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CONVERSION OF CELLULOSIC AND WASTE
POLYMER MATERIAL TO GASOLINE

By
James L. Kuester

March 28, 1979

Work Performed Under Contract No. EY-76-S-02-2982

College of Engineering and Applied Sciences
Arizona State University
Tempe, Arizona

U. S. DEPARTMENT OF ENERGY

Division of Industrial Energy Conservation

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James L. Kuester
Principal Investigator

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Interim Report to the
Alternate Materials Utilization Branch
Division of Industrial Energy Conservation
U. S. Department of Energy
Contract No. EY-76-S-02-2982.*000

March 28, 1979

College of Engineering and Applied Sciences
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SUMMARY

The present status and future plans for a project to convert cellulosic (biomass) and waste synthetic polymer materials to quality liquid fuels is presented. A thermal gasification approach is utilized followed by catalytic liquid fuels synthesis steps. Potential products include a medium quality substitute for natural gas or liquid fuel equivalents of diesel fuel, kerosene or high octane gasoline. The process appears very flexible with regard to ability to handle different sources of feedstock.

Results to date indicate quality products can be produced. Product yields need to be improved with the main thrust centered on improvement of pyrolysis gas composition. This will be a major effort in the new contract period. Other items to be addressed are study of alternate, economic feedstocks, waste stream characterization, and liquid fuels synthesis and tailoring with particular attention on the effects of alternate feedstocks.

A description of a proposed 10 ton/day pilot plant is presented with flow sheet, material balance and cost estimates.

DESCRIPTION OF PROCESS

The conversion system is research scale with a feedstock capacity of about 25 lbs/hr. The system is operated continuously and, for the most part, is a simulator of commercial scale processing. The system is capable of processing a variety of cellulosic and/or waste polymer feedstocks. Potential products include a medium quality gas, paraffinic fuels such as diesel or kerosene and/or a high octane gasoline. An additional potentially marketable product is an alcohol-water stream. The system is conveniently divided into two sections: (1) gasification, and (2) liquid fuels synthesis. The conversion scheme is depicted in Figure 1 and the processing equipment in Figure 2.

Gasification. The gasification system consists of a dual fluidized bed arrangement where one bed serves as the pyrolysis reactor and the other as a combustor to heat a solid medium continuously transferred between the two vessels. The advantage of the fluidized bed is efficient heat transfer. The dual system keeps the pyrolysis gas free of combustion products thus avoiding the necessity of extensive gas clean-up steps. In the pyrolysis reactor, feedstock is fed continuously via a feeding device into the bed operating at a controlled temperature in the 1000 - 2000 °F range and a slight positive pressure (0 - 5 psig). The feeding devices available include a vibrating screw conveyor and a rotary type valve. A pressurized hopper and sparge gas (pyrolysis gas) are used in conjunction with each device. The feedstock is flashed to a gas at these conditions. The gas is passed through an overhead system consisting of a cyclone separator (to remove char and other solid particles), scrubbing system (to cool the gas and remove tars, etc.), filter (to remove solid fines) and a compressor. From the compressor discharge, the gas is split to the pyrolysis reactor (fluidizing and sparge gas) and the liquid fuels synthesis system. Storage tanks are available to contain the generated pyrolysis gas and allow operation of the liquid fuels reactors independent of the gasification system.

The combustor on a larger scale would be fueled with air and recycle char plus off gases from the downstream reactors. In the research unit, propane and oxygen are used to feed the combustor. The solid heat transfer media can be inert or catalytic. The overhead system on the combustor is essentially the same as that for pyrolysis (cyclone, scrubber, filter).

Liquid Fuels Synthesis. The liquid fuels synthesis system consists of a modified Fischer-Tropsch catalytic reactor and a catalytic reformer. A prehydrogenation step is also under investigation between the Fischer-Tropsch and reforming reactors. The modified Fischer-Tropsch reactor is fed raw pyrolysis gas (primarily hydrogen, ethylene, carbon monoxide, methane, ethane and carbon dioxide). A cobalt-alumina catalyst is

BASIC CHEMICAL CONVERSION SCHEME

Figure 1.

CONVERSION SYSTEM SCHEMATIC

Figure 2.

utilized (developed on the project). The bed is fluidized (to achieve temperature control with significant exothermic heat of reaction) at a temperature of ca. 300°C and a pressure of ca. 125 psig. A resistance wire heater temperature control system is utilized. The gas product is cooled and two liquid phases are produced -- a hydrocarbon phase and a alcohol-water phase. Uncondensed gases with significant amounts of methane and ethane are available for recycle to the gasification system (but not implemented on the research scale system). The hydrocarbon/water phase ratio is about 10/1 on a volumetric basis. The water phase contains about 15% (wt.) low molecular alcohols (primarily n-propanol). The hydrocarbon phase is paraffinic in nature with some branched compounds, olefins and aromatics. This phase could probably be tailored to match paraffinic fuels such as kerosene and diesel. It is unsuitable however as an internal combustion engine fuel due to low octane rating. The required structural complexity (and thus high octane) is achieved in a fixed bed catalytic reformer using a commercial reforming catalyst. For this step, the reformer feed (Fischer-Tropsch hydrocarbons manually decanted from the water phase) is pumped to the reformer operating at a temperature of ca. 400°C and pressure of ca. 500 psig. Product gases are cooled and the high octane product collected in a trap. A resistance wire temperature control system is employed. Off gases are available for recycle (but not implemented).

EXPERIMENTAL FACILITIES

The basic conversion system consists of a thermal gasification (pyrolysis) system and liquid fuels synthesis system (Figure 3). All units are operated continuously. Auxiliary equipment includes size reduction equipment, extensive analytical instrumentation, control hardware and computation equipment.

Conversion System. The gasification system consists of a dual fluidized bed arrangement where one bed is operated as a gasifier and the other as a heat source for a solids medium continuously transferred between the two beds. Auxiliary gasification equipment includes cyclone separators, scrubbers, solids feed systems and a recirculation compressor.

The liquid fuels synthesis system converts the untreated pyrolysis gas to liquid fuels via catalytic chemical reactors. Auxiliary equipment includes condensers and liquid traps.

Analytical.

1. Hewlett-Packard Model 5781A Chromatograph. Gas or liquid feeds with hydrogen flame and thermal conductivity detector systems and temperature programmed oven.
2. Carle Model AGC-111 Chromatograph with thermal conductivity detector and automatic on-line sampling system.
3. Waters Associates Model ALC201/GPC501 Liquid Chromatograph with differential refractometer detector system and high pressure pumping system.
4. Varian Nuclear Magnetic Resonance T60A and XL-100-12. Multi nuclear capabilities: F^{19} , P^{31} , B^{11} , C^{13} . Fourier Transform interface, spin decoupling capability.
5. Atlas Mass Spectrometer CH4-B (medium resolution) and SM1-B (high resolution).

Control Systems. An extensive array of modular electronic control equipment is used to control the conversion system. The basic units include controllers, set stations, recorders, transmitters, transducers and control valves.

Computer Facilities. The conversion system and analytical equipment are tied to a real time computer (Digital Equipment Corp. PDP-11) for data logging and reduction purposes. The batch computers that are available for process modelling work are UNIVAC 1110 and IBM 370.

Other Facilities. Size reduction equipment (ball mills, pulverizers, screeners, etc.) is available for proper sizing of solid feedstock, heat transfer media and catalysts. In addition other equipment that normally would be located in a chemistry/chemical engineering department (ovens,

Figure 3. Conversion System.

wet chemistry apparatus, etc.) are available. An excellent development shop is available in the College of Engineering for fabrication of specialty items.

RESULTS TO DATE

The work on the project can be divided into two categories, (1) equipment design and development, and (2) factor and optimization studies.

Equipment Design and Development. Equipment modifications, improvements and automation have been in steady progress since the start of the contract (June, 1976). In many cases, commercial equipment is not available at this scale thus requiring local fabrication. This is true for essentially all of the processing vessels (reactors, cyclones, scrubbers, traps, etc). Other items (instrumentation, compressor, feeders, analyzers, etc) have been purchased. The gasification system is the most complex due to extensive solids handling (feeding, hot solids transfer, tar removal). No unsurmountable operational problems remain although appreciable adjustments are often required when exploring new conditions (new feedstocks, new operating condition ranges, etc). No attempt has been made to run the gasification system beyond about 8 hours. There would be little difficulty in maintaining the system at operating temperature for long period of times, i.e., a shutdown would normally be associated with feedstock conversion (e.g., feeding problems, tar removal problems, etc). Operationally, the liquid fuels synthesis reactors are quite reliable. The modified Fischer-Trapsch reactor has been operated continuously for about a 50 hour period. Some of the equipment would not be practical on a commercial scale however (resistance wire heaters, manual decanting). It is assumed that these steps would be improved at the next scale of development. An unplanned shutdown in these reactors would probably be due to a catalyst problem (low activity, carbon formation, etc).

Factor Studies and Optimization. The overall process objective is to maximize liquid fuel yields subject to constraints on product quality (composition, physical properties) and operating conditions (factors). The factors available for manipulation are those associated with any typical chemical reactor system -- temperature, pressure, residence time, feed composition, catalyst composition and characteristics. In addition, reactor mixing and solid contact patterns can play important roles in achieving desired process performance.

Several feedstocks have been explored including paper chips, Eco-Fuel II (Combustion Equipment Associates) and kelp waste. The paper chips and Eco-Fuel II appear to be similar in performance. The kelp waste (Kelco Co.) was dropped from consideration due to the large amount of inert filter aid (Perlite) present in the material.

Catalyst development initially was concentrated on the liquid fuels synthesis reactors. An extensive effort was made to develop a catalyst

to use in conjunction with the mixed pyrolysis gas feed stream. A modified Fischer-Tropsch type was finally selected (cobalt-alumina). Several commercial reforming catalysts were tested with material from American Cyanamid currently under study. The use of water gas shift and tar cracking catalysts in the pyrolysis reactor are currently under investigation.

Factor and optimization studies have followed the strategy depicted in Figure 4. A number of studies have been made in all three reaction systems (pyrolysis, Fischer-Tropsch and reformer). Results from three typical, recently completed experiments will be presented here. The factors studied, responses and design are as follows:

<u>Reactor</u>	<u>Factors</u>	<u>Responses</u>	<u>Design</u>
Pyrolysis	Temperature Feed rate	Gas phase yield Gas composition	2^2 factorial plus base pt.
Fischer-Tropsch	Feed composition (CO, H ₂ , C ₂ H ₄)	Liquid hydrocarbon yields	2^3 factorial plus base pt. and central composite pts.
Reforming	Temperature Pressure	Product yields Octane numbers	2^3 factorial plus base pt. and central composite pts.

For each system an experimental strategy diagram is presented with current progress indicated. Following this diagram is a diagram indicating the present capabilities of the system. The particular experimental design layout and results follows for each system. The significance of each experiment is as follows:

- 1) Pyrolysis - higher temperatures and feed rates favor gas phase yields and ethylene content (virtues). New compressor and 10" beds ordered to achieve this.
- 2) Fischer-Tropsch - need to produce more ethylene, less carbon monoxide and more hydrogen in feed gas from pyrolysis step to maximize liquid hydrocarbon yields. Water gas shift reaction and tar cracking catalyst being studied to achieve this (as well as higher temperatures and feed rates).
- 3) Reforming - in general, factors that increase yield tend to decrease octane number. Compromise region discovered that will produce high yields (> 80%) with acceptable octane (> 80). Prehydrogenation studies in progress for possible further improvement.

Figure 4. Experimental Strategy.

PYROLYSIS REACTOR

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Figure 5.

PYROLYSIS

Feedstock: Paper chips
Temperature, °C: 600-800
Pressure, psig: 0-5
Residence Time, sec: <2
Feed Rate, lb/hr: 10-20
Heat Transfer Media: Sand
Catalyst: None
Reactor Dimensions: 4" ID x 4'

Yield, lb.gas/lb.feedstock: .75-.85

Gas Composition, mole %:

H ₂	10-20
C ₂ H ₄	5-15
CO	35-50
CH ₄	10-15
C ₂ H ₆	1-5
CO ₂	15-30

Figure 6.

PYROLYSIS FACTORIAL EXPERIMENT

x_1 = reactor temperature, $^{\circ}\text{C}$

x_2 = feed rate, lbs/hr

Uncoded:	-1	0	+1
x_1	580	688	788
x_2	3.1	4.8	7.8

Figure 7.

Table 1.
PYROLYSIS GAS COMPOSITION
 (Mole %)

Run	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	O ₂	N ₂	CH ₄	CO
#1 Base	19.4	10.3	4.24	1.30	0	1.15	10.80	52.8
#2 Base	22.2	10.3	3.50	1.25	0	1.50	10.85	50.4
#3 Base	20.8	8.8	3.70	1.19	0	2.00	10.55	52.9
+1,+1	21.7	7.8	5.65	.84	0	.32	11.65	52.0
+1,-1	25.3	7.9	2.90	.40	.16	2.92	12.20	48.2
-1,+1	13.3	13.2	2.45	1.08	0	4.67	7.60	57.7
-1,-1	14.3	14.5	1.70	.95	.25	10.50	6.30	51.5

Table 2.
PYROLYSIS GAS YIELDS
 (lbs gas/lbs feed x 100)

Run	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO	TOTAL
#1 Base	1.11	12.9	3.38	1.11	4.91	42.0	65.4
#2 Base	1.42	14.6	3.15	1.21	5.57	45.3	71.2
#3 Base	1.15	10.8	2.86	.98	4.68	40.9	61.3
+1,+1	1.61	12.9	5.90	.94	6.90	54.2	82.4
+1,-1	1.36	9.4	2.21	.32	5.32	36.7	55.4
-1,+1	.50	10.9	1.28	.61	2.27	30.2	45.7
-1,-1	.78	17.1	1.27	.84	2.69	38.6	61.3

Figure 8.

FISCHER-TROPSCH

Feed Rate, SCFH: 14

Inlet Composition, mole %:

H ₂	43
C ₂ H ₄	21
CO	21
CH ₄	4
C ₂ H ₆	--
CO ₂	11

Temperature, °C: 255

Pressure, psig: 125

WHSV, g. feed/(hr)(g catalyst): 0.9

Catalyst: Cobalt-alumina

Reactor Dimensions: 2" ID x 6'

Hydrocarbon Liquid Yields, ml/hr: 132

Water-Alcohol Liquid Yields, ml/hr: 32

Hydrocarbon Phase Composition, wt. %:

Straight Chain Saturated	48
Straight Chain Unsaturated	8
Branched Chain	26
Cycloparaffins	1
Aromatics	9
Unknowns	8

Figure 9.

x_1 = mole% CO
 x_2 = mole% H₂
 x_3 = mole% C₂H₄

CODED:	-1.68	-1	0	+1	+1.68
UNCODED:					
x_1	2.28	7.14	14.28	21.42	26.28
x_2	23.71	28.57	35.71	42.85	47.71
x_3	2.28	7.14	14.28	21.42	26.28

Figure 10.

EXPERIMENTAL RESULTS (FISCHER TROPSCH)
(ml/hr liquid hydrocarbons)

19

At base point, 38, 44, 39, 41

Figure 11.

REFORMING REACTOR

20

Figure 12.

REFORMING

Inlet Composition, wt %:

Straight Chain Saturated	45-55
Straight Chain Unsaturated	5-10
Branched Chain	20-30
Cycloparaffins	0-2
Aromatics	4-10
Unknowns	--

Temperature, °C: 480

Pressure, psig: 500

WHSV, g. feed/(hr)(g catalyst): .16

Reactor Dimensions: 2" ID x 2'

Product Liquid Yields, vol.out/vol.in: .5-.8

Product Composition, wt %:

Straight Chain Saturated	15-25
Straight Chain Unsaturated	0
Branched Chain	19-23
Cycloparaffins	<1
Aromatics	40-50
Unknowns	--

Octane Number: 80-100

Specific Gravity: .7-.8

ASTM 10% (°F): 50-100

ASTM 90% (°F): 300-500

Figure 13.

x_1 = reactor temperature, °F

x_2 = reactor pressure, psig.

x_3 = liquid feed flow rate, ml/min

CODED:	-1.68	-1	0	+1	+1.68
UNCODED:					
x_1	808	825	850	875	892
x_2	416	450	500	550	584
x_3	0.16	0.5	1.0	1.5	1.84

Figure 14.

EXPERIMENTAL RESULTS (REFORMER)

(Yield/Octane No.)

23

At base point: 68/78, 68/79, 70/78

Figure 15.

A summary of the present status of the project is given below:

	<u>Pyrolysis</u>	<u>Fischer-Tropsch</u>	<u>Reforming</u>
	CO = 35-55 mole %	35 gals/ton cellulose	30 gals/ton cellulose
<u>Present:</u>	H ₂ = 10-20 C ₂ H ₄ = 5-10		
	CO = 21 mole %	100 gals/ton cellulose	>60 gals/ton cellulose
<u>Optimal:</u>	H ₂ = 28 C ₂ H ₄ = 21		

Thus the "optimal" feed gas composition to the modified Fischer-Tropsch step requires less carbon monoxide, more hydrogen and more ethylene. Also an improvement in reformer yield would be desirable. The major thrust (in progress) to achieve improved pyrolysis gas composition is the use of catalysts and redesign of the gasification reactors.

The dual fluidized bed gasification system is uniquely set up for the use of catalysts to manipulate the pyrolysis reactions. Current candidates are a tar cracking catalyst, water gas shift catalyst (to adjust the CO and H₂ content) and dolomite (to strip CO₂ in an exothermic reaction). Major developments are expected in these studies.

The original gasification fluidization system consisted of 4" ID reactors using 20-60 mesh heat transfer media particles. With this size particle, the temperature differential between the combustor and pyrolysis reactor was about 300°F. Thus the pyrolysis reactor temperature was limited to about 1500°F (since combustor temperatures above 1800°F can result in melting of metal parts). To reduce the temperature differential and allow for desired higher pyrolysis temperatures, a smaller solid particle is called for (~ 100-200 mesh). This should increase solid transfer rates and achieve improved fluidization patterns (with an expected improvement in gas composition and yields). To implement this concept at present feedstock rates, a larger diameter reactor and combustor is called for (to prevent solids blowover and maintain transfer loop surge capacity for decreased bed height). Ten inch diameter reactors are currently being installed for this purpose. A new, larger capacity compressor has also been purchased to handle the expected increase in gas phase yields.

Reformer yield improvements are centered on the following experiments: (1) optimization of operating variables (temperature, pressure, residence time) for reforming catalyst, (2) study of use of isomerization catalysts and (3) study of use of prehydrogenation catalysts.

CONTINUING RESEARCH

The projected work for the contract renewal period (June 1, 1979 to May 31, 1980) can be grouped into the following tasks: (1) feedstock studies, (2) gasification system optimization, (3) waste stream characterization, and (4) liquid fuels synthesis and tailoring.

Feedstock Studies. The study will be limited to industrial cellulosic wastes to include wood wastes, cotton gin trash and guayule bagasse. Paper chip feed will be used as appropriate to establish base conditions. An additional study will be to study the effect of degree of moisture in the feed on the process performance. It is anticipated that the main difficulty here will be feed control reliability problems with increasing amount of moisture. Finally, an attempt will be made to correlate the characteristics of the feedstock (chemical analysis etc.) with gasifier yields and composition.

Gasification System Optimization. New 10" fluidized beds (currently being installed) will be tested with the objective of allowing for investigation of higher reactor temperatures (1500 - 2000°F) and increased feed rates. From prior studies, the indication is that gas phase yields and composition will be improved in the new ranges of these two factors. In addition, studies on the use of catalysts in the pyrolysis step will be continued. This will include both water gas shift catalysts (with steam addition) and tar cracking catalyst. Both fixed bed (at the reactor overhead discharge) and fluidized bed (replacing inert heat transfer medium) modes will be explored. The objective is to reduce carbon monoxide in favor of hydrogen (water gas shift) and reduce tar formation in favor of gas phase yield improvement (tar cracking).

Waste Stream Characterization. A characterization and environmental assessment of potential waste streams will be conducted. The candidates are combustor off gas, scrubber liquid discharge and modified Fischer-Tropsch water phase trap discharge. The primary stream of concern is the pyrolysis reactor scrubber liquid discharge (tars, aldehydes, furfural, etc.). The thrust here will be to minimize the formation of potential wastes in the gasification step (via catalyst usage for example).

Liquid Fuels Synthesis and Tailoring. Three subtasks will be addressed: (1) catalyst activity, (2) diesel fuel production capability, and (3) high octane gasoline yield improvement. Catalyst activity is difficult to study without continuous, around the clock operation. Nevertheless, an attempt will be made to characterize the catalyst life and the effect of alternate feedstocks for both the modified Fischer-Tropsch and reforming steps. Of particular concern is the effect of

possible trace poisons in the feedstock (from pesticides, etc). A major emphasis will also be placed on producing a diesel type fuel product from the modified Fischer-Tropsch reactor. This will involve achieving control of both product structure and molecular weight. Past studies have shown that the product is very paraffinic in nature like diesel fuel but insensitive to changes in process conditions. Thus tailoring of the product may require some physical separation development work. Some analytical development work is also anticipated here (cetane number, etc.). Finally, an effort will be made to improve catalytic reformer yields. Past studies have shown that a volumetric yield drop of about 20% occurs to achieve commercial octane rating. Isomerization and prehydrogenation studies in progress will be continued in this regard.

A task schedule follows.

<u>Task</u>	0	2	4	6	8	10	12
1. feedstock studies							
2. gasification system optimization							
3. waste stream characterization							
4. liquid fuels synthesis and tailoring							
5. report							

PROPOSED PILOT PLANT AND PROJECTED ECONOMICS

The pilot unit at Arizona State University has demonstrated that the concept of producing liquid hydrocarbon fuels from cellulosic materials by flash pyrolysis is feasible. However, this unit is too small to permit obtaining certain engineering data which are required for designing commercial scale plants. To obtain such data it will be necessary to build a larger scale pilot plant. A 10 ton (feedstock)/day unit should be large enough for this purpose. Process flow diagrams (ANL 184-1 and ANL 184-2) as well as a material balance and capital cost estimate are included in this report. These are for a 10 ton/day plant with gasoline producing facilities.

The logical steps toward commercialization of this process are:

1. Design, construct, and operate a 10 ton/day pilot plant.
2. Design, construct, and operate a 300 ton/day demonstration plant.
3. Design, construct and operate a 1000 ton/day commercial plant.

The capacities indicated are in tons of bone dry feedstock. These plants may or may not include facilities for producing gasoline. Data obtained to date indicate that gasoline yield is too low and the final product should be a fuel oil (essentially no. 2 furnace oil). Proposed work at ASU may show that it is possible to raise the gasoline yield to acceptable levels. The 10 ton/day unit should be designed for future addition of gasoline facilities.

There is, of course, no economic justification for a 10 ton/day unit and a 300 ton/day unit is just above the breakeven point. The 1000 ton/day unit can be justified. An economic evaluation of such a unit is included.

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STREAM NUMBER	1	2	3	4
TEMPERATURE, °F	70	1400	1400	800
PRESSURE, PSIA	14.7	16	15	15
TOTAL FLOW RATE	833.3	5268.8	5241.5	27.3
COMPONENT				
H ₂				
N ₂		3625.7	3625.7	
O ₂				
CO				
CO ₂		1114.1	1114.1	
H ₂ O		501.8	501.8	
CH ₄				
C ₂ H ₄				
C ₂ H ₆				
C ₃ H ₈				
C ₄ H ₁₀				
C ₂ H ₅ OH				
1-PROPANOL				
2-PROPANOL				
CELLULOSE	833.3			
C ₅ -C ₁₉ N-ALKANES				
C ₅ -C ₁₉ BRANCHED				
ALKANES				
C ₅ -C ₁₀ ALKENES				
CYCLOPARAFFIN				
AROMATICS				
CONDENSABLES				
CHAR				
ASH		27.3		27.3
TAR				
C ₃ -C ₇ SAT.				
C ₃ -UP SAT.				
C ₃ -C ₇ BRANCHED				
PARAFFINS				
C ₇ -UP BRANCHED				
PARAFFINS				
UNKNOWN				

FLOW RATES ARE IN POUNDS PER HOUR

STREAM NUMBER	5	6	7	8
TEMPERATURE, °F	1100	800	700	70
PRESSURE, PSIA	16	15	15	14.7
TOTAL FLOW RATE				
COMPONENT	1595.9	1547.2	48.8	17635.5
H ₂	34.1	34.1		
N ₂				
O ₂				
CO	358	358		
CO ₂	657.1	657.1		
H ₂ O	51.3	51.3		17635.5
CH ₄	34.1	34.1		
C ₂ H ₄	358	358		
C ₂ H ₆				
C ₃ H ₈				
C ₄ H ₁₀				
C ₂ H ₅ OH				
1-PROPANOL				
2-PROPANOL				
CELLULOSE				
C ₅ -C ₁₉ N-ALKANES				
C ₅ -C ₁₉ BRANCHED				
ALKANES				
C ₅ -C ₁₀ ALKENES				
CYCLOPARAFFIN				
AROMATICS				
CONDENSABLES	54.6	54.6		
CHAR	48.8		48.8	
ASH				
TAR				
C ₃ -C ₇ SAT.				
C ₃ -UP SAT.				
C ₃ -C ₇ BRANCHED				
PARAFFINS				
C ₇ -UP BRANCHED				
PARAFFINS				
UNKNOWN				

STREAM NUMBER	13	14	15	16
TEMPERATURE, °F	100	80	536	85
PRESSURE, PSIA	50	250	147	40
TOTAL FLOW RATE				
COMPONENT	694.7	694.7	694.7	694.7
H ₂	16.3	16.3	3.1	3.1
N ₂				
O ₂				
CO	170.7	170.7	64.8	64.8
CO ₂	313.3	313.3	237.6	237.6
H ₂ O	7.6	7.6	50.2	50.2
CH ₄	16.3	16.3	37	37
C ₂ H ₄	170.7	170.7	21.6	21.6
C ₂ H ₆			92.6	92.6
C ₃ H ₈				
C ₄ H ₁₀				
C ₂ H ₅ OH			0.5	0.5
1-PROPANOL			11.8	11.8
2-PROPANOL			0.5	0.5
CELLULOSE				
C ₅ -C ₁₉ N-ALKANES			83.9	83.9
C ₅ -C ₁₉ BRANCHED ALKANES			14.9	14.9
C ₅ -C ₁₀ ALKENES			45.2	45.2
CYCLOPARAFFIN			2.4	2.4
AROMATICS			15.9	15.9
CONDENSABLES				
CHAR				
ASH				
TAR				
C ₃ -C ₇ SAT.				
C ₃ -UP SAT.				
C ₃ -C ₇ BRANCHED PARAFFINS				
C ₇ -UP BRANCHED PARAFFINS				
UNKNOWN			12.7	12.7

STREAM NUMBER	17	18	19	20
TEMPERATURE, °F	70	70	70	70
PRESSURE, PSIA	39.7	39.7	39	39
TOTAL FLOW RATE				
COMPONENT	456.7	238	174.7	63.3
H ₂	3.1			
N ₂				
O ₂				
CO	64.8			
CO ₂	237.6			
H ₂ O		50.2		50.2
CH ₄	37.0			
C ₂ H ₄	21.6			
C ₂ H ₆	92.6			
C ₃ H ₈				
C ₄ H ₁₀				
C ₂ H ₅ OH		0.5		0.5
1-PROPANOL		11.8		11.8
2-PROPANOL		0.5		0.5
CELLULOSE				
C ₅ -C ₁₉ N-ALKANES		83.9	83.9	
C ₅ -C ₁₉ BRANCHED ALKANES		14.9	14.9	
C ₅ -C ₁₀ ALKENES		45.2	45.2	
CYCLOPARAFFIN		2.4	2.4	
AROMATICS		15.9	14.9	
CONDENSABLES				
CHAR				
ASH				
TAR				
C ₃ -C ₇ SAT.				
C ₃ -UP SAT.				
C ₃ -C ₇ BRANCHED PARAFFINS				
C ₇ -UP BRANCHED PARAFFINS				
UNKNOWN		12.7	12.7	0.3

STREAM NUMBER	21	22	23	24
TEMPERATURE, °F	70	900	900	85
PRESSURE, PSIA	614.7	614.7	614	39.7
TOTAL FLOW RATE	174.7	174.7	174.7	174.7
COMPONENT				
H ₂			TRACE	TRACE
N ₂				
O ₂				
CO				
CO ₂				
H ₂ O				
CH ₄			23.6	23.6
C ₂ H ₄				
C ₂ H ₆			4.8	4.8
C ₃ H ₈			44.1	44.1
C ₄ H ₁₀			40.2	40.2
C ₂ H ₅ OH				
1-PROPANOL				
2-PROPANOL				
CELLULOSE				
C ₅ -C ₁₉ N-ALKANES	83.9	83.9		
C ₅ -C ₁₉ BRANCHED ALKANES	14.9	14.9		
C ₅ -C ₁₀ ALKENES	45.2	45.2		
CYCLOPARAFFIN	2.4	2.4	0.3	0.3
AROMATICS	15.9	15.9	11.5	11.5
CONDENSABLES				
CHAR				
ASH				
TAR				
C ₃ -C ₇ SAT.			14.7	14.7
C ₃ -UP SAT.			16.6	16.6
C ₃ -C ₇ BRANCHED PARAFFINS			11.1	11.1
C ₇ -UP BRANCHED PARAFFINS			3.5	3.5
UNKNOWN	12.4	12.4	4.3	4.3

STREAM NUMBER	25	26	27	28
TEMPERATURE, °F	85	85	75	70
PRESSURE, PSIA	39.7	39.7	39.7	39.7
TOTAL FLOW RATE				
COMPONENT	112.7	62	569.4	4727.1
H ₂	TRACE		3.1	
N ₂				3625.7
O ₂				1101.4
CO			64.8	
CO ₂			237.6	
H ₂ O				
CH ₄	23.6		60.6	
C ₂ H ₄			21.6	
C ₂ H ₆	4.8		97.4	
C ₃ H ₈	44.1		44.1	
C ₄ H ₁₀	40.2		40.2	
C ₂ H ₅ OH				
1-PROPANOL				
2-PROPANOL				
CELLULOSE				
C ₅ -C ₁₉ N-ALKANES				
C ₅ -C ₁₉ BRANCHED ALKANES				
C ₅ -C ₁₀ ALKENES				
CYCLOPARAFFIN		0.3		
AROMATICS		11.5		
CONDENSABLES				
CHAR				
ASH				
TAR				
C ₃ -C ₇ SAT.		14.7		
C ₃ -UP SAT.		16.6		
C ₃ -C ₇ BRANCHED PARAFFINS		11.1		
C ₇ -UP BRANCHED PARAFFINS		3.5		
UNKNOWN		4.3		

ESTIMATE SUMMARY
10 Ton/Day Flash Pyrolysis
Pilot Plant

	<u>Material</u>	<u>Labor</u>	<u>Total</u>
Vessels	344250	148050	492300
Mechanical	191050	82250	273300
Instruments	79800	34450	144250
Total Equipment	615100	264750	879850 ⁽¹⁾
Piping			175000
Structural			200000
Electrical			120000
Painting			20000
Total Direct Cost			1394850
Engineering ⁽²⁾			111600
Overhead ⁽³⁾			572400
TOTAL COST			2078850

(1) Includes foundations, supports, chutes, vents, insulation and installation

(2) Includes fee

(3) Includes ANL supervision

PRELIMINARY COST ESTIMATE
10 Ton/Day Flash Pyrolysis
Pilot Plant

ITEM	DESCRIPTION	MATERIAL	LABOR	TOTAL
A Vessels				
101	Combustor	130000	55900	185900
102	Pyrolyzer	130000	55900	185900
103	Combuster scrubber	3400	1450	4850
104	Pyrolyzer scrubber	3400	1450	4850
105	Fischer-Tropsch reactor	3500	1500	5000
106	Reformer	3000	1300	4300
201	Aftercooler	950	450	1400
202	F-T condenser	2500	1100	3600
203	Reformer interchanger	2000	850	2850
204	Reformer cooler	3500	1500	5000
501	Truck hopper	1000	450	1450
502	Storage silo	25000	10750	35750
503	Day Bin	7500	3200	10700
504	Ash receiver	3650	1550	5200
505	Separator	3650	1550	5200
506	Alcohol/water storage TK	8200	3500	11700
507	Intermed prod. storage TK	900	400	1300
508	Product storage TK	5900	2550	8450
509	Comb. scrubber decanter	3100	1350	4450
510	Pyrolyzer scrubber decanter	3100	1350	4450
TOTAL VESSELS		344250	148050	492300

ITEM	DESCRIPTION	MATERIAL	LABOR	TOTAL
B Mechanical				
430-1	Exhauster	2000	850	2850
430-2	Air compressor	8500	3650	12150
430-3	Instrument air comp.	4250	1850	6100
430-4	Low pressure comp.	10000	4300	14300
430-5	High pressure comp.	30000	12900	42900
430-6	Comb. fluidiz. blower	4500	1950	6450
431-1	Comb. scrubber pump	1000	450	1450
431-2	Pyrol. scrubber pump	1000	450	1450
431-3	Reformer feed pump	2200	950	3150
431-4	Loading pump	1500	650	2150
432-1	Storage conveyor	33000	14200	47200
432-2	Feed conveyor	22000	9450	31450
432-3	Star valve	3500	1500	5000
432-4	Star valve	2000	850	2850
432-5	Pyrolyzer feeder	6000	2600	8600
432-6	Char feeder	5000	2150	7150
432-7	Star valve	2000	850	2850
432-8	Star valve	2000	850	2850
435-1	Combustor cyclone	5000	2150	7150
435-2	Pyrolyzer cyclone	1500	650	2150
435-3	Gas filter	800	350	1150
435-4	Gas filter	800	350	1150
435-5	F-T cyclone	1500	650	2150
436-1	Liquid cyclone - Comb.	500	200	700
436-2	Liquid cyclone - Pyrol.	500	200	700
442-1	Instr. air dryer	4600	2000	6600
442-2	Hot oil system	23600	10150	33750
442-3	Reformer heater	11500	4950	16450
437-1	Dryer	300	150	450
	TOTAL MECHANICAL	191050	82250	273300

ITEM	DESCRIPTION	MATERIAL	LABOR	TOTAL
C INSTRUMENTS				
	G.C. recorder/integrator	5000	2150	7150
	G.C. recorder/integrator	1000	450	1450
	10-point temp. recorder	2500	1100	3600
	Temp. programmable G.C.	10000	4300	14300
	Isothermal G.C.	2500	1100	3600
	Controllers	20400	8800	29200
	DP & AP cells	9800	4200	14000
	TC transmitters	2100	900	3000
	Transducers	5100	2200	7300
	Control valves	12400	5350	17750
	3-pen recorders	9000	3900	12900
	TOTAL INSTRUMENTS	79800	34450	114250
	TOTAL EQUIPMENT	615100	264750	879850

An economic evaluation has been made for a 1000 ton/day commercial plant producing fuel oil only. The basis for this evaluation is:

1. Plant goes on stream in 1985 after 30 month design and construction, operated at 90% load-factor (328.5 days/year).
2. Process efficiency improves 25% (from 21 wt% to 26.25 wt%) from pilot plant and demonstration plant operation experience.
3. Product prices and costs escalate until plant goes on stream and then hold constant over life.
4. The large amount of feedstock needed is beyond waste availability. Feed must be gathered from forest scrap, deadwood, mill residue, etc. and be prepared for use, at a 1979 cost of \$10/ton.
5. An Investment Tax Credit was considered, at 10% of investment.

The discounted cash flow rate of return (DCF) was calculated at 4 levels of escalated product price: an estimated progressive rate and at constant 6%, 8% and 10%/yr rates until 1985. The resultant rates of return are 11.0%, 6.9%, 9.4% and 11.4% respectively.

A real project, privately owned and operated, would most likely use a form of accelerated depreciation for tax purposes. A 16 year-life, sum-of-the-years-digits methods was used and the results analyzed. As expected, the early decreases in taxable income and income tax create a larger cash flow during the first part of the project life, increasing DCF rates. The rates are (for the 4 cases mentioned above) 12.3%, 7.9%, 10.6% and 12.3% respectively.

Details of Investment, Operating and Maintenance Expenses, Feedstock Costs, Product Revenues and Basic Assumptions are attached.

COMMERCIAL PLANT
Light Fuel Oil Production From Waste Biomass
1000 TPD Plant, Start-up in 1985

ESTIMATED CAPITAL REQUIREMENT

Capital in \$10 ⁶ "Current" Calendar Year Year of Operations	1982.5 -2½ to -2	1983 -2 to -1	1984 -1 to 0	1985 0
Land	0			0
Facilities Price (1979)	2.85	13.55	11.90	28.30
Escalation Factor(1)	1.47	1.53	1.64	(1.57)
Installed Cost	4.19	20.73	19.53	44.45
Construction Interest (2)	0.23	1.60	3.82	5.65
Working Capital(3)	-	-	0.60	0.60
TOTAL CAPITAL REQUIRED	4.42	22.33	23.95	50.70

- (1) O&G J, Nelson Index-Ref'y Mat'ls (1977-62) at 7%, Ref'y Constr'n (1977-6) at 7.6%; are 7.30%/yr
 (2) At 11% on average balance
 (3) At 45 days ave. cash opg exp.

CASH FLOW FROM INVESTMENT PERIOD

Cash Expended	4.42	22.33	23.35	50.10
Less: Tax-loss, Int. Exp	(0.11)	(0.80)	(1.91)	(2.82)
Invest Tax Cred. 10%	(0.44)	(2.23)	(2.34)	(5.01)
Net Cash Expenditure	3.87	19.30	19.10	42.27
Add. Wkg Cap.	-	-	0.60	0.60
TOTAL CASH EXPENDED	3.87	19.30	19.70	42.87

ESTIMATED ANNUAL OPERATING & MAINTENANCE EXPENSES

Expenses in \$10³ (1985)

Payroll:	Manager	1 @ \$48 X 1.30	62
(O&GJ,	Engineer	1 @ \$36 X 1.30	47
1-1-79,	Operating Crews	4/shift @ 8hr/shift @ \$11.80/hr @ 986 shifts X 1.30	484
Refy Opg	Instrument Tech.	1 @ 32 X 1.30	42
Labor	Laborers	2 @ 48 X 1.30	62
7.25%/yr	Office	1 @ 14.6 X 1.3	19
escal'n	Maintenance	(2.25% Pit)	1000
	SUB-TOTAL		1716
Supplies:	Operating (15% O.L.)		73
	Maint. (2.25% Pit)		1000
	SUB-TOTAL		1073
Electricity (6.4%/yr)			389
Outside Services (1.073) ⁶ X60			95
Local Taxes/Insur - 3% Pit.(effective escl'n 10%/yr, actual 3%/yr)			1327
	TOTAL		\$4600

$$\text{Wkg Cap.} = 4600 \times \frac{45}{365} = 0.57, \text{ use } \$0.60 \times 10^6$$

COMMERCIAL PLANT
Light Fuel Oil from Waste Biomass

1. Estimated Feedstock Cost

"Waste" biomass composite - basically timber, dead-wood, non-commercial timber, mill-residue, etc. for escalation: Oil & Gas Journal, 1-1-79, lumber composite index, 1977-62, = 7.8%/yr

Then, \$10/T (1979) X (1.078)⁵ = \$14.55/TON - 1985

1000 TPD X 328.5 days/year X \$14.55/TON = \$4.78 X 10⁶/yr

2. PRODUCT REVENUES:

a. Price Escalation:

1. Estimated: 1979 - 15% (OPEC), 1980-12%, 1980 - 10%, 1982-4, 8%/yr
= 1.785 factor 1.785 X 42¢/gal 1979 = 75¢/Gal/1985

2. a. 6%/yr (very conservative) - mid-1985 = 61.3¢/Gal

b. 8%/yr (still somewhat conservative) = 69.3¢/Gal

c. 10%/yr (high) = 78.0¢/Gal

b. Total Revenues

A. 100 TPD feed X 328.5 $\frac{\text{days}}{\text{year}}$ X 0.2625 $\frac{\text{Prod}}{\text{Feed}}$ X $\frac{293 \text{ gal}}{\text{ton}}$ = Gal/yr

1. Gal/yr X \$0.75/Gal = \$18.95 X 10⁶/yr

2. a. X 0.613/Gal = \$15.49 X 10⁶/yr

b. X 0.693/Gal = \$17.51 X 10⁶/yr

c. X 0.780/Gal = \$19.71 X 10⁶/yr

COMMERCIAL PLANT
Economic Assessment

Conventional Comparison - Find DCF rate

Net Cash Investment (in \$10⁶ "current")(1)

<u>Year</u>	<u>Amount</u>
-2½ to -2	3.87
-2 to -1	19.30
-1 to -0	19.10
0	0.6

(1) Net Cash Expenditures assumes owner can utilize interest expenses during construction as tax-deduction from other income and can fully use 10% investment tax credit annually as is promulgated for multi-year coal gasification construction proposals.

Annual Operations: (in \$10⁶, 1985)

	<u>Est.</u>	<u>6%/yr</u>	<u>8%/yr</u>	<u>10%/yr</u>
Product price escalation				
Revenues	18.95	15.49	17.51	19.71
Less: O&M	4.60			
Feedstock	4.78			
Depreciation	2.50			
Taxable Income	7.07	3.61	5.63	7.83
Net Income	3.54	1.81	2.81	3.91
After Tax Cash Flow	6.04	4.31	5.31	6.41

DCF Equation:

$$-(3.87(1+i)^{2.25} + 19.30(1+i)^{1.5} + 19.1(1+i)^{0.5} + 0.6) + ATCF_k (P/A)_{20}^i = 0$$

DCF Rate	11.0%	6.9%	9.4%	11.8%
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COMMERCIAL PLANT
Additional DCF rate
Use of Accelerated Depreciation

Sum of the years digits depreciation (SOYD) 16 year life.

1. Any given year - "k" where n = depreciable life

$$\text{Depreciation rate} = \frac{.2(n-k)}{(n)(n+1)}$$

2. Uniform Annual Equivalent over life: (UAE)

$$\text{UAE SOYD} = \frac{2}{(n)(n+1)} \left[n(P/A)_n^i - (P/G)_n^i \right] (A/P)_{20}^i *$$

n = 16:

$$\text{Then UAE SOYD} = 0.00735 \left[16(P/A)_{16}^i - (P/G)_{16}^i \right] (A/P)_{20}^i$$

3. With depreciable plant = \$50.10 x 10⁶

$$\text{UAE SOYD} = \$0.368235 \times 10^6 \left[16(P/A)_{16}^i - (P/G)_{16}^i \right] (A/P)_{20}^i$$

4. For 4 cases (in \$10⁶)

Product price escl'n	Est.	6%/yr	8%/yr	10%/yr
Operating Margin	9.57 **	6.11	8.13	10.33
Less: Depr VAE	0.368 x []			
Taxable Income	9.57 - 0.368 []	6.11 - 0.368 []	8.13 - 0.368 []	10.33 - 0.368 []
Income Tax	4.785 - 0.184 []	3.055 - 0.184 []	4.065 - 0.184 []	5.165 - 0.184 []
Net Income	" - " "	" - " "	" - " "	" - " "
ATCF	4.785 + 0.184 []	3.055 + 0.184 []	4.065 + 0.184 []	5.165 + 0.184 []
DCF rate	12.3%	7.9%	10.6%	13.2%

$$[] = \left[16(P/A)_{16}^i - (P/G)_{16}^i \right] (A/P)_{20}^i$$

* $(P/A)_n^i$ = Present value of uniform annual series at rate "i" over "n" years

$(P/G)_n^i$ = Present value of uniform annual arithmetical gradient at rate "i" over "n" years

$(A/P)_n^i$ = Uniform annual equivalent series from a present amount at rate "i" over "n" years

Further Study:

- Assumptions:
- 1) Commercial Plant (100 TPD) $\$28.3 \times 10^6$ (1979)
 - 2) Operations: Start 1985
 - 3) On-stream 90% - 328.5 days/yr
 - 4) Efficiency improves from 21 wt-% prod. to 26.25 wt-% (25% of present) through pilot plant, demo. plant process improvement
 - 5) Tipping Fee: The size of commercial plant means great effort to collect, gather and prepare feedstock - it will be a cost (1979) of \$10 instead of revenues.
 - 6) Assume inflation rates for all costs, intervening years, i.e. - plant invest., O&M product
 - 7) Derive ATCF using allowable accelerated deprec., Investment Tax Credit (10%), and any tax-loss carry forward or back provisions (+7, -3)
 - 8) ITC on multiyear proj. allowed annually (coal gasif.), parent takes,