 825A
Washington, DC 20006
202/833-5656

Walter L. Berry, Jr.
The Rust Engineering Company
1130 South 22nd Street, P.O. Box 101
Birmingham, AL 35201
205/254-4101

Kurt Bevernitz
Gulf Science and Technology
9009 West 67th Street
Merriam, KS 66202
913/722-3200 x370

Larry A. Bingham
Consultant
3235 6th
Boulder, CO 80302
303/447-1668

Harvery W. Blanch
University of California - Berkeley
Lawrence Berkeley Labs
Berkeley, CA 94720
415/642-1387

George F. Bonnici
Duvant (IDP - Renault)
One Old Country Road
Carle Place, NY 11514
516/248-0880

Neil L. Book
University of Missouri - Rolla
Dept. of Chemical Engineering
Rolla, MO 65401
314/341-4422

Buford L. Brauninger
12978 Bellerive Ests Dr.
Creve Coeur, MO 63141

Neal J. Briggi
Novo Laboraotries, Inc.
59 Danbury Road
Wilton, CT 06897
203/762-2401

Ronald E. Brooks
General Electric - CRD
Building K-1, Room 3B12
P.O. Box 8
Schenectady, NY 12301
518/385-8640

Barbara Broomfield
Argonne National Lab
9700 S. Cass Avenue EES012
Argonne, IL 60439
312/972-3723

Attendees - Biomass Energy Systems

Peter Brown
Bio-Gas of Colorado
5611 Kendall Ct.
Arvada, CO 80002
303/422-4354

Harold Brownell
Forintek Canada Corp.
800 Montreal Road
Ottawa, Ontario, Canada K1S 2B8

Henry R. Bungay
Rensselaer Polytechnic Institute
Dept. of Chem. & Env. Engineering
Troy, NY 12186
518/270-6376

Frank Burton
Gates Rubber Company
999 South Broadway
P.O. Box 5887
Denver, CO 80217

Robert M. Busche
DuPont Co., Cent. R&D Dept.
E356/381
Wilmington, DE 19898
302/772-3700

C

Clayton D. Callihan
Louisiana State University
Chemical Engineering Dept.
Baton Rouge, LA 70803
504/388-1426

Carol A. Camalio
Bio-Energy Council
1625 Eye Street, NW
Washington, DC 20006
202/833-5656

Stanley B. Carpenter
University of Kentucky
Dept. of Forestry
Lexington, KY 40546
606/258-4609

Michelle Carr
Golden Daily Transcript
1000 10th Street
Golden, CO 80401
303/279-5541

Charles C. Carson
Sandia Laboratories
ox 5800
lbuquerque, NM 87108
505/264-4344/6477

Robert P. Chambers
Aubrun University
Chemical Engineering
Auburn, AL 36830
205/486-4827

Clement J. Chiang
Sandia Lab
Albuquerque, NM 87108
505/265-9472

W. Howard Church
Naval Post Graduate School
Monterey, CA 93940
408/646-2471

David P. Chynoweth
Institute of Gas Technology
3424 S. State Street
Chicago, IL 60616
312/567-3715

J. Coffman
Wright - Malta Corporation
Malta Test Station, Plains Road
Ballston Spa, NY 12020

Louis F. Conde
University of Florida
School of Forest Res. & Conservation
342-B N-Z Hall
Gainesville, FL 32611
904/392-1850

Allyn W. Coombs
Real Land Inc.
293 Russel Street
Hadley, MA 01035
413/584-3751

Donna R. Cooper
Editor, Energy Information
P. I. Corporation
1375 Delaware
Denver, CO 80201
303/825-2181

Elizabeth R. Coppinger
Ecotope Group
2332 E. Madison
Seattle, WA 98112
206/322-3753

Attendees - Biomass Energy Systems

Raymond Costello
Mittel Hauser Corporation
400 1st Street, NW
Washington, DC 20001
202/783-1061

George A. Coulman
Cleveland State University
Dept. of Chemical Engineering
Cleveland, OH 44115
216/687-2569

Joe R. Cowles
University of Houston
Department of Biology
Houston, TX 77004
713/749-3135

John L. Cox
Gas Research Institute
10 West 35th Street
Chicago, IL 60616
312/567-6640

John E. Cummings
Electric Power Research Institute
3412 Hillview Avenue, P.O. Box 10412
Palo Alto, CA 94303
415/855-2166

D
Rathin Datta
Exxon Research
Bldg. 8-216
Linden, NJ 07036
201/420-5600

John Davis
Iotech Limited
220 Laurier Avenue West
Ottawa, Ontario, Canada K1P 5Z9
613/238-4126

Leland G. Desmon
Peabody Gordon-Piatt
Box 650
Winfield, KS 67156
316/221-4770

Ray Desrosiers
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1436; FTS 327-1436

Richard E. DeZeeuw
California Energy Commission
1111 Howe Avenue
Sacramento, CA 95825
916/920-6033

Donald D. Dickenson
Holly Sugar Corporation
P.O. Box 1052
Colorado Springs, CO 80901
303/471-0123

James P. Diebold
Department of the Navy
Naval Weapons Center - Code 3244
China Lake, CA 93555
714/939-7242

Robert C. Dobbs
Canadian Forestry Service
Department of Envornment
Ottawa, Ontario, Canada K1A 0E7
997-1878

Richard D. Doctor
Science Applications, Inc.
1211 W. 22nd Street - Suite 901
Oakbrook, IL 60521
312/655-5960

Jimmie D. Dodd
Texas Ag. Expeirment Station
TAMRF, F. E. Box H
College Station, TX 77843
713/846-7731

Linda Sutliff Dolan
Seattle City Light
1015 Thrid Avenue, Rm. 926
Seattle, WA 98104
206/625-3590

Larry Douglas
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1470; FTS 327-1470

Gordon L. Dugan
University of Hawaii at Manoa
Dept. of Civl Engineering
Holmes Hall 383 - 2540 Dole Street
Honolulu, HI 96822

Carl R. Duisberg
Technikos
1120 19th Street, NW - Suite 601
Washington, DC 20036
202/466-8825

E
W. Brian Earl
University of Canterbury
Christurch, New Zealand

Attendees - Biomass Energy Systems

Donald P. Easler
Institute of Gas Technology
1825 K Street NW - Suite 218
Washington, DC 20006
202/785-3511

Richard W. Edsinger
Boeing Computer Services Co.
565 Andover Park West, M/S 9C-01
Tukwila, WA 98188
206/575-5068

Douglas C. Elliott
Pacific Northwest Lab
Battelle Memorial Institute
P.O. Box 999
Richland, WA 99352
509/946-2026

Sabri Ergun
Lawrence Berkeley Lab
University of California - Bldg. 77F
Berkeley, CA 94720
415/486-6666

Paul R. Erickson
Rexnord Inc.
5103 W. Beloit Road
Milwaukee, WI 53214
414/643-2688

Veryl G. Eschen
Science Applications Inc.
7935 E. Prentice Avenue
Englewood, CO 80111
303/773-6900

Igor Evangelidi
SNC Montreal
1 Complexe Desjordins
Case Postale 10
Montreal, Canada H5B 1CB

F

R. S. Farnham
Soil Science Department
University of Minnesota
St. Paul, MN 55108
612/373-1447

Herman F. Feldmann
Battelle - Columbus Laboratories
505 King Avenue
Columbus, Ohio 43201
614/424-4732

Peter J. Felker
Dept. of Soil and Environmental Sciences
University of California - Riverside
Riverside, CA 92521
714/787-3143

John Ferchak
University of Pennsylvania
323 Anatomy/Chemistry/G3
Philadelphia, PA 19104
215/243-8038

Oliver A. Fick
International Paper Company
P. O. Box 3860
Portland, OR 97208
503/243-3266

Carlos Figueroa
Lawrence Berkeley Lab. Bldg. 934-16
University of California
Berkeley, CA 94720
415/486-6766

Bryant E. Fitch
Auburn University
Chemical Engineering
Auburn, AL 36830
205/826-4827

John F. Flagg
UOP, Inc.
10 UOP Plaza
Des Plaines, IL 60016
312/391-3331

Michael C. Flickinger
Purdue University
Laboratory of Renewable Resources Eng.
A. A. Potter Center
West Lafayette, IN 47907
317/494-6151 Sta. 227

Kenneth R. Foote
General Electric Co.
Court Street Plant #2
Syracuse, NY 13221
315/456-7545

Malcolm D. Fraser
Inter Technology/Solar Corp.
100 Main Street
Warrenton, VA 22186
703/347-7900

Attendees - Biomass Energy Systems

Douglas J. Frederick
North Carolina State University
School of Forest Resources
Box 5488
Raleigh, NC 27650
919/737-2891

Robert G. Friel
Trinity College
Department of Computer Science
Dublin 2 Ireland
772941 Ext. 1765

Janis T. Fritts
Pyros Inc.
#2 Prof. Drive
Gaithersburg, MD 20760
301/948-5686

G

James L. Gaddy
University of Missouri
Rolla, MO 65401
314/341-4416

Ekkehart Gasper
PRC/ Energy Analysis Co.
7600 Old Springhouse Rd.
McLean, VA 22102
703/893-1820

Abraham P. Gelbein
Chem Systems Inc.
Research Center
One Evans Street
Fairfield, NJ 07006
201/575-8820

Richard K. Genung
Oak Ridge National Laboratory
P. O. Box X, CTD
Oak Ridge, TN 37803
615/574-6830

Joel C. Goldman
Woods Hole Oceanographic Institution
Woods Hole, MA 02543
617/548-1400

Frank J. Goldner
U. S. Department of Energy
Forrestal Bldg.
Washington, D. C. 20585
202/252-6503

Irving S. Goldstein
North Carolina State University
Dept. of Wood & Paper Science
Raleigh, NC 27650
919/737-3181

Judith J. Gordon
The MITRE Corporation/Metrek Division
1820 Dolley Madison Boulevard
McLean, VA 22102
703/827-6000

Michael S. Graboski
Colorado School of Mines
Golden, CO 80401
303/279-3033 Ext. 2769

Daniel Grafstein
Exxon Research & Engineering Co.
P. O. Box 8
Linden, NJ 07036
201/474-3111

Marvin I. Greene
Chem Systems Inc.
1 Evans St.
Fairfield, NJ 07006
201/575-8820

Harry P. Gregor
Columbia University
Dept of Chemical Eng. & Applied Chemistry
New York, NY 10027
212/280-4716

Karel Grohmann
Jet Propulsion Laboratory 125-124
California Institute of Technology
4800 Oak Grove Dr.
Pasadena, CA 91103
213/354-7215

J. Charles Grosskreutz
SERI
1536 Cole Boulevard
Golden, CO 80401
303/231-1303

H

James Halligan
University of Missouri
Rolla, MO 65401
314/341-4151

Attendees - Biomass Energy Systems

James W. Hanover
Michigan State University
Dept. of Forestry
East Lansing, MI 48824
517/355-0090

Rolf C. Hansen
Swedish Trade Office
P. O. Box 27459
Houston, TX 77027
713/626-8850

Lars Hansson
National Swedish Board for Energy Source Dev.
Box 1103
S-163 12 Spanga
Sweden
08752 03 60

John B. L. Harkness
Argonne National Laboratory
EES Division
Argonne, IL 60439
312/972-7636

Ira M. Harmon
Steamboat Solar Systems
Box 323
Steamboat Springs, CO 80477
303/879-3076

Jerome P. Harper
Argonne National Laboratory
Energy & Environmental Systems Div.
9700 S. Cass Avenue
Argonne, IL 60439
312/972-3399

S. M. Harris
U. S. Dept. of Energy
Washignton, DC 20545
202/376-9425

Carolyn M. Hart
Sandia Labs
Box 5800
Albuquerque, NM 87185

Andrew G. Hashimoto
U. S. Dept. of Agriculture
Agrigultural Eng. Research Unit
P. O. Box 166
Clay Center, NE 68933
402/762-3241

Robert J. Hathaway
A. E. Staley Co.
Decatur, IL 62525
217/423-4411 Ext. 2160

R. Wayne Hendren
Research Triangle Institute
Chemistry and Life Sciences Division
P. O. Box 12194
Research Triangle Park, NC 27709
919/541-6518

Mark Henriquez
Lawrence Berkeley Laboratory
1 Cyclotron Road
Bldg. 90 Room 3125
Berkeley, CA 94720
415/486-6933

Julie E. Hermsen
Rensselaer Polytechnical Institute
Dept. of Chemical & Env. Engineering
Troy, NY 12181
518/270-6376

Charles E. Hewett
The Resource Policy Center
Dartmouth College
Hanover, NH 03755
603/646-3551

Bill Hoagland
SERI
1536 Cole Blvd.
Golden, CO 80401
303/231-1368

Michael Hohmann
Mid-American Solar Energy Center
1256 Trapp Rd.
Eagan, MN 55121
612/452-5830

Mark Holtzapple
Univeristy of Pennsylvania
197 Towne Bldg
Philadelphia, PA 19104
215/243-8354

Richard Hood
GM Transportation Systems Division
Warren, MI 48090
313/575-8520

Attendees - Biomass Energy Systems

Irwin Hornstein
Office of Energy, Ag. Int'l Dev.
Department of State
Washington, DC 20523
703/325-9020

Randel L. Hoskins
Rocket Research Company
York Center
Redmond, WA 98052
206/885-5000

Emile Houle
Bechtel National, Inc.
50 Beale Street
San Francisco, CA 94105
415/768-8363

T. A. Huibers
Hydrocarbon Research, Inc.
Research & Development Division
P. O. Box 6047
134 Franklin Corner Road
Lawrence Township, NJ 08648

Arthur E. Humphrey
University of Pennsylvania
107 Towne Bldg.
Philadelphia, PA 19104
215/243-7244

Danial L. Ingold
Creare Incorporated
Box 71
Hanover, NH 03755
603/643-3800

Robert Inman
SERI
1536 Cole Boulevard
Golden, CO 80401
303/231-1481

Ron Isaacson
Argonne National Laboratory
9700 S. Cass Avenue
Argonne, IL 60439
312/972-7542

Robert J. Isler
Brookhaven National Laboratory
Bldg. 120
Upton, NY 11943
516/345-2405

Toru Iora
The Aerospace Corporation
20030 Century Boulevard
Germantown, MD 20767
301/428-2754

J

Dan Jantzen
SERI
1536 Cole Boulevard
Golden, CO 80401
303/231/1203

Thomas W. Jeffries
USDA - Forest Service
Forest Products Laboratory
P. O. Box 5130
Madison, WI 53705
608/257-2211, Ext. 462

Robert F. Johnston
Johnston Associates
Pretty Brook Rd.
Princeton, NJ 08540
609/924-3131

Roy H. W. Johnston
Trinity College
18 Westland Road
Dublin 2 Ireland
766210

Everett N. Jones
TSISSA Corporation
New Lane
West Tisbury, MA 02575

Jerry L. Jones
SRI International
333 Ravenswood Ave.
Menlo Park, CA 94025
415/326-6200, Ext. 3534

Keith Jones
E. G. & G.
P. O. Box 1625
Idaho Falls, Idaho 83401
208/526-9876

Maurice W. Jones
Hydrocarbon Research, Inc.
134 Franklin Corner Road
P. O. Box 6047
Lawrenceville, NJ 08648
609/896-1300

Attendees - Biomass Energy Systems

David C. Junge
Oregon State University
Dept. of Mechanical Engineering
Corvallis, OR 97331
503/754-4902

K

Anthony Y. Kam
SRI International
333 Ravenswood Ave.
Menlo Park, CA 94025
415/326-6200

Keith E. Kawaoka
The Aerospace Corporation
20030 Century Blvd.
Germantown, MD 20767
301/948-0984

Gary R. Kent
Universal Services Unlimited
Box 151
Thompson Falls, MT 59873
406/827-4722

Sonja N. Kerr
Pennsylvania State University
135 Land & Water
University Park, PA 16802
814/803-0120

James D. Kerstetter
National Center for Appropriate Technology
P. O. Box 3838
Butte, MT 59701
406/494-4572

Jeannette W. King
Denver Research Institute
2390 So. York St.
Denver, CO 80210
303/753-2911

Robert G. Kispert
Dynatech R/D Company
99 Erie St.
Cambridge, MA 02139
617/868-8050

Donald L. Klass
Institute of Gas Technology
3424 South State Street
IIT Center
Chicago, IL 60616
312/567-3783

Herbert E. Klei
University of Connecticut
Chemical Engineering Dept.
Storrs, CT 06268
203/486-2756

H. John Knapp
Grenco, Inc.
2035 SW 58th Ave.
Portland, OR 97221
503/297-2536

David L. Koch
Desert Research Institute
Bioresources Center
P. O. Box 60220
Reno, Nevada 89506
702/673-4750, Ext. 232

Victor R. Koch
EIC Corporation
55 Chapel Street
Newton, MA 02158
617/965-2710

Herbert M. Kosstrin
Energy Resources Co., Inc.
185 Alewife Brook Parkway
Cambridge, MA 02138
617/661-3111

A. P. Kozacik
30361 Rainbow Hill Rd.
Golden, CO 80401

Timothy T. Ku
University of Arkansas
Dept of Forestry
Monticello, AR 71655
501/367-2835

James L. Kuester
Arizona State University
College of Engineering
Tempe, AZ 85281
602/965-5071, Ext. 3313

Attendees - Biomass Energy Systems

L Fred M. Laing
University of Vermont
Botany Department
Burlington, VT 05401
802/656-2930 or 899-4923

Donald M. LaRue
EG & G Idaho
P. O. Box 1625
Idaho Falls, ID 83401
208/526-0509

Jai Moon Lee
Chemical Engineering Department
Cleveland State University
Cleveland, OH 44115
216/687-3539

John C. Lee
Brookhaven National Lab
Associated Universities Inc.
Upton, NY 11973
516/345-2248

Yoon Y. Lee
Auburn University
Chemical Engineering Department
Auburn, AL 36830
205/826-4827

Terry Lenz
Colorado State University
Department of Agr. & Chem. Engr.
Fort Collins, CO 80523
303/491-5252

Wayne A. LePori
Texas A&M University
Agricultural Engineering Department
College Station,TX 77843
713/845-1131

Michael P. Levi
North Carolina State University
School of Forest Resources
Raleigh, NC 27607
919/737-3386

Daniel J. Lizdas
Hamilton Standard/United Technologies Corp.
Windsor Locks, CT 06096
203/623-1621

John H. Lohnes
Dartmouth College
Hanover, NH 03755
603/646-3551

Luis Eduardo Lima
FDTE
Cidade Universitaria
Armando de Salles Oliveira
Setor Amarelo, Bloco N-2
Sao Paulo-SP
011/211-4308

James C. Linden
Colorado State University
Agr. & Chem. Engineering
Fort Collins, CO 80523
303/491-8291

Edward S. Lipinsky
Battelle
505 King Avenue
Columbus, OH 43201
614/424-6526

Hemrik Lumdberg
National Swedish Board for Energy Source Development
ESSIWGESTRAHKET II
Stockholm, Sweden S-12647
08-56 2710

Marianne K. Luppold
Arthur D. Little, Inc.
Acorn Park
Cambridge, MA 02140
617/864-5770

M Elizabeth A. MacDonald
TSISSA Corp.
New Lane
West Tisbury, MA 02575

Thomas R. McAdams
BSP Division, Envirotech Corp.
One Davis Drive
Belmont, CA 94002
415/592-4060

James F. McBrayer
Oak Ridge National Laboratory
P. O. Box X
Oak Ridge, TN 37830
615/574-7317 FTS 624-7317

Thomas A. McCaskey
Auburn University
Animal & Dairy Science
Auburn, AL 36830
205/826-4020

Attendees

Attendees - Biomass Energy Systems

Gary David McGinnis
Mississippi State University
P.O. Drawer FP
State College, MS 39762
601/325-2116

Kenneth W. McLeod
Savannah River Ecology Lab
Drawer E
Aiken, SC 29801
803/450-2472

Gary C. Magruder
University of Missouri - Rolla
VMR Chem. Engr. Dept.
Rolla, MO 65401
314/341-4460

Otis J. Manar
The Rust Engineering Company
P. O. Box 101
Birmingham, AL 35201
205/254-4534

Mary Mandels
U.S. Army NARAD COM
FSL
Natick, MA 01760
617/653-1000 ext.2381

George W. Mann
The Dow Chemical Company
A-2301 Building
Freeport, TX 77541
713/238-4257

Barbara Mantovani
Purdue University
Student-Agricultural Engineering Dept.
West Lafayette, IN 47901
317/749-2971

Evandro Mantovani
Purdue University - Student
Agricultural Engineering Dept.
West Lafayette, IN 47901
317/749-2971

Jodie B. Marler
Arkansas Energy Conservation & Policy Office
960 Plaza West Bldg.
Little Rock, AR 72205
501/371-1730

Paul Marnell
Brookhaven National Lab
Building 120
Upton, NY 11973
516/345-2405

Ana Maria Martinez
Instituto De Investigaciones Electricas
APDO. Postal 475
Cuernavaca, Morelos Mexico
41351

Daniel H. Mason
California Department of Food & Agr.
1220 N.St. #104
Sacramento, CA 95814
916/322-6832

Richard C. Mason
CMDNJ-Rutgers Medical School
P. O. Box 101
Piscataway, NJ 08854
201/463-4553

Nancy Mathews
Technical Consultant to
Congressman Richard Ottinger
House Science Com. B-374 RHOB
Washington, DC 20515
202/225-9117

Lon J. Mathias
Auburn University
Chemistry Department
Auburn, AL 36830
205/826-4043

Helio Mattar
Instituto de Pesquisas Tecnològicas do
Estado de S. Paulo S/A IPT
Cidade Universitària
05508 São Paulo SP BRAZIL
011/268-2211

G. L. Mauldin
New Mexico Solar Energy Institute
Box 3 SOL
Las Cruces, NM 88003
505/646-1846

Bill Maze
Proteus, Inc.
321 So. Bridge
Visalia, CA 93277
209/733-5423

R. Kenneth Merkey
Fibrex, Inc.
293 Russell St.
Hadley, MA 01035
413/584-3751

Attendees - Biomass Energy Systems

Eilif V. Miller
U.S. Department of Agriculture
SEA, Cooperative Research
Washington, DC 20250
202/447-6347

Tom Milne
SERI
1536 Cole Blvd.
Golden, CO 80401
303/231-1483

Akira Mitsui
University of Miami
4600 Rickenbacken Causeway
Miami, FL 33149
305/350-7341

Edwin W. Mogren
Colorado State University
Department of Forest and Wood Sciences
Fort Collins, CO 80523
303/491-6519

Bland S. Montenecourt
Rutgers State University of New Jersey
P. O. Box 231
New Brunswick, NJ 08903
201/932-9829

Antonio R. Moreira
Colorado State University
Agr. & Chem. Engineering Dept.
Fort Collins, CO 80523
303/491-5252

Frederick E. Moreno
Energy Resources Co.
445 So. San Antonio Rd.
Los Altos, CA 94022
415/941-3931

Gregory Morris
University of California
Energy & Resources Group
Bldg. T-4 Rm. 100
Berkeley, CA 94720
415/642-6703 or 1640

Lyle K. Mudge
Battelle-Northwest
P. O. Box 999
Richland, WA 99352
509/946-2268

Vincent G. Murphy
Department of Agr. & Chem. Engineering
Colorado State University
Fort Collins, CO 80523
303/491-5252

Wayne K. Murphey
Texas Ag Experiment Station
TAMU - Forest Science
College Station, TX 77843
713/845-5033

Richard C. Myerly
Union Carbide Corporation
P.O. Box 8361
So. Charleston, WV 25303
304/747-4034

N Eric T. Nelson
Gilbert Associates Inc.
Box 1498
Reading, PA 19608
215/775-2601

Franklin H. Newman
The Aerospace Corporation
20030 Century Blvd.
Germantown, MD 20767
301/428-2774

Jim W. Niewald
Texas Energy Advisory Council
7703 N. Lamar
Austin, TX 78752
512/475-5588

Richard E. Nightingale
Battelle - Northwest
P.O. Box 999
Richland, WA 99352
509/946-2597

Edward J. Nolan
General Electric Co.
Box 8555
Philadelphia, PA 19101
215/243-8354

Wheeler J. North
California Institute of Technology
Kerckhoff Marine Laboratory
101 Dahlia Street
Corona del Mar, CA 92625
213/795-6811; 714/673-9894

Attendees - Biomass Energy Systems

John M. Nystrom
Northeast Solar Energy Center
70 Memorial Drive
Cambridge, MA 02142
617/661-3500

O
Ed O'Hair
The Montana Energy and MHD R&D Institute
P.O. Box 3809
Butte, MT 59701
406/494-6293

Yildirim "Bill" Omurtag
University of Missouri - Rolla
301 Harris Hall
Rolla, MO 65401
314/341-4556

Daniel O'Neil
Georgia Tech
EES
Atlanta, GA 30332
404/894-9641

Dennis M. Oster
Bonneville Power Adminstration
P.O. Box 3621
Portland, OR 97208
503/234-3361 x4037

Ralph P. Overend
Biomass Energy Project, National Res.
Room 206, M-50, Montreal Road
Ottawa, Ontario, Canada K1A 0R6
613/993-3405

P
William Pace
Bio-Gas of Colorado
5611 Kendall Ct.
Arvada, CO 80002

Birgitta Palmberger
National Swedish Board of Energy Source Dev.
Box 1103 Spanga
Stockholm, Sweden
08/7520360

Worth H. Percival
United Stirling
1629 K Street, NW Suite 801
Washington, DC 20006
202/466-7286

Gene R. Petersen
Jet Propulsion Lab
[illegible]800 Oak Grove Dr.
[illegible]sadena, CA 91103
213/354-3791

Timothy D. Placek
Auburn University
Chemical Engineering
Auburn, AL 36830
205/826-4827

Fannie Posey
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1202; FTS 237-1202

John H. Posselius
Agricultural Engineering
Michigan State University
East Lansing, MI 48824
517/353-9448

Stelios M. Pneumaticos
Energy - Mines & Resources Canada
580 Booth Street
Ottawa Ontario Canada

James L. Prouty
Arthur G. McKee & Company
10 S. Riverside Plaza
Chicago, IL 60606
312/454-3857

E. Kendall Pye
University of Pennsylvania
Dept. of Biochemistry & Biophysics
Medical School
Philadelphia, PA 19104
215/243-8038

R
Robert Rabson
DOE/Div. of Biological Energy Research
Germantwon Bldg.
Washington, DC 20545
202/233-2873; FTS 233-2873

John M. Radovich
University of Oklahoma
202 W. Boyd - Rm. 23
Norman, OK 73019
405/325-5811

Harvey L. Ragsdale
Emory University
Biology Department
Atlanta, GA 30322
404/329-4218

J. W. Ranney
Oak Ridge National Laboratory
P.O. Box X
Oak Ridge, TN 37830
615/574-7364; FTS 624-7364

Attendees - Biomass Energy Systems

Tom Reed
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1484; FTS 327-1484

Terrence A. Renner
Argonne National Lab
7900 S. Cass Avenue
Argonne, IL 60439
312/972-4535

Ronald R. Rich
Minnesota Energy Agency
150 E. Kellogg, #980
St. Paul, MN 55101
612/296-6893

B. C. Robb
Sverdrup & Parcel & Associates, Inc.
800 N. 12th Blvd.
St. Louis, MO 63101
314/436-7600

P. A. Holly Roberts
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1391; 327-1391

Ronnie S. Roberts
Georgia Tech
School of Chemical Engineering
Atlanta, GA 30332
404/894-2889

Ernest E. Robertson
The Biomass Energy Institute Inc.
304-870 Cambridge Street
Winnipeg, Manitoba, CANADA R3M 3S7
204/284-0472

Jean Robieux
Scientifique Laboratoires de Marcoussis CR-CGE
Route de Nozay
91460 Marcoussis FRANCE
(1) 901.20.02

Robert W. Roig
International Research & Tech. Inc.
7655 Old Springhouse Rd.
McLean, VA 22101
703/821-8810

Gary L. Rolfe
University of Illinois
211 Mumford Hall
Urbana, IL 61801
217/333-7489 or 0240

David S. Ross
SRI International
Menlo Park, CA 94025
415/326-6200 x2430

Donald L. Ross
SRI International
333 Ravenswood Avenue
Menlo Park, CA 94025
415/326-6200 x3195

George O. Rudkin
ICI Americas Inc.
ICI Americas Inc.
Wilmington, DE 19897
302/575-8301

Ben Russel
ECON Company
P.O. Box 821
Alexander City, AL 35010
205/329-8424

John H. Ryther
Woods Hole Oceanographic Institution
Woods Hole, MA 02543
617/548-1400

S Roy M. Sachs
University of California
Department of Env. Hort.
Davis, CA 95616
916/

Robert L. Sajdak
Forestry Department
Michigan Tech. University
Houghton, MI 49931
906/487-2338

John E. Sanderson
Dynatech R/D Company
99 Erie Street
Cambridge, MA 02139
617/868-8050

Robert L. San Martin
US DOE
600 E Street NW
Washington, DC 20545
202/376-4761

Kathryn A. Saterson
Arthur D. Little, Inc.
Acorn Park
Cambridge, MA 02174
617/864-5770

Attendees - Biomass Energy Systems

Larry Schaleger
Lawrence Berkeley Lab
University of California Bldg. 77F
Berkeley, CA 94720
415/486-5944

Susan Schellenbach
Bio-Gas of Colorado
5611 Kendall Ct.
Arvada, CO 80002

Gary F. Schiefelbein
Battelle - Northwest
P.O. Box 999
Richland, WA 99352
509/946-2140

Fred A. Schooley
SRI International
333 Ravenswood Avenue
Menlo Park, CA 95014
415/326-6200

Herbert A. Schroeder
Colorado State University
Forest & Wood Science
Ft. Collins, CO 80521
303/491-6631

Torghy S.R. Schütt
Commission on Energy R&D
SVEAUAGEN 9-11, S-11157
Stockholm, Sweden
08-7633741

Eion G. Scott
General Electric Company
Court Street Plant Z
Syracuse, NY 13066
/456-7740

Jerry Scott
ECON Company
P.O. Box 821
Alexander City, AL 35010
205/329-8424

L. John Sealock, Jr.
Battelle - Northwest
P.O. Box 999
Richland, WA 99352
509/946-3635

Manu Seth
Lawrence Berkeley Lab
University of California - Bldg. 77E
Berkeley, CA 94720
415/231-9570

Willian H. Seward
Mittlehauser Corporation
5120 Belmont Rd. Suite G
Downer's Grove, IL 60515
312/964-8164

Robert A. Shade
Boise Cascade Corporation
One Jefferson Square
Boise, ID 83707
208/384-7217

John E. Shively
Lawrence Berkeley Lab
University of Calfornia - Bldg. 77F
Berkeley, CA 94720
415/486-5944

Sharon Shoemaker
Cetus Corporation
600 Bancroft Way
Berkeley, CA 94710
415/549-3300

S. E. Shumate, II
Oak Ridge National Laboratory
P.O. Box X
Oak Ridge, TN 37830
615/574-6832

William Shustor
Rensselaer Polytechnic Institute
Dept. of Chemical & Env. Engineering
Troy, NY 12180
518/270-6376

C. Ross Silversides
Biomass Energy Project, National Res.
Room 206, M-50, Montreal Road
Ottawa, Ontario, Canada K1A 0R6
613/993-3405

Anthony J. Sinskey
Massachusetts Institute of Technology
Dept. of Nutrition & Food Science
Rm. 56-121
Cambridge, MA 02139
617/253-6721

James N. Skeen
Fernbank Science Center
156 Heaton Park Dr., NE
Atlanta, GA 30307
404/378-4311

Attendees - Biomass Energy Systems

Geoffery Slaff
University of Pennsylvania
197 Towne Bldg.
Philadelphia, PA 19104
215/243-8354

Richard L. Smit
General Motors Corporation
Technical Center
Warren, MI 48090
313/575-8433

Clayton Smith
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1200; FTS 327-1200

John T. Smolik
Corning Glass Works
Corning, NY 14830
607/947-4220

Todd C. Sneller
Nebr. Ag. Products Inc. Utilization Comm.
301 Centennial Mall South
Lincoln, NE 68509
402/471-2941

Michael Soltys
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1476; FTS 327-1476

Thomas C. Sommer
Mueller Associates, Inc.
1900 Sulphur Spring Rd.
Baltimore, MD 21227
301/247-5666

William E. Sopper
Pennsylvania State University
108 Land & Water Research Bldg.
University Park, PA 16802
814/863-0291

Thomas J. Stadt
Michigan State University
126 Natural Resource Bldg.
East Lansing, MI 48824
517/355-0090

G. Stanford
Agri-City, Inc.
Greenhills Center
R. 1, B 861
Cedar Hill, TX
214/296-1955

K. Steinbeck
University of Georgia
School of Forest Resources
Athens, GA 30602
404/542-1376

Richard S. Stevenson
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1094

Richard P. Stringer
Gilbert Associates, Inc.
P.O. Box 1498
Reading, PA 19603
215/775-2600 x4300

Carl L. Strojan
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1093

Herb Suer
Desert Research Institute
P.O. Box 60220
Reno, NV 89506
702/972-1676

Barbara A. Summers
Biomass Energy Project, National Res.
Rm. 206, M-50, Montreal Road
Ottawa, Ontario, Canada K1A 0R6
613/993-3405

Donald W. Sundstrom
University of Connecticut
N-139
Storrs, CT 06268

T Abu Talib
MITRE Corporation/METREK Division
1820 Dolley Madison Blvd.
McLean, VA 22102
703/827-6625

Roy E. Taylor
University of Idaho
Dept. of Agricultural Engineering
Moscow, ID 83843
208/885-6182

George T. Tsao
Laboratory of Renewable Resources Eng'g
Purdue University
West Lafayette, IN 47907
317/494-6151 x233

Attendees - Biomass Energy Systems

Wayne Turnacliff
Bio-Gas of Colorado
5611 Kendall Ct.
Arvada, CO 80002
303/422-4354

Jim L. Turpin
University Arkansas
Chemical Engineering Dept.
Fayetteville, AR 72701
501/575-4951

U David M. Updegraff
Colorado School of Mines
Golden, CO 80401
303/279-0300 x2633

V Charles W. Vail
Ultrasystems, Inc.
7926 Jones Branch Dr.
McLean, VA 22101
703/821-1960

Robert I. Van Hook
Oak Ridge National Laboratory
P.O. Box X, Bldg. 1505
Oak Ridge, TN 37830
615/574-7374

Fred Varani
Bio-Gas of Colorado
5611 Kendall Ct.
Arvada, CO 80002
303/422-4354

Walter Vergars
Promon International Services
540 Madison Avenue
New York, NY 10022
607/256-4403

Ruxton Villet
Solar Energy Research Institute
1536 Cole Bvld.
Golden, CO 80401
303/231-1000; FTS 327-1000

W Thomas P. Wagner
Pacific Gas & Electric Co.
77 Beale Street - Rm. 2328
San Francisco, CA 94106
415/781-4211 x1501

Thomas A. Walbridge
V.P.I. & S.U.
210 Cheatam Hall
Blacksburg, VA 20461
703/961-6391

Richard B. Wallace
U.S. Dept. of Energy - Region X
1992 Federal Bldg.
Seattle, WA 98174
206/442-7290

Edward I. Wan
Science Applications, Inc.
1764 Old Meadows Lane
McLean, VA 22101
703/821-4563

Daniel E.C. Wang
MIT
Room 16-114A, Dept. of Nutrition & Food Science
Cambridge, MA 02139
617/253-2126

Sprague H. Watkins
Crown Zellerbach
Central Research
Camas, WA 98607
206/834-4444

Paul F. Weaver
Solar Energy Research Insitute
1536 Cole Blvd.
Golden, CO 80401
303/231-1928; FTS 327-1928

Nathan E. Welch
Sverdrup/ARO Inc.
Tullahoma, TN 37388
615/455-1186

Don H. White
University of Arizona
Dept. of Chemical Engineering
Tucson, AZ 85721
602/626-5280

Richard A. Wilde
DOE
Idaho Operations Office
550 2nd Street
Idaho Falls, ID 83401
FTS 583-1807

LaVergne D. Williams
Harbor Branch Foundation
Rt. 1, Box 196
Ft. Pierce, FL 32450
/465-2400

Lynn A. Williams
Dept. of Vit. & Enology
University of California
Davis, CA 95610
916/752-7706

Attendees - Biomass Energy Systems

John Willson
Burns & Roe, Inc.
1850 K Street, NW - Suite 220
Washington, DC 20006
202/659-2690

Donald Lee Wise
Dynatech R/D Company
99 Erie Street
Cambridge, MA 02139
617/868-8050

Rex A. Wisehart
Castlewood Corporation
709 W. Littleton Blvd.
Littleton, CO 80120
303/794-4366

Wolfgang Witly
Am Hirtenhaus 3
8901 Bibung
West Germany
0821/488038

Gerry W. White
Energy Management Consultants
7956 W. 5th Ave.
Lakewood, CO 80226
303/232-0522

Ron White DOE
DOE - Div. of Advanced Energy Systems Policy
Rm 6 E 068 Forestal Bldg.
Washington, DC 20585
202/252-6433

Zachary S. Wochok
Plant Resources Institute
400 Wakara Way
Salt Lake City, UT 84108
801/466-4741

Joseph C. Woodford
Woodford Manufacturing Company
2121 Waynoka Road
Colorado Springs, CO 80915
303/574-0600

Y
Merold Yates
Illinois Farm Bureau
1701 Towanda
Bloomington, IL 61701
309/557-3251

Jackson Yu
Bechtel National, Inc.
50 Beale Street
San Francisco, CA 94105
415/768-2971

Z
Oskar R. Zaborsky
National Science Foundation
1800 G Street, NW
Washington, DC 20550
202/632-7398

Gerald L. Zachariah
University of Florida
Agriculture Engineering Dept.
9 Frasier Rogers Hall
Gainesville, FL 32611
904/392-1864

SUPPLEMENT

Otis P. Armstrong
US Resourc Development
131 S. Jefferson Bldg. 7
Loveland, CO 80537
303/667-5252

Gary L. Baughman
Colorado School of Mines Research Inst.
P.O. Box 112
Golden, CO 80401
303/279-2581 x247

Frank Baylin
Solar Energy Research Insitute
1536 Cole Blvd.
Golden, CO 80401
303/231-1028; FTS 237-1028

Bill Belew
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1223; FTS 237-1223

William D. Bensema
National Bureau of Standards
325 S. Broadway
Boulder, CO 80303
303/499-1000 x3465

Romeu Boto Dantas
Coperbo Companhia Permambucana de
Borracha Sintetica
Rua do Hospicio, 601
231-1290 Recife, PE Brazil

Ronald L. Dickenson
SRI International
333 Ravenswood Avenue
Menlo Park, CA 94025
415/326-6200

Attendees - Biomass Energy Systems

Sam Flaim
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1065; FTS 327-1065

Charles O. Galvin
Texas Energy Advisory Council
7703 N. Lamar
Austin, TX 78752
512/475-5588

Joseph T. Hamrick
Aerospace Research Corporation
5454 Jae Valley Road
Roanoke, VA 24014
703/342-2961

Paphael Katzen
Raphael Katzen Associates
1050 Delta Avenue
Cincinnati, OH 45208
513/861-7500

Stephen C.T. Lien
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1925; FTS 327-1925

James Nolfi
Associates In Rural Development
109 S. Winnocski - Suite 203
Burlington, VT 05401
802/658-3890

Jack E. Olson
SPM Group Inc.
14 Inverness Dr. E, Unit 6B
Englewood, CO 80112
303/770-1201

John T. Pfeffer
University of Illinois
Dept. of Civil Engineering
Bldg. 3230
Urbana, IL 61801
217/333-6965

Phactuel M. Rego
Organization of American States
1735 I Street, NW Room 1140
Washington, DC 20006
202/381-8117

Jody E. Schultz
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1872; FTS 327-1872

Michael Seibert
Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
303/231-1879; FTS 327-1879

Lyle F. Hodde
Rural Route 2
Hamberg, IA 51640
816/389-1324

Robert L. Garten
Catalytica Associates Inc.
3255 Scott Blvd. - Suite 7E
Santa Clara, CA 95051
408/243-9243

SUPPLEMENT II

Attendees - Biomass Energy Systems

Beverly J. Berger
United States DOE
Washington, DC 20545
202/376-4876

David H. Dawson
Forestry Sciences Laboratory
U.S. Department of Agriculture
P. O. Box 898
Rhinelander, WI 54501
715/362-7474

David Duncan
Holly Sugar Corporation
P. O. Box 1052
Colorado Springs, CO 80901
303/471-0123

Alvin G. Keene
Gorham International Inc.
Gorham, ME 04038
207/892-6761

Perry L. McCarty
Stanford University
Civil Engineering Department
Stanford, CA 94305
415/497-3504

Andrew C. Nyce
Gorham International Inc.
Gorham, ME 04038
207/892-6761

Agenda

3rd ANNUAL BIOMASS ENERGY SYSTEMS CONFERENCE

Colorado School of Mines
Golden, Colorado

June 4-7, 1979

AGENDA

CHAIRMAN: Dr. Robert San Martin
Department of Energy

TECHNICAL CHAIRMAN: Larry Douglas
Solar Energy Research Institute

Monday, June 4, 1979

5:00 - 9:00 p.m.	-Registration	Twin Towers

Tuesday, June 5, 1979

7:30 - 9:00 a.m.	Registration	Green Center Lobby

SESSION I: INSTITUTIONAL OVERVIEW

Green Center

L. J. Douglas, SERI
Chairman

9:00 - 9:15 a.m.	Welcome & Announcements	L. J. Douglas, SERI
9:15 - 10:00 a.m.	Keynote Address	R. Ottinger, (D-NY)
10:00 - 10:15 a.m.	COFFEE BREAK	
10:15 - 10:40 a.m.	DOE Overview	R. San Martin, DOE
10:40 - 11:05 a.m.	DOE Policy Statement	S. Harris, DOE
11:05 - 11:30 a.m.	SERI Overview	C. Grosskreutz, SERI
11:30 - 12:00 noon	SERI Role	C. Smith, SERI
12:00 - 1:30 p.m.	LUNCH (on own)	

SESSION II: THREE CONCURRENT SESSIONS

Green Center

SESSION IIA: BIOMASS PRODUCTION

R. Inman, SERI
Chairman

1:30 - 1:45 p.m.	Biomass Production Overview	R. Inman, SERI
1:45 - 2:10 p.m.	Freshwater and Marine Plants for Biomass Production	J. Ryther, Woods Hole
2:10 - 2:35 p.m.	Bioengineering Aspects of Inorganic Carbon Supply to Mass Algal Cultures	J. Goldman, Woods Hole

2:35 - 3:00 p.m.	Biological Investigations of Marine Farms	W. North, Cal Tech
3:00 - 3:45 p.m.	COFFEE BREAK	
3:45 - 4:10 p.m.	Membrane Processes of Separation and Concentration in Biomass Production	H. Gregor, Columbia Univ
4:10 - 4:35 p.m.	Increasing the Biomass Production of Short Rotation Coppice Forests	K. Steinbeck, University of Georgia
4:35 - 5:00 p.m.	Fuels from Microalgae Biomass	J. Benemann, University of California/Berkele

SESSION IIB: BIOMASS REFINING
H. Bungay, Rensselaer
Chairman

1:30 - 1:50 p.m.	Fuels from Fermentation of Biomass	H. Bungay, Rensselaer
1:50 - 2:20 p.m.	Direct Microbiological Conversion of Biomass to Ethanol and Chemicals	D. Wang, MIT
2:20 - 2:40 p.m.	Production of Liquid Fuels from Cellulosic Biomass	E. K. Pye, University of Pennsylvania
2:40 - 3:00 p.m.	Selective Solvent Extraction in Utilization of Stored Solar Energy in Cellulosic Biomass	G. Tsao, Purdue Universi
3:00 - 3:20 p.m.	COFFEE BREAK	
3:20 - 3:40 p.m.	Process Development Studies on the Bioconversion of Cellulose and Production of Ethanol	H. Blanch, Lawrence Berkeley Laboratory
3:40 - 4:00 p.m.	Improvements in Cellulase Production by Trichoderma Through Mutation	B. Montenecourt, Rutger
4:00 - 4:20 p.m.	Acid Hydrolysis of Cellulosic Biomass	H. Grethlein, Dartmouth
4:20 - 4:40 p.m.	Design, Fabrication and Operation of a Biomass Fermentation Facility	D. O'Neil, Georgia Tech
4:40 - 5:00 p.m.	Liquid Fuels from Biomass	J. E. Sanderson, Dynate

SESSION IIC: THERMAL CHEMICAL CONVERSION
S. Ergun, Lawrence Berkeley Lab
Chairman

1:30 - 1:45 p.m.	Introductory Remarks and Overview of Biomass Liquefaction	S. Ergun, Lawrence Berkeley Lab
1:45 - 2:10 p.m.	Operation of the Biomass Liquefaction Facility	W. L. Berry, Jr., The Rust Engineering Co.
2:10 - 2:30 p.m.	Correlative Assessment of Process Development of Biomass Liquefaction	C. Figueroa, University of California
2:30 - 2:55 p.m.	Modified Extruder for Feeding Cellulosic Slurries to Pressure Systems	D. White, Universit Arizona

Session IIC (cont'd)

2:55 - 3:15 p.m.	Pretreatment of Biomass Prior to Liquefaction	L. L. Schaleger, Lawrence Berkeley Lab
3:15 - 3:30 p.m.	COFFEE BREAK	
3:30 - 3:55 p.m.	Bench Scale Research in Biomass Liquefaction in Support of the Albany, Oregon, Experimental Facility	D. C. Elliott, Battelle Pacific Northwest Labs
3:55 - 4:10 p.m.	Bench Scale Supporting Development for Biomass Liquefaction	M. Seth, University of California
4:10 - 4:35 p.m.	Thermochemical Conversion of Biomass to Gasoline	J. Diebold, Naval Weapons Center
4:35 - 4:55 p.m.	Economic Feasibility Assessments of Biomass Liquefaction	S. Ergun, University of California, LBL
4:55 - 5:15 p.m.	Summary of Current Status of Biomass Liquefaction and Schedule Activities	S. Ergun, University of California, LBL
6:30 - 7:30 p.m.	Social Hour (Cash Bar)	Freidhoff Hall
7:30	Banquet "The Burning Issue"	Speaker: Paul Bente, Bio-energy Council

Wednesday, June 6, 1979

SESSION III: TECHNICAL OVERVIEW* — Green Center
C. Smith, SERI
Chairman

9:00 - 9:45 a.m.	Biomass Processes Systems Evaluation and Financial Analysis	R. Katzen, Katzen Eng'g.
9:45 - 10:15 a.m.	Wood Energy Systems	D. Meadows, Dartmouth
10:15 - 10:30 a.m.	Technical Information Dissemination Program	W. Corcoran, SERI
10:30 - 11:00 a.m.	Preliminary Characterization of TASE Biomass Technologies	J. Harper, Argonne
11:00 - 11:30 a.m.	Energy Forest Treatment of Municipal Waste	G. Stanford, Greenhills Environmental Center
11:30 - 12:00 noon	Market Penetration Model	F. Schooley, SRI, Int'l.
12:00 - 1:30 p.m.	LUNCHEON	

*During this session there will be no set time for a coffee break, but coffee will be available all morning.

SESSION IV: THREE CONCURRENT SESSIONS

SESSION IVA: BIOMASS PRODUCTION

J. Ranney, Oak Ridge
Chairman

1:30 - 1:55 p.m.	Solar Hydrogen Production	B. Hoagland, SERI
1:55 - 2:20 p.m.	Analysis of the Photosynthesis Energy Factory	M. Fraser, InterTechnology Solar Corp.
2:20 - 2:45 p.m.	Potential Water Quality Impacts from Large Scale Crop Residue Harvesting	M. Torpy, Argonne
2:45 - 3:10 p.m.	Carbohydrate Crops As a Source of Fuels	E. Lipinsky, Battelle Columbus Labs
3:10 - 3:30 p.m.	COFFEE BREAK	
3:30 - 4:00 p.m.	Savannah River Energy Plan	A. Talib, MITRE
4:00 - 4:30 p.m.	Herbaceous Species Screening Program	K. Saterson, Arthur D. Little, Inc.

SESSION IVB: BIOMASS REFINING

L. Douglas, SERI
Chairman

1:30 - 1:50 p.m.	Hemicellulose Recovery and Pentose Utilization in a Biomass Processing Complex	R. Chambers, Auburn University
1:50 - 2:10 p.m.	Improvement of Yields and Rates Using Immobilized Enzymes during the Hydrolysis of Cellulose to Glucose	H. Klei, Connecticut University
2:10 - 2:40 p.m.	Bioconversion of Plant Biomass to Ethanol	R. Brooks, General Electric Company
2:40 - 3:00 p.m.	Enzymatic Saccharification of Waste Cellulose	M. Mandels, U.S. Army Natick Research Labs
3:00 - 3:20 p.m.	COFFEE BREAK	
3:20 - 3:40 p.m.	An Overview of Fuel Gas Production via Anaerobic Fermentation of Selected Municipal Wastes, Agricultural Residues, and Crop-Grown Biomass	D. Wise, Dynatech
3:40 - 4:00 p.m.	Experimental Anaerobic Fermentation Facility	D. Lizdas, Hamilton Standard
4:00 - 4:20 p.m.	Engineering Analysis of Anaerobic Digestor Concepts	E. Ashare, Dynatech
4:20 - 4:40 p.m.	Low Cost Methane Generator for Small Farms	W. Jewell, Cornell University

Session IVB (cont'd)

4:40 - 5:00 p.m.	Methane from Agricultural Residues Process Conversion Efficiencies	J. Pfeffer, University of Illinois

SESSION IVC: THERMAL CHEMICAL CONVERSION

G. F. Schiefelbein, Battelle Labs
Chairman

1:30 - 1:45 p.m.	Overview of SERI Thermochemical Conversion Program	T. Reed, SERI
1:45 - 2:10 p.m.	Biomass Energy Systems Program - An Overview of Thermochemical Conversion Activities	G. F. Schiefelbein, Battelle Labs
2:10 - 2:35 p.m.	Development of Wood as an Alternative Fuel for Large Power Generating Systems	J. Hamrick, Aerospace Research Corporation
2:35 - 3:00 p.m.	Advanced Systems Demonstration for Utilization of Biomass as an Energy Source	B. Milligan, Wheelabrator Cleanfuel Corporation
3:00 - 3:15 p.m.	COFFEE BREAK	
3:15 - 3:40 p.m.	Thermal Conversion of Biomass in a Rotary Kiln System	J. Turpin, University of Arkansas
3:40 - 4:05 p.m.	Synthetic Fuels from a Large Gasifier	V. Flanigan, University of Missouri-Rolla
4:05 - 4:30 p.m.	The SGFM Biomass Gasifier	S. Beck, Texas Tech Univ.
4:30 - 4:55 p.m.	Catalyzed Steam Gasification of Biomass	J. Coffman, Wright-Malta Corporation
4:55 - 5:30 p.m.	Catalytic Gasification	L. Mudge, Battelle Pacific Northwest Labs
7:00 - 9:00 p.m.	DISCUSSION GROUPS	Green Center
	Biomass Refining	H. Bungay, Rensselaer Chairman
	Anaerobic Digestion	D. Wise, Dynatech Chairman
	Biomass Production	R. Inman, SERI Chairman
	Thermochemical Conversion	T. Reed, SERI Chairman
	Biomass Utilization as a Fuel	Chairman

Thursday, June 7, 1979

SESSION V: THREE CONCURRENT SESSIONS

Green Center

SESSION VA: BIOMASS PRODUCTION

R. Inman, SERI
Chairman

9:00 - 9:50 a.m.	Research and Development of Intensively Cultured Plantations for Maximum Biomass Production	D. Dawson, USDA/ Forest Service
9:50 - 10:15 a.m.	To be announced	
10:15 - 10:30 a.m.	COFFEE BREAK	
10:30 - 10:55 a.m.	Energy and Chemicals from Woody Species in Florida	L. Conde, University of Florida
10:55 - 11:20 a.m.	Non-Commercial Woody Plants as Potential Biomass Fuel Producers: An Ecological Rational for Their Selection	H. Ragsdale, Emory University
11:20 - 12:00 noon	Species Selection and Silvicultural Systems for Producing Fuels from Woody Biomass in the Southeastern United States	D. Fredrick, No. Carolina State University
12:00 - 1:30 p.m.	LUNCH (on own)	

SESSION VB: BIOMASS REFINING

D. Jantzen, SERI
Chairman

9:00 - 9:25 a.m.	A.S.E.F. Solid Waste to Methane Gas	P. Ware, Waste Management
9:25 - 9:50 a.m.	Design and Evaluation of a Methane Gas System for a Hog Farm	W. Turnacliff, Bio-Gas of Colorado, Inc.
9:50 - 10:15 a.m.	Heat Treatment of Biomass for Increasing Biodegradability	P. McCarty, Stanford University
10:15 - 10:30 a.m.	COFFEE BREAK	
10:30 - 10:55 a.m.	Anaerobic Fermentation of Beef Cattle and Crop Residues	A. Hashimoto, U.S. Dept. of Agriculture
10:55 - 11:20 a.m.	The Operation of 183 m^3 Anaerobic Digestor at the Monroe State Dairy Farm	E. Coppinger, Ecotope Group
11:20 - 12:00 noon	Pilot Plant Demonstration of an Anaerobic, Fixed-Film Bioreactor for Wastewater Treatment	R. Genung, Oak Ridge National Lab
12:00 - 1:30 p.m.	LUNCH (on own)	

day, June 7, 1979 (cont'd.)

SESSION VC: THERMAL CHEMICAL CONVERSION

G. F. Schiefelbein, Battelle Labs
Chairman

9:00 - 9:25 a.m.	Conversion of Forest Residue to a Methane-Rich Gas	H. Feldmann, Battelle Columbus Labs
9:25 - 9:50 a.m.	Thermal Conversion of Biomass to Gaseous Products	D. Garrett, Garrett Energy R & D
9:50 - 10:15 a.m.	Thermochemical Catalytic Conversion of Biomass into Gas and Liquid Fuels	R. L. Garten, Catalytica Associates, Inc.
0:15 - 10:30 a.m.	COFFEE BREAK	
0:30 - 10:55 a.m.	Research and Evaluation of Biomass Resources/Conversion/Utilization Systems	R. Stringer, Gilbert Associates, Inc.
0:55 - 11:20 a.m.	Wood Residuals Fuel for Thermal Chemical Conversion	A. Nyce, Gorham International, Inc.
1:20 - 11:45 a.m.	Biomass Based Methanol Processes	E. Wan, Science Applications, Inc.
1:45 - 12:10 p.m.	Potential of Wood Gasification for Industrial Application	A. Talib, MITRE
2:10 - 1:30 p.m.	LUNCH (on own)	

SESSION VI: SUMMARY AND OPEN FORUM

S. Harris, DOE
Chairman

1:30 - 1:50 p.m.	Biomass Energy Systems: A Critical Economic Analysis	A. Kam, SRI Int'l.
1:50 - 2:10 p.m.	A Summary of an Economic Study of the Biological Production of Liquid Fuels	A. Humphrey, University of Pennsylvania
2:10 - 2:30 p.m.		E. Manual, Brookhaven National Lab
2:30 - 4:00 p.m.	Open Discussion/Summary	

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