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ANAEROBIC FERMENTATION OF AGRICULTURAL RESIDUE: POTENTIAL FOR IMPROVEMENT AND IMPLEMENTATION

FINAL REPORT

February 1978

**Prepared for
U.S. DEPARTMENT OF ENERGY
Assistant Secretary for Energy Technology
Division of Solar Technology**

Under Grant No. EY-76-S-02-2981

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ANAEROBIC FERMENTATION OF AGRICULTURAL RESIDUE: POTENTIAL FOR IMPROVEMENT AND IMPLEMENTATION

ABSTRACT

This report contains the results of one year of activity by a multi-disciplinary team at Cornell University that attempted to evaluate the potential of rapidly improving the technology of anaerobic fermentation of agricultural residues and methods of implementing it in existing agricultural operations. The main objectives of this study were to: identify simple and low cost anaerobic fermentor design criteria that would be appropriate in small agricultural operations, develop high rate fermentor concepts that would enable multiple product recovery from the reactor, expand the information base particularly in the area of temperature influence on the process, and to review sociological and economic issues relating to implementation of fermentation technology.

This study has identified several major anaerobic fermentation concepts which illustrate that the technology may be rapidly improved. A simple reactor design utilizing an unmixed plug flow concept was shown to be comparable to the more complex completely mixed reactor when using dairy cow residue. This design should enable the use of low cost reactor materials for small scale agriculture. A high rate thermophilic reactor designed to encourage flotation of particulate solids illustrated that liquid, solid, and gaseous products can be generated within the anaerobic fermentor thus eliminating an additional dewatering unit process.

A third reactor concept involved extension of the anaerobic attached microbial film expanded bed to the treatment of cow manure slurries. By concentrating the viable microbial mass in an attached film on small inert particles it was possible to operate this unit at much higher flow rates than would be possible with conventional processes. At a three hour hydraulic retention time, 35°C, with 2 percent total solids feed, the reactor was operating in a stable condition producing gas at a rate of 7 l/l-day with effluent volatile acids below 300 mg/l as acetic acid. This is the highest rate of methane generation recorded for this type of substrate to date.

Comprehensive thermophilic fermentation studies (60°C) indicated that the increased temperature resulted in little improvement in total quantity or the rate of yield of gas over that obtained with mesophilic fermentation with reactor retention periods greater than 10 days. A significant improvement was noted for reactor retention periods less than 5 days with feed total solids greater than 6 percent of the wet weight.

Finally, other areas where preliminary data was obtained included: surveys to identify sociological issues, computer model economic comparison of the cost of anaerobic fermentation as compared to conventional residue treatment technology, microbiological examination of reactors operated, and nutrient analysis of dairy manure before and after anaerobic fermentation.

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LIST OF ABBREVIATIONS

AFEB	- attached film expanded bed reactor
A/S	- air to solids ratio (mass/mass)
BA	- bicarbonate alkalinity (mass/volume)
BOD ₅	- biochemical oxygen demand, 5-day (mass/volume)
BVS	- biodegradable volatile solids
BY	- bacterial yield (weight) per weight of volatile solids destroyed
COD	- chemical oxygen demand (mass/volume)
CST	- capillary suction time
CSTR	- completely stirred tank reactor
DAF	- dissolved air flotation
DCW	- digestion of the cell wall (%)
E _b	- efficiency of biodegradable volatile solids removal (%)
FA	- free ammonia (mass/volume)
FRT	- float residence time
gpm	- gallons per minute
G/S	- gas to solids ratio (mass/volume)
G _b	- methane production, based on biodegradable volatile solids (BVS) concentration change (volume CH ₄ /volume reactor/day)
gm	- gram
G _t	- methane production, based on total solids (TS) concentration change (volume CH ₄ /volume reactor/day)
HRT	- hydraulic retention time

LIST OF ABBREVIATIONS (CONTINUED)

k	- substrate removal coefficient
kcal	- kilocalorie
kg	- kilogram
K_s	- half velocity coefficient (mass/volume)
kWh	- kilowatt-hour
ℓ	- liter
L_k/ADR or L/ADR	- lignin to acid detergent residue ratio
m	- meter
mg	- milligram
mℓ	- milliliter
MMBtu	- million Btu
MPN	- most probable number
P	- weight of protein synthesized
PB	- weight of protein per weight of bacteria
PD	- fraction of volatile solids degraded
Q	- flow rate (volume/time)
R	- refractory fraction
S_1	- effluent total volatile solids (mass/volume)
S_{b1}	- effluent biodegradable volatile solids (mass/volume)
S_{r1}	- effluent refractory volatile solids (mass/volume)
S_0	- influent total volatile solids (mass/volume)
S_{b0}	- influent biodegradable volatile solids (mass/volume)
S_{r0}	- influent refractory volatile solids (mass/volume)

LIST OF ABBREVIATIONS (CONTINUED)

SRT	- solids retention time
STP	- standard temperature and pressure (1 ATM, 0°C)
t	- time
TA	- total alkalinity (mass/volume)
TDDM	- total digestible dry matter (%)
TDOM	- true digestibility of organic matter (%)
TKN	- total kjeldahl nitrogen (mass/volume)
TS	- total solids
TVA	- total volatile acids (mass/volume)
TVS	- total volatile solids (mass/volume)
μ	- specific growth rate of microorganisms (time^{-1})
$\bar{\mu}$	- maximum specific growth rate of microorganisms (time^{-1})
UHD	- unmixed horizontal displacement reactor
V	- volume
VS	- volatile solids
VS_A	- volatile solids added
VS_D	- volatile solids destroyed

SYSTEMS ANALYSIS
OBJECTIVE FUNCTION VARIABLES
(CHAPTER VIII)

CC	- labor cost per unit
CD	- cost per unit of diesel fuel
CE	- cost of a unit of electricity
CK	- potassium unit value
CM	- cost per unit of methane
CN	- unit value of nitrogen
CP	- value per unit of phosphorus
C_x	- units of annual labor for unit x
D_x	- units of diesel fuel for system x per year
E_x	- annual units of electricity for system x
FC_x	- fixed annual cost of system x
K_x	- number of units of potassium for system x per year
M_x	- annual production of methane for system x
$NASC_x$	- net annual system cost of system x
N_x	- annual production of nitrogen for system x
P_x	- phosphorus produced by system x per year

ACKNOWLEDGEMENTS

This final report was developed at the request of the U.S. Department of Energy's Project Manager, Dr. Roscoe F. Ward. Much of the data and analyses included are of a preliminary nature since this report was originally scheduled for completion in June, 1979 according to the original proposal for this study. The unexpected change in emphasis resulted in premature termination of a large number of experiments being conducted on a multi-year time schedule.

Presently, field scale development of the simplified reactor discussed in this report is under study at Cornell University.

Major authors for each section are noted at the end of each chapter. Since the major input to this project was in the form of manpower, the following distribution of manpower support is indicative of the amount of effort expended in various areas (values in percent of total manpower contributed to the project):

1. Plug flow concept - 30 (bench-scale - 10, pilot-scale - 15, semi-solid fermentor - 5);
2. High rate liquid solid separation - 27 (thermophilic internal separator - 24, expanded bed - 3);
3. Integrated topics - 33 (microbiological studies - 8, information integration and transfer - 14, social and economic impact - 6);
4. Conventional and basic information - 10.

The authors wish to thank J. D. Krizek for drafting the figures and T. J. Sampson for assisting in preparing and typing the manuscript.

SUMMARY AND CONCLUSIONS

Anaerobic fermentation of organic residues is a unique technology among those presently available that can produce significant quantities of clean renewable energy at a price that may be competitive with those for existing gaseous fuels. This study illustrated that there is great potential for rapid improvement and implementation of this method of processing organic residues. A number of unique developments reported here illustrate two major areas of application that show promise for future development. First, low cost anaerobic fermentors for small scale farming operations may be technically feasible and the operation may be simpler than would be anticipated based on existing literature on sewage sludge fermentation. Presently this phase is being expanded from the pilot-scale laboratory work to large field-scale experimentation at Cornell University. Secondly, high rate anaerobic fermentor designs capable of multi-product recovery have been tested and shown to be feasible.

STUDY RATIONAL AND APPROACH

The conversion of organics to methane and carbon dioxide via anaerobic fermentation is an attractive technology for a number of reasons:

- There are significant quantities of organic residues which are presently being wasted and which represent a potentially significant energy supply;
- application of the technology assists in controlling pollution such as odor and nutrient control;
- comparison of the quantity of energy that can be generated from organic residues in existing farming operations indicates that a large fraction of the petroleum fuels used for food production

would be replaced by methane. Thus the possibility of creating energy self-sufficient food production centers appears to be achievable;

- anaerobic fermentation results in the development of useful by-products such as single cell protein as well as a high quality automatically separated form of energy;
- although the near term application of this technology appears to be limited to existing animal manures and crop residues, extension of this technology into energy farming of the land and the oceans holds promise of supplying large quantities of clean renewable fuels and other food products.

An earlier comprehensive study reported on the technical, practical and economic feasibility of using anaerobic fermentation on small dairies (40 and 100 cows) and medium sized beef feedlots (1000 cows) (Jewell et al., 1976). The net cost of methane generated on small operations was two to four times the existing cost of natural gas, but economies of scale decreased the cost for the larger operations to less than the cost of natural gas. Several significant qualifiers were added to these conclusions. The cost of converting the mixture of methane and carbon dioxide (often referred to as biogas) and using it in a conventional manner greatly increased the cost of the energy. On the positive side, it was noted that few improvements had been made in the technology since the mid 1930's, and that additional benefits might be derived from the technology such as pollution control or by-product recovery. The study reported here attempted to provide an indication of the improvement that could be achieved with emphasis on the fermentation process itself and the products that could be derived from fermented dairy cow manure.

An overall summary of transformations that occurs with various organic residues is illustrated in Figure 1 and the potential value of the by-products is shown in Figure 2. These two figures illustrate some of the applications and some of the problems addressed during this study. For example, for a ton of wet sewage sludge there is very little excess energy provided because of the large quantity of water in relation to the organic matter which is biodegradable. If it is assumed that a dilute slurry (7 percent dry solids)

Figure 1. Mass and energy transformations in the anaerobic fermentation of municipal and agricultural residues: sewage sludge, solid waste, crop residues, and dairy cow manure.

Figure 2. Hypothetical by-products developed during anaerobic fermentation of freshly produced animal manures.

must be fed to the reactors, a large quantity of water must be added to dry organics such as crop residues. This limitation raises the question of the reason for the need for dilute organics to support the fermentation process. After fermentation is complete there are significant by-products available for recovery, reuse, or disposal. Figure 2 indicates that for a wet ton of animal wastes processed by anaerobic fermentation, a total of about \$5.00 of energy is generated, but the total value of all of the by-products is more than three times the energy value. Thus the central questions regarding the adoption of the technology relate to the kind of reactor(s) that are affordable and can be operated on specific types of operations and the efficiencies of the conversion and by-product separation processes. A small dairy with a total residue product flow to the fermentor of 0.5 tons (dry) per day may need a different type system than a beef feedlot with 50,000 head of beef that generates a dry residue at the rate of more than 100 tons per day. This study has defined the potential of improving anaerobic fermentation as it would apply to agricultural residues in four general areas:

- development of a simplified low cost reactor design for small scale operations,
- identification of high rate multiple by-product recovery reactors,
- expansion of basic information, and
- review of the capability of integrating anaerobic fermentation into existing agricultural operations.

SIMPLIFIED LOW COST REACTOR DESIGN - PLUG FLOW CONCEPT

The simplest design concept that could be developed would be an unmixed unit with the flow controlled by gravity capable of treating concentrated organics. A low cost reactor would also require the use of soil-supported structures, and inexpensive materials that could be erected by the farmer at the organic generation site. One adaptation of these types of restrictions would be to use a soil trench lined with flexible lagoon liner material, covered with this same type of material and fed at one end. No intense mixing would be utilized and flow through the unit would be achieved by gravity. Although such design criteria has been shown to create fermentation problems

with residues such as sewage sludge little was known regarding the restrictions of this design when applied to uniform materials free of toxic constituents such as agricultural residues. This design concept, called the plug flow design for convenience, was the focus of small scale laboratory studies at temperatures varying from 20°C to 60°C, and a large scale 5 cow pilot plant.

Laboratory Simulation of Plug Flow Reactors

Two types of reactors were used to simulate the plug flow conditions in the laboratory. One was constructed of five foot long six inch diameter tubes, and the second was four completely mixed reactors in series. The long continuous tubes were found to be unsuitable for small scale studies with highly fibrous substrate such as dairy manure.

In long term biodegradability studies, the maximum portion of volatile solids found to be converted to methane and carbon dioxide was 44 to 51 percent of the total. This resulted in a maximum potential gas yield of about 7 cubic feet of gas producible per pound of organic solids added to the reactor or about 3.5 ft³ of methane per pound of solids fed at 35°C.

In all cases with the small scale reactors and the pilot plant there were no serious start-up problems. Although there was a slight depression in the pH as the acid formers developed initially faster than the methane formers, the buffering capacity of the dairy manure was sufficient to maintain it above 6.5 in all cases.

Gas production rates in the 20 and 30-day HRT plug flow tube reactors at 35°C averaged 1.3 and 1.1 volume of gas per volume of reactor per day, or 5.4 and 6.5 ft³ per lb TS fed, respectively; indicating efficient conversion of the available organic carbon.

The plug flow simulators consisting of four completely mixed reactors connected in series were used to develop data on the impact of high feed rates since this would be the critical condition in the first zone of a plug flow reactor. The following data are representative of the small scale series

studies fed a substrate containing 80 gm/l total solids.

Temp., °C	Total Retention Period, Days	Series Reactor HRT Days	Solids Destruction Efficiency % VS	Gas Production Rate, vol/total Reactor vol.-day
22°C	12	3	14	0.27
35°C	10 (40 % recycle)	2.5	29	1.32
35°C	4 (20 % recycle)	1.0	29.1	1.7
60°C	12	3	-	1.6

Even at high feed rates with a one-day HRT (with recycle) the reactors appeared to be operating in a stable condition with volatile acid concentrations of 3200 mg/l as acetic acid.

Pilot Plant Plug Flow Reactor

The 5600 liter plug flow pilot plant reactor was fed solids at 7.8 percent for a 30-day HRT at 35°C for 150 days, and then the feed solids were increased to the "as-produced" concentration which varied between 10 and 12 percent TS. The volatile solids reduction efficiency was 32 percent and 25 percent for the 8 percent and 12 percent feed conditions, respectively. Gas production averaged 0.9 vol/vol-day for the 8 percent solids feed condition and 1.0 vol/vol-day for the 10 to 12 percent solids feed. The methane fraction in all samples varied between 55 and 61 percent, with higher percentages resulting at the longer retention periods. Although floating fiber accumulated at the 8 percent total solids feed, no float formed at the higher solids concentration. Thus the larger scale pilot plant was successfully operated without mechanical mixing and with undiluted dairy cow residues. This design information opens the possibility of implementing low cost reactors in many agricultural operations. This design concept is presently being tested with a 65 cow capacity unit.

Anaerobic Fermentation of Semi-Solid Materials

As noted in Figure 1 a large amount of liquid needs to be added to many organic residues in order to obtain a liquid slurry solids concentration that is recommended for fermentation. Thus the availability of anaerobic fermentors capable of producing methane at solids concentrations significantly higher than 10 percent would expand the application of this technology.

In order to investigate the influence of moisture content on the fermentation process, mixtures of dairy cow manure and straw with dry solids concentrations ranging from 5 to 35 percent were examined in laboratory reactors. At solids retention times less than 20 days and with solids concentration greater than 25 percent typical reactor failure occurred with depressed pH values and low methane production. The gas production rate at 22 percent solids was 1.6 ℓ/ℓ -day (52 percent methane) at 20-day SRT and 35°C. Preliminary conclusions thus indicate that animal manure and bedding combinations up to 25 percent solids can be successfully fermented to methane and carbon dioxide. Fermentation at higher solids appears to be inhibited. Additional studies are being conducted on this topic.

HIGH RATE MULTIPLE BY-PRODUCT RECOVERY REACTOR CONCEPTS

Thermophilic Internal Liquid-Solid Separation

Earlier investigations indicated that efficient liquid-solid separation would be difficult to achieve with anaerobically fermented animal residue slurries. One concept which was hypothesized was to utilize the gas production to float the fiber to the surface and to achieve internal liquid-solid separation. Operation of this unit at high temperatures should favor this reaction since the viscosities of the slurry would be decreased and gas production increased. A 100 liter, one foot diameter, 9 foot high laboratory reactor was constructed to simulate the flotation reaction. All studies were conducted at 50°C.

In general this reactor concept was shown to be capable of efficient separation of gaseous, liquid and particulate phases. Using a 5 percent total solids feed and a 5-day HRT, a float solid of 14 percent was achieved, and the underflow solids concentration varied from 1 to 3 percent. The highest gas production that occurred was 2.4 ℓ/ℓ -day with 56 percent methane content.

Anaerobic Attached Film Expanded Bed Reactor

There are a number alternatives that have been examined to increase the rate of through-put in an anaerobic reactor. One means would be to separate the active bacteria from the flow and maintain them within the reactor. Until recently the tendency for microbes to form films and remain separate from the food supply passing them has only been studied in applications such as the trickling filter. This is the first study to examine the reaction of an attached film to a particulate solids feed of dilute animal residues.

The attached film expanded bed process consists of an upflow filter composed of small inert particles that become naturally coated with a film of microorganisms. The unit is usually operated in an upflow mode to slightly expand the bed and distribute the flow around the media.

A 8.2 liter reactor was tested with diluted dairy cow manure at 1 and 2 percent TS. This unit was operated for nearly 10 months under five loading conditions with HRT varying from 1.2 to 0.15 days. The organic solids loading rate varied from 8.3 to 133 kg TS/m³/day (0.5 to 8.3 lb TS/ft³/day).

Total volatile solids destruction efficiencies of 40 percent was achieved at an HRT of 1.2 days using a 2 percent TS in the attached microbial film reactor, whereas equal efficiencies were achieved at a 10-day HRT in a control reactor that was a conventional completely mixed unit. This process was found to be stable and capable of producing gas at a rate of 7 vol/vol-day at an HRT of 0.15 days with no indication that the unit was being stressed. At a loading rate of 133 kg TS/m³/day (an HRT of about 3 hours) the effluent volatile acids were 280 mg/ ℓ as acetic acid. These data represent the highest rate of methane production from a resistant organic such as cow manure that has been observed.

The data developed in this section of the study clearly emphasizes the potential of this technology. Much additional study will be required to define the limitations of the attached microbial film reactors as well as to define the engineering design specifications.

FUNDAMENTAL STUDIES OF ANAEROBIC FERMENTATION

In comparison to sewage sludge, relatively few studies have been conducted on the anaerobic fermentation of agricultural residues. Because of the lack of information on such materials as dairy manure, there is a tendency to extrapolate results of studies on substrates such as sewage sludge or municipal solid wastes to other substrates. Such extrapolations are possible only when the existing experimental data has been carefully defined with fermentor controlling parameters identified. Unfortunately, there remain significant questions regarding the impact of substrate composition, viscosity, nutrients, toxic materials, microbiology, and other characteristics so that such extrapolations should be accepted as approximations. One of the most important questions relates to the impact of increased temperature on methane production. Previous studies reported increased methane yields and higher rates at thermophilic conditions. For this reason a series of fermentation experiments were conducted with dairy cow manure at 60°C.

Long term biodegradability studies indicated that the biodegradability fraction varied from 50 to 53 percent of the feed volatile solids at 60°C. An average biodegradable value of 51 percent of the volatile solids was used in the kinetic analyses in this study. This biodegradability fraction is similar to that observed earlier with mesophilic studies.

All semi-continuously fed thermophilic reactors operated well at feed concentrations varying from 4 to 10 percent in reactors with HRT's varying from 5 to 30 days. Total volatile solid destruction efficiencies were lower than expected varying from 19.1 to 33.1 percent and gas production rates varied from 0.75 to 3.3 ℓ/ℓ -day.

The units operated with 8 and 10 percent feed solids concentrations resulted in fairly stable single stage completely mixed reactor operation down to an HRT of 2.5 days, whereas reactors fed 4 and 6 percent solids failed at 2.5-days HRT. Gas production rates approached 4 volume per reactor volume per day with 2.5-day HRT experiments. Volatile solid destruction efficiencies decreased to 22 percent at these shorter retention times.

The empirical, first order reaction rate design model was found to be suitable in describing the data generated in the thermophilic study.

These results do not show large improvements in the quantity of methane yield or the rate of yield. At 10-days HRT or more, mesophilic fermentation exhibits an equal or greater volatile solids destruction efficiency with dairy cow manure. The most significant advantage of the higher temperatures would result with high feed solids and the low HRT of 2.5 days. Mesophilic units would have difficulty operating under these conditions.

Other fundamental areas discussed in the text include microbiological analyses, nutrient analyses of the fermentor effluents, and a review of kinetic modeling.

INTEGRATION INTO EXISTING AGRICULTURAL OPERATIONS

Economics of Adoption

General acceptance and rapid implementation of energy producing fermentation technology will depend largely upon the economic feasibility of integrating these systems into existing agricultural residue management schemes. Evaluation of the many manure handling alternatives, their economic evaluation, and comparison to addition of the fermentation system is beyond the scope of this study. However, an existing computer program was adapted to evaluate the economics to provide some insight into the competitiveness of the technology with existing manure management systems.

Net annual costs for simple and commonly used residue handling systems on a 100 cow dairy varied from around \$1600 for a simple tractor scraper-solid manure spreader system to nearly \$5700 for a gutter cleaner-solid manure spreader combination. Incorporation of conventional treatment systems such as an anaerobic lagoon increases the annual cost to as much as \$8500. In many instances the credits obtained from incorporating an anaerobic fermentor nearly completely offset the waste handling and/or treatment costs.

This analysis indicated that the energy value and potential income from using anaerobic fermentation holds promise even for small agricultural operations.

Social Considerations

The adaptation of new technologies to existing social and economic systems is a complex process. Even in the most seemingly straight forward cases, adjustments cannot be made quickly, and the role of any one set of technical experts is rather limited. A preliminary survey was conducted to identify the social issues surrounding the practical feasibility of incorporating methane production from dairy and beef feedlot residues.

One of the most important observations made during the interviews of dairymen in Vermont and New York is that a major change is now in progress in residue handling in the dairy industry. Most farmers are generally aware of the various alternative systems available or in testing stages, and many have plans underway to convert their own systems. However, many are also looking forward to the improvement of new developments such as methane generation as possible low cost residue handling/disposal systems.

In general, the farmers interviewed were aware of methane fermentation technological, most knew the state of development of the technology, and the economics were not the primary concern. If these systems can be demonstrated to be an improvement in the residue handling problems of small to medium sized agricultural operations, there would appear to be a significant market readily available.

CONCLUSIONS

In conclusion, this short study has shown that significant improvements in anaerobic fermentation are achievable and that these improvements will reduce the cost to levels which will make the technology attractive in many existing agricultural operations. These studies have been conducted with bench scale and large pilot plant reactors. Thus the data must be carefully evaluated before it is applied to full scale installations.

The following improvements have been shown to be achievable in this study:

- Simple unmixed plug flow gravity operated reactors are as efficient as the conventional completely mixed reactors when operated with dairy cow residues. This design enables the use of inexpensive reactor materials such as soil supported structures and lagoon liners which can be erected by the operator.
- Alternatives to liquid-solid separation such as internal flotation of the fiber with thermophilic fermentation produces multiple products (gases, liquid slurry, and solids) comparable to that achieved with a reactor combined with a unit process such as a centrifuge.
- The anaerobic attached microbial film process has been shown to be one way to achieve a high rate reaction with animal residues. This is achieved by concentrating the active microbial population in the reactor so that flow rates of up to 20 to 30 times that considered acceptable can be used. Further development of this reactor concept should result in a stable methane generation process with HRT values measured in hours instead of days.
- Finally, many fundamental scientific and sociological issues will be created by the development of anaerobic fermentation technology in agriculture. Several of these issues are reviewed in the following chapters. Significant improvements and actual implementation of the technology will be achieved most rapidly by multi-disciplinary teams working closely together.

CHAPTER I

INTRODUCTION AND OBJECTIVES

The availability of a clean, reasonably low cost energy at quantities sufficient to support developed and developing countries' needs represents one of the most important concerns today. Most experts agree that the primary fuels upon which our society depends, namely liquid petroleum and natural gas, have been nearly completely consumed. Because of the diminishing supply of these common fuels, it is estimated that new sources of fuels must be identified, developed, commercialized and fully implemented within one or two decades. One of the most disturbing aspects of this problem is the fact that both developed countries and undeveloped countries have not been limited by a fundamental resource such as energy in history, nor have they had to organize themselves to solve such a widespread common problem within the time frame that exists. Anaerobic fermentation of natural organics to provide substitute natural gas represents one technology which may be capable of supplying clean renewable energy in the near future. This concept is further strengthened when it is noted that pollution control from animal manures can be achieved while extracting energy, conserving nutrients, and synthesizing protein with this process. In fact, it is ironic that today this technology is probably the most well known concept by the layman of any which has been applied in only a few full scale installations on a world-wide basis. This report was intended to show the potential for improving and implementing the technology in the near future.

BACKGROUND

The complicated issues surrounding energy supply estimates are confusing and do not lead to a clear definition of where future energy supplies must be obtained. Some hold great hope that clean fusion power will be available in

the distant future. Others argue that fusion will never be a usable source of power. The only technology that would appear to have the capability of meeting the projected energy growth in the near future is nuclear fusion power. The increasing costs of construction and operation, the time of delivery, the net energy production, and risks to the environment are issues that raised questions as to whether this technology could even serve as an interim power source until cleaner and less expensive technologies are developed. The issue becomes further cloudy when it is noted that the U.S. has coal reserves with energy content to supply its projected energy needs for several hundred years. Since countries such as South Africa are presently using coal to manufacture gasoline, it is clear that the technology also exists to provide the various forms of energy. However, there appear to be economic problems in developing this resource. New mining regulations and the concern for maintaining environmental quality in areas where the coal is mined and burned will add significantly to the cost of this energy source. Although the cost and availability of new energy sources are highly complex, several points appear to stand out in developing judgments as to which energy sources should receive increased attention for development. First, it is likely that natural gas and petroleum energy will continue to increase in price for a period longer than ten years. The cost (in present-day value) of energy could easily be more than ten times the present cost within several decades. Thus the cost of natural or substitute gas could be \$40 per 10^6 kcal (or \$10 per 10^6 Btu) or more. Today, when large energy users of natural gas exceed their quotas, they must pay up to \$24 per 10^6 kcal. Any sizeable energy source that could produce usable energy for less than this value today should be considered as a potential viable source. If the alternative energy source could be obtained without degrading environmental quality and can be converted to usable forms without large-scale production of pollutants, it would have significant advantages over conventional sources such as coal. These are areas in which methane production from agricultural residues hold promise; and it has been projected that more efficient control of agricultural residues using anaerobic fermentation can produce a significant amount of energy at a present cost less than \$10 per 10^6 kcal (\$2.00 per 10^6 Btu), for some large animal raising operations.

The ideal energy source would be rapidly renewable and non-polluting. The Project Independence report estimated that the U.S. could obtain nearly 1.0 percent of its energy use in 1985 from renewable sources and that this would increase to 25 percent by the year 2020 (ERDA 49, 1975). Of this total renewable energy, methane production was estimated to be capable of producing about 50 percent in 1985 and 25 percent in 2020. However, it should be noted that there is a lack of comprehensive information regarding anaerobic fermentation of agricultural residues. Most projections and cost estimates have been developed with technology that is old or that which has been developed for sewage sludge.

Although a considerable amount of information has been developed which aids in understanding the detailed biochemistry and microbiology of the methane generation process, an anaerobic reactor built in 1935 for treatment of animal residues would have the same generation capacity as one would have if built using all available information today. The purpose of this study was to determine the potential for improvement of the technology of methane generation from agricultural residues.

The total quantity of agricultural residues is nearly 20 times the organic wastes in garbage and domestic sewage, and averages nearly 60 pounds per person in the U.S. each day (Eliassen, 1969). If all of this were converted to methane, this would supply more than 20 percent of the total U.S. energy consumption. The technology is not only attractive because it produces a clean fuel from a renewable resource, but it holds promise of producing a significant quantity of energy.

If the utilization of animal wastes for energy generation proved successful, the next logical development would be to utilize solar energy fixed in crops as a major energy source. This concept is attractive since it can be projected that if high yielding crops now available were grown on 17 percent of the land area, the resulting renewable energy would be equal to the total U.S. demand. The anaerobic fermentation bioconversion process has the potential to control pollution by stabilizing organic wastes, to conserve nutrients, can be used in existing agricultural operations, and produces a

replacement for natural gas. A research team at Cornell University, has attempted to determine the technical, economic, and practical feasibility of using this bioconversion technology in the most common animal production operation--cattle production (dairies and beef feedlots) (Jewell et al., 1976). The Cornell study generated some optimistic data regarding the potential of using anaerobic fermentation of animal wastes to produce an energy source for agricultural operations while reducing their pollution potential.

First it was estimated that the quantity of energy that can be made available in dairies and beef feedlots exceeds that used in the operation. In beef feedlots of 1000-head or more, the quantity of energy generated as methane gas was projected to exceed that used on-site by more than six times. This energy accounting approach indicated that a nearly energy self sufficient food production unit might become a reality if technologies such as methane production and utilization were implemented. The cost of generating the gas was estimated to vary from a high of about \$18 per million kcal, (\$4.50 per million Btu) for a 40-cow dairy to around \$4 million kcal, (\$1.00 per million Btu) for a 1000-head beef feedlot. In instances where efficient pollution control is required, the incremental cost of methane generation should be the additional cost above that of other residue management alternatives. In some cases, the cost of energy generation would be zero and would be treated as an income used to decrease the costs of residue management. The anaerobic fermentation process with characteristics of odor control, organic stabilization, and methane production offers a highly useful technology. However, even though the potential of the technology is great, it is far from being developed to a point where U.S. business and agriculture can or will assume responsibility for its development.

There is a need for additional background data on this technology, more fundamental information, and full scale demonstrations are required. During the initial study (Jewell et al., 1976) several unique modifications to existing anaerobic fermentation technology were identified which would appear to hold promise for significant improvements in production capability or reduction in capital costs. The year-long study described in this report is an extension of Cornell's effort to develop an increased information base which may be used to improve anaerobic fermentation technology, as applied to the

conversion of agricultural residues to a usable form of energy.

OBJECTIVES

The overall premise of this study was that existing anaerobic fermentation technology for agricultural residues had changed little since 1936, and that any major activity in this area should include a comprehensive evaluation of the potential for improving this technology. "Improvement" here meant to develop less expensive designs, operations that decreased the labor requirement, and/or designs which enabled recovery of other energy containing by-products other than methane. Earlier work by the Cornell University team emphasized several important factors: the majority of animal residues are produced in smaller scale operations with the small dairy being the most common unit. The total capital cost for these installations would have to be less than \$20,000 to \$30,000 per unit, thus imposing a significant economic constraint. No fermentor design presently available could be implemented for this cost. Thus a new and unique design basis must be developed for these small scale operations.

On the other hand, when given a large quantity of raw material such as would be available on larger animal raising operations, the value of products other than methane could be significant. It would be important in these cases to be able to separate and recover the gases, liquids, and solid matter. The single most important concern here is the problem of separating the fermentor effluent. Unfortunately, a short series of tests showed that this would be a difficult problem. This data suggested that a highly sophisticated reactor be sought which would be capable of high rate methane production and multi-product recovery. Thus two major lines of research were suggested that would serve to illustrate the potential for rapid improvement and implementation of the technology:

1. Development of the basis for a highly simplistic, low-maintenance reactor with nearly automatic operation was needed; and
2. A highly complex high rate gas production unit capable of recovering gases, liquids, and particulate solids was needed.

Both of these two needs were pursued in this study whose general goal was to demonstrate the feasibility of futuristic methane generation reactors for agricultural operations that would result in significant improvement over existing technology.

In order to provide the many facets of information needed to support the dual direction of this study, a number of objectives needed to be satisfied in three major areas. These areas and the specific objectives of these portions of the study are given below:

- I. Develop a simplified low cost anaerobic reactor design.
  - A. Determine the feasibility of the use of a fermentation system with low mixing and heating requirements, such as would be the case with an unheated or minimal heated plug flow reactor design.
  - B. Test the feasibility of developing a semi-solid fermentation system.
- II. Develop high rate reactors capable of multiple product recovery.
  - A. Determine the feasibility of developing a high rate anaerobic fermentor that would be capable of achieving internal liquid-solids separation without a centrifuge or other major unit operations.
  - B. Determine the possibilities of developing a high flow rate system capable of achieving long microbial residence with residues with solids higher than 7 percent.
- III. Develop a comprehensive understanding of the technology so that it could be more easily implemented and managed in agriculture.
  - A. Identify and collect anaerobic fermentation process design information required to define the following: kinetics of decomposition, and inhibition conditions, and temperature effects.
  - B. Identify new management techniques and develop predictive methods for reactor operation by clarifying the fundamental

understanding of the ecology of specific anaerobic fermenting bacteria and by developing an analytical tool capable of direct quantitative measurement of the numbers of viable methanogenic bacteria.

- C. To insure that the designs which are identified will cause minimum interruption to food production and maximum benefits; the incorporation of this technology in dairies and beef feedlots will be examined in the total systems concept.

Successful development of portions of this study were perceived to represent the blocks of information needed to provide the impetus for widespread adoption of anaerobic fermentation technology by agriculture. It is projected that definition of a simplistic reactor would enable individuals to construct a system at minimal costs and would provide a clean, renewable energy source capable of meeting all simple energy demands on the food-raising operation, such as heating and cooking.

Recovery of components in the liquid and particulate solids from anaerobically fermented animal waste offers the interesting potential of converting all the residue into useful products. However, efficient separation of liquid from the solids in fermented residue promises to be a difficult problem.

The following chapters summarize the information that was developed in the above mentioned areas. On-going efforts in 1978 involve only one of the above objectives--development of a simplified low cost reactor.

CHAPTER II

CONVENTIONAL COMPLETELY MIXED ANAEROBIC FERMENTATION

INTRODUCTION

In comparison to sewage sludge, relatively few studies have been conducted on the anaerobic fermentation of agricultural residues. Because of the lack of information on such materials as dairy manure, there is a tendency to extrapolate results of studies on substrates such as sewage sludge or municipal solid wastes to other substrates. Such extrapolations are possible only when the existing experimental data has been carefully defined with fermentor controlling parameters identified. Unfortunately, there remain significant questions regarding the impact of substrate composition, viscosity, nutrients, toxic materials, and other characteristics so that such extrapolations should be accepted as approximations. These approximations are of little use in providing accurate comparisons of varying design approaches or in defining such problems as ammonia toxicity concentrations. Thus it was necessary to evaluate the anaerobic fermentation of dairy waste under controlled conditions in order to provide detailed comparison data for some components of this study.

Earlier studies on municipal solid wastes (Pfeffer, 1974) and beef cattle manure (Bryant et al., 1975; Ashare et al., 1977) raised the possibilities of significantly increasing the rate and quantity of gas yield from a substrate by increasing the reactor temperature. This portion of the study focused on the solids reduction possible and gas production achievable using thermophilic anaerobic fermentation at 60°C. The purpose of this study was three-fold: to study the conventional completely mixed anaerobic fermentation of dairy cow manure at thermophilic temperatures, to provide control data for the thermophilic liquid-solid separation flotation fermentor, and to provide data to support a kinetic modeling approach.

The specific objectives of this study were to:

1. Determine the effect of thermophilic temperatures on the performance of anaerobic fermentation of dairy cow manure;
2. Determine the refractory fraction of dairy cow manure when it is anaerobically fermented at thermophilic temperatures;
3. Examine the results of this study in terms of a mathematical design model;
4. Provide a basis of comparison of other studies on the thermophilic fermentation of dairy cow manure; and to
5. Compare the results of high temperature fermentation obtained in this study to mesophilic digestion of the same substrate.

LITERATURE REVIEW

The most common form of the anaerobic fermentation process is a single-stage, completely mixed, semi-continuous flow system. In this type of reactor, the substrate is introduced, and effluent is drawn off periodically in a manner such that a constant volume of material is always maintained in the system. No recycling of solids is employed.

The environmental and operational parameters of concern, when fermenting any type of organic material, are presented in Table 1 (McCarty, 1964). Many of the operational parameters correspond to environmental parameters. By varying the method of operation, these systems can be operated at or near optimum environmental conditions.

Environmental Conditions--pH, Alkalinity, and Volatile Acids

The three environmental parameters of pH, alkalinity, and volatile acid concentration are all interrelated, with pH being the most important. Stable anaerobic treatment will result if the pH is between 6.6 and 7.6 with an optimum range being 7.0 to 7.2 (McCarty, 1964). Outside of this range, the process is thought to be less efficient. Alkalinity in these units provides

TABLE 1. OPERATIONAL AND ENVIRONMENTAL PARAMETERS AFFECTING ANAEROBIC FERMENTATION.

Environmental Parameters	Operational Parameters
1. pH	1. Composition of Substrate
2. Alkalinity	2. Retention Time
3. Volatile Acid Concentration	3. Substrate Concentration
4. Temperature	4. Organic Loading Rate
5. Nutrient Availability	5. Degree of Mixing
6. Toxic Substances	6. Heating

the "buffering capacity," so that if there is a small volatile acid accumulation, the pH of the reactors will not be adversely affected. The volatile acid concentration in the system is an important parameter which should be monitored. If the volatile acid concentration increases, this indicates that the system is not in equilibrium and the methane forming bacteria in the system may be inhibited by some environmental parameter(s). Thus, if the volatile acid concentration continues to accumulate, a pH drop will eventually result and the reactor will "fail."

Temperature and Heating

Two distinct temperature ranges are optimum for anaerobic fermentation, these being the mesophilic temperature range, 30 to 40°C, and the thermophilic temperature range, 43 to 60°C. Heat must be supplied to the reactor so that these temperatures can be maintained within the system.

Nutrient Availability and Composition Of Substrate

For optimum microbial growth in anaerobic systems, many organic and inorganic nutrients must be supplied. These nutrients include carbon, hydrogen, nitrogen, phosphorus, and sulfur (Loehr, 1966), in addition to trace minerals. Dairy cow manure is rich in these nutrients; therefore, addition of any of these will seldom be necessary.

Toxic Substances

Many substances, if present in a high enough concentration, can be toxic to the anaerobic fermentation process. These substances include sulfides (>200 mg/l), soluble heavy metals (>1 mg/l), alkalinity with metals such as sodium (5000-8000 mg/l), potassium (4000-10,000 mg/l), calcium (2000-6000 mg/l), and magnesium (1200 to 3500 mg/l) and ammonia (1700-4000 mg/l) (EPA 1974).

Mixing

Mixing in anaerobic reactors is provided to accomplish many functions, among which are : 1) breakup of scum layer; 2) distribution of heat;

3) distribution of influent in the reactor; 4) distribution of toxic materials or metabolic products; and 5) prevention of settling (Loehr, 1974; WPCF Manual of Practice No. 16, 1968).

Retention Time, Substrate Concentration, and Organic Loading Rate

The three parameters of retention time, substrate concentration, and organic loading rate are intimately related. Figure 3 illustrates the relationship between these three parameters. The organic loading rate is defined as the quantity of organic material supplied to the reactor per day, expressed in this study as grams of volatile solids added per liter of reactor per day. Two distinct retention times exist, these being the hydraulic retention time (HRT) and the solids retention time (SRT). The HRT represents the theoretical residence time of the liquid in the system while the SRT is the theoretical residence time of the solids. In systems where no solids are recycled into the reactor from the effluent, the HRT and SRT should be equal. The substrate concentration, in this study, will represent the total solids (expressed as percent solids) of the influent material.

Thermophilic digestion of municipal domestic sludge has been investigated in a number of laboratory studies (Malina, 1961; Pohland and Bloodgood, 1963; Rowe, 1971) and at larger scale installations (Garber, 1954; O'Hara and Colbaugh, 1975). In addition, laboratory studies have examined anaerobic digestion of dairy cow manure at mesophilic temperatures (Morris, 1976; Fong, 1973; Gramus et al., 1971; Halderson, 1972; Jewell et al., 1976). However, comparatively little work has been done at thermophilic temperatures on the fermentation of dairy cow manure (Bryant et al., 1976). A summary of three different studies on thermophilic fermentation of dairy cow manure is presented in Table 2. Bryant et al., (1976) observed the effect that high loading rates had on the process. TVS removal efficiencies of 17.7 to 31.9 percent were achieved, while methane production rates of 1.0 to 3.0 liters of methane per liter of reactor volume per day were observed. Laboratory thermophilic fermentations were also reported by Fong (1973) and Jewell (1976), although both these studies employed lower organic loadings than those utilized by Bryant. It is difficult to make direct comparisons between studies

Figure 3. Organic loading rate as a function of HRT in a completely mixed fermentor receiving dairy cow manure, assuming VS/TS equals 0.88.

TABLE 2. SUMMARY OF DATA ON THERMOPHILIC ANAEROBIC FERMENTATION OF DAIRY COW MANURE FROM VARIOUS DIFFERENT STUDIES.

HRT days	Temp. °C	Organic Loading		T.V.S. Reduction %	Methane Production		Ref.
		Solids Conc. % T.V.S.	gm T.V.S./ day		ml CH ₄ gm VS added	l CH ₄ /l reactor day	
10	52	2.40	2.4	28.0	117	--	Fong, 1973
15	52	3.60	2.4	28.0	81	--	"
10	52	3.20	3.2	24.5	122	--	"
15	52	4.80	3.2	25.0	85	--	"
10	52	4.00	4.0	20.5	117	--	"
15	52	6.00	4.0	23.0	82	--	"
3	60	6.00	20.0	23.7	118	2.36	Bryant <u>et al.</u> , 1973
6	60	6.00	10.0	26.1	137	1.33	"
9	60	6.00	6.67	31.9	151	1.02	"
3	60	8.00	26.7	20.8	105	2.80	"
6	60	8.00	13.3	25.9	137	1.79	"
9	60	8.00	8.89	30.2	135	1.19	"
3	60	10.00	33.3	22.2	87	2.90	"
6	60	10.00	16.7	26.7	111	1.84	"
6	60	12.00	20.0	27.2	82	1.63	"
9	60	12.00	13.3	29.6	111	1.49	"
6	60	14.0	23.3	17.7	45	1.03	"
9	60	14.0	15.6	23.8	74	1.10	"
5	60	--	17.6	19.0	--	--	Jewell, 1976
10	60	--	2.56	24.0	--	--	"

of this nature, since many external factors can influence certain dairy cow manure characteristics. The composition of dairy cow manure is subject to certain variables which include the type and size of the animals, stage of lactation, feed rations, and the time of year at which the study was conducted. These factors may cause substantial changes in the organic and inorganic content, the biodegradability of the manure and the percent solids at which the manure is generated. Therefore, care must be exercised in comparing the results of one study to another.

THEORETICAL CONSIDERATIONS

An empirical design model will be used in this study to analyze the data generated at steady state from the thermophilic completely mixed reactors. The purpose of developing and using an empirical model are two-fold: to determine if the reactor performance data can be described through a set of simple mathematical equations, and for use in design of systems operating under similar conditions.

The parameter used as a measure of substrate concentration will be total volatile solids (TVS). This parameter provides more accurate measurements than the COD analysis because of sampling errors in relation to the nature of the analyses. When dealing with a highly concentrated particulate slurry, dilutions of greater than 100:1 are required to determine the COD. Because of the high dilutions required, obtaining a representative sample becomes difficult. Thus the sampling techniques and dilutions required can introduce substantial experimental error into the COD test. The TVS test requires no dilutions, and thus utilization of this test is felt to be a more practical measurement of organic conversion or destruction efficiencies.

Biodegradability Studies

Highly concentrated animal manures such as dairy cow manure contain a large fraction of nonbiodegradable or refractory organic material. The TVS are comprised of two fractions, biodegradable volatile solids (BVS) and refractory volatile solids. The definition of the refractory fraction of the organic volatile solids can be stated as that portion of the TVS that are

resistant to microbial attack over long periods of time and remain after the rate of degradation of the TVS decrease to a very low level (Jewell, 1968). The following conceptual modeling approach was developed by Jewell (1968) and used by Morris (1972) to describe anaerobic fermentation of dairy manure. Equation 1 describes the two fractions that comprises the TVS (Jewell, 1968):

$$S_o = S_{bo} + S_{ro} \quad \text{Eq. 1}$$

where: S_o = influent TVS conc. (gm/l)
 S_{bo} = influent BVS conc. (gm/l)
 S_{ro} = influent refractory volatile solids conc. (gm/l)

Since the refractory fraction of the TVS is resistant to microbial attack, the influent and effluent refractory solids concentrations remains constant. Thus the effluent from the system can be described by the following equation:

$$S_1 = S_{b1} + S_{r1} \quad \text{Eq. 2}$$

where: S_1 = effluent TVS conc. (gm/l)
 S_{b1} = effluent BVS conc. (gm/l)
 S_{r1} = effluent refractory volatile solids conc. (gm/l)

and: $S_{ro} = S_{r1} = S_r \quad \text{Eq. 3}$

The refractory fraction of the TVS in the influent wastes stream can thus be described as:

$$R = (S_r)/(S_o) \quad \text{Eq. 4}$$

where R is the ratio of the influent refractory volatile solids to the influent TVS.

To determine the refractory fraction (R) of dairy cow manure total volatile solids, a method used by Anthonisen *et al.*, (1968) and Wood *et al.*, (1974) and modified by Morris (1976) will be employed. In this method the dairy cow manure is anaerobically fermented in a batch reactor. Samples are withdrawn at various intervals and analyzed for TVS. The assumption is that as the solids retention time (SRT) approaches infinity, the biodegradable fraction of the manure will be destroyed, leaving only the refractory fractions remaining. A plot of S_1/S_0 versus $1/S_0$ (SRT) is made (as shown in Figure 4). This will produce a linear relationship with the ordinate intercept being the refractory fraction of the organic matter.

Measurement of the biodegradable fraction of the residue is necessary since it is the driving force for microbial substrate utilization. Thus, performance of an anaerobic reactor to produce methane gas should be based upon its ability to react with BVS and not TVS. The subsequent empirical design model will be developed to relate the performance of these systems to their capability of converting BVS to methane gas.

Empirical Design Model

A mass balance approach will be used to describe a first order reaction for the removal of organic wastes in an anaerobic fermentor. Many investigators have used similar models to describe organic material destruction in activated sludge processes (McKinney, 1962; Goodman and Englande, 1974), aerobic treatment of poultry manure (Prakasam *et al.*, 1974), and in anaerobic digester systems (Morris, 1976; Gaddy *et al.*, 1974). The design approach incorporating the biodegradable concept combined with a mass balance, and first order decay coefficient will be used in this study (Jewell *et al.*, 1976, Morris, 1976).

The assumption is made that the influent flow rate and effluent flow rate are equal, therefore the reactor volume remains constant. The flow rate and reactor volume are related in the following expression to determine the liquid residence time in the reactor:

$$\text{HRT} = V/Q \qquad \text{Eq. 5}$$

Figure 4. Graphical determination of the refractory fraction, R .

where: Q = flow rate (ℓ/day)
                V = reactor volume (ℓ)
               HRT = hydraulic retention time (days)

In a completely-mixed single-staged system where recycle of solids is not practiced, it is noted that:

$$\text{HRT} = \text{SRT}$$

where: SRT = solids retention time (days)

A mass balance relationship describing the change in substrate concentrations in the reactor may be expressed as follows:

$$\text{Change of Substrate Conc. in Reactor} = \left[\begin{array}{c} \text{Influent} \\ \text{Substrate Conc.} \end{array} \right] - \left[\begin{array}{c} \text{Effluent} \\ \text{Substrate Conc.} \end{array} \right] - \left[\begin{array}{c} \text{Substrate} \\ \text{Removal} \end{array} \right]$$

Assuming first order substrate utilization kinetics, the mass balance equation may be explicitly written:

$$V \frac{dS_b}{dt} = QS_{bo} - QS_{b1} - KVS_{b1} \quad \text{Eq. 6}$$

where: $V \frac{dS_b}{dt}$ = change in BVS with time (g/day)
                K = substrate removal coefficient (day^{-1})

At steady state dS_b/dt will equal zero, and substituting equation 5 into equation 6 results in:

$$S_{bo} - S_{b1} = KS_{b1} (\text{HRT})$$

Rearranging terms and solving for the effluent BVS yields:

$$S_{b1} = S_{bo} / [K(\text{HRT}) + 1] \quad \text{Eq. 7}$$

The HRT can be expressed as:

$$\text{HRT} = (S_{b0} - S_{b1})/K(S_{b1}) \quad \text{Eq. 8}$$

and the substrate removal coefficient, K, is shown to be:

$$K = [S_{b0} - S_{b1}]/\text{HRT}(S_{b1}) \quad \text{Eq. 9}$$

The approach shown in equation 8 was first shown to be valid in 1968 by Jewell under aerobic conditions and subsequently by Foree and McCarty (1968) under anaerobic conditions with decomposing algae. Equation 8 indicates that the substrate removal coefficient, K, can be determined by plotting $(S_{b0} - S_{b1})/\text{HRT}$ versus (S_{b1}) , as shown in Figure 5. This equation states that this plot will produce a straight line that will go through the origin and has a slope equal to K.

The efficiency of BVS removal can similarly be described by:

$$E_b = 100(S_{b0} - S_{b1})/S_{b0} \quad \text{Eq. 10}$$

where: E_b = efficiency of BVS removal (%)

By using equations 1, 2, and 3 and substituting these into equations 7, 8, 9, and 10, a similar set of equations that incorporate the biodegradable concept can be developed to describe effluent TVS, TVS removal efficiency, and HRT and K in terms of TVS:

$$S_1 = S_0(1 - R)/[K(\text{HRT}) + 1] - RS_0 \quad \text{Eq. 11}$$

$$K = (S_0 - S_1)/\text{HRT}(S_1 - RS_0) \quad \text{Eq. 12}$$

$$\text{HRT} = (S_0 - S_1)/K(S_1 - RS_0) \quad \text{Eq. 13}$$

Figure 5. Graphical determination of the substrate removal coefficient, K.

$$E_t = 100(S_0 - S_1)/S_0 \quad \text{Eq. 14}$$

where: E_t = efficiency of TVS removal (%)

Lastly, the rate of change of substrate concentrations in the reactor with time, or the rate of waste utilization can be described as:

$$\frac{\Delta S}{\Delta t} = \frac{dS}{dt} = (S_0 - S_1)/\text{HRT} \quad \text{Eq. 15}$$

It has been determined through stoichiometry that for each gram of COD as complex organics stabilized, 0.35 liters of methane gas are produced at STP (McCarty, 1964). If it is assumed that the ratio between biodegradable COD reduction and biodegradable solids reduction remains constant throughout the anaerobic fermentation process, and knowing that the ratio of COD to VS for most complex organics is 1.4, the quantity of gas production per unit volume of reactor can be predicted on the basis of volatile solids destruction:

$$G_b = 0.5(S_{b0} - S_{b1})/\text{HRT} \quad \text{Eq. 16}$$

and

$$G_t = 0.5(S_0 - S_1)/\text{HRT} \quad \text{Eq. 17}$$

where: G_b = methane production at STP (volume CH_4 /volume reactor/day)
 G_t = methane production at STP (volume CH_4 /volume reactor/day)

It should be noted that G_b will equal G_t , since $(S_{b0} - S_{b1})$ will equal $(S_0 - S_1)$.

A summary of the design equations presented in this section is shown in Table 3. These equations will be used in evaluating the performance of the thermophilic reactor operated in this study.

TABLE 3. SUMMARY OF DESIGN EQUATIONS DETERMINED FROM THE EMPIRICAL MODEL.

Design Parameter	Equation	Equation No.
Effluent Volatile Solids		
Biodegradable	$S_{b1} = S_{bo}/[K(HRT) + 1]$	(7)
Total	$S_1 = S_o(1 - R)/[K(HRT) + 1] - RS_o$	(11)
HRT		
Biodegradable	$HRT = (S_{bo} - S_{b1})/KS_{b1}$	(8)
Total	$HRT = (S_o - S_1)/K(S_1 - RS_o)$	(13)
K		
Biodegradable	$K = (S_{bo} - S_{b1})/HRT(S_{b1})$	(9)
Total	$K = (S_o - S_1)/HRT(S_1 - RS_o)$	(12)
Efficiency		
Biodegradable	$E_b = 100(S_{bo} - S_{b1})/S_{bo}$	(10)
Total	$E_t = 100(S_o - S_1)/S_o$	(14)
Methane Production		
Biodegradable	$G_b = 0.5(S_{bo} - S_{b1})/HRT$	(16)
Total	$G_t = 0.5(S_o - S_1)/HRT$	(17)
Rate of Waste Utilization		
General	$ds/dt = (S_o - S_1)/HRT$	(15)

MATERIALS AND METHODS

Four thermophilic semi-continuously fed completely mixed anaerobic reactors were used in this portion of the study. Each reactor was 15 liters in volume and each was operated at a different percent total solids feed, namely 4, 6, 8, and 10 percent total solids.

These four reactors were operated on a semi-continuously fed basis, using a fill and draw technique. Each day an aliquot of the slurry was withdrawn from each reactor and replaced with an equal volume of feedstock. Since no solids recycle was employed, and since the reactor contents were completely mixed during wasting and feeding, the hydraulic retention time (HRT) was equal to the solids retention time (SRT).

Thermophilic temperatures (60°C) were maintained in these reactors. This was accomplished by placing these reactors in a constant temperature environmental chamber. Figure 6 presents a schematic of one of these reactors.

Manure Feedstock

The manure used as a feedstock for all of the fermentors described in this report was obtained from one source at Cornell University's teaching and research dairy located at Harford, New York. The manure was obtained from an area housing approximately 120 graded Holstein cows, with approximately 60 cows in mid-lactation and 60 cows in late lactation stages. The number of dry cows varied between zero and ten.

The diet of the mid-lactation cows consists of 48 percent corn and hay silage, 45 percent brewers grain and 7.0 percent cracked corn, on a dry matter basis. The feed has a total digestible nutrient (TDN) value of about 71 percent and a protein content of 15.5 percent. The late-lactation and dry cows were fed left-over ingredients, which include corn and hay silage, brewers grain, cracked corn, soybean meal and small amounts of poultry and dairy cow manure in varying proportion. This diet varied on a day to day basis.

Figure 6. Schematic of the thermophilic reactor apparatus.

The barn itself is an open-ventilation, free-stall area in which floor scrapers, shown in a photograph in Figure 7, draw the manure down to a cross conveyor at the end of the barn. The residue on the floor consists of a mixture of urine, feces and a small amount of wasted feed.

Manure was stored in a refrigerator at 4°C to minimize degradation. Prior to feeding, the manure was removed from the refrigerator and mixed with hot tap water to obtain the desired percent solids concentration. Four different ratios of manure and water were made up daily, one for each of the four reactors. For short HRT conditions when a large portion of the digester contents was replaced with feedstock, the raw manure was preheated to the exact reactor temperature to prevent thermal shock to the microorganisms.

Reactor Operation and Management

Prior to feeding the reactor, each reactor was vigorously mixed by hand to provide a homogenous effluent, and the prescribed amount of manure was then removed from the unit. The reactor was then filled with the raw manure, and, when necessary, was brought up to the 15 liter volume with additional water. Adjustment of the reactor volume was necessary because moisture would be carried out of the reactor with the gas, thus slowly reducing the volume of the reactor contents. The reactor was then manually mixed again to insure a uniform distribution of the feed.

The gas produced by each reactor was measured by a wet test gas meter. The wet test meters were kept at ambient temperatures (20°C) and all raw gas production figures were recorded at this temperature. When this information was summarized, the gas production figures were corrected to STP (1 ATM pressure and 0°C) as indicated in the data tables. Gas was collected and analyzed for CO₂ and CH₄ content in gas collection tubes.

These four thermophilic reactors were operated for a total of 14 conditions at HRTs ranging from 2.5 to 30 days. A summary of all conditions at which these reactors were operated is presented in Table 4.

Figure 7. Delta scraper system used to collect manure and urine substrate for all project fermentors, Cornell University.

TABLE 4. CONDITIONS AT WHICH THE THERMOPHILIC COMPLETELY MIXED REACTORS WERE OPERATED.

HRT (days)	Digester 1 4% T.S. Feed	Digester 2 6% T.S. Feed	Digester 3 8% T.S. Feed	Digester 4 10% T.S. Feed
2-1/2			X	X
3-1/2	X	X		
5	X	X	X	X
10	X	X	X	X
20		X		
30				X

TABLE 5. ANALYSIS SCHEDULE FOR THERMOPHILIC REACTORS.

Analysis Parameter	Influent Feed	Effluent			Gas	
		Daily	Bi-Weekly	Steady State	Daily	Steady State
Total Solids	X		X	X		
Total Volatile Solids	X		X	X		
TKN	X			X		
NH ₄ N	X		X	X		
pH	X	X	X	X		
Alkalinity	X		X	X		
Volatile Acids			X	X		
Total Gas Measurement					X	X
Gas Composition						X
Temperature		X	X	X		
CST	X			X		

In addition to the reactors operated during this study, the total biodegradability of the manure was determined in long term batch experiments. Effluent from each thermophilic, semi-continuous fermentor was placed in a 1.5 liter batch reactor and allowed to react for long periods at 60°C. Effluent was transferred only when the semi-continuous reactors were operating at steady state and only one day's effluent was used. The batch fermentors were manually mixed on a daily basis. Over a period of 50 to 70 days, weekly samples were withdrawn and analyzed for total and total volatile solids. These data were then analyzed, and the total biodegradability of this material was determined.

Table 5 presents a listing of all analyses performed on the influent feed, liquid effluent, and gas. The influent feed was analyzed each time a new batch of feed was obtained, usually about once per week. The liquid effluent was analyzed only for temperature and pH on a daily basis; however, more complete analysis were performed on a bi-weekly basis, and when the units were thought to be at steady state. Gas production was measured daily and analyzed for composition at steady state.

Total and total volatile solids, pH, alkalinity, volatile acids, gas measurement and composition, and temperature were all determined according to Standard Methods (1971). Total kjeldahl nitrogen (TKN) and ammonia nitrogen were analyzed according to Wallace and MacKenzie (1954) and the CST test was performed according to Baskerville and Gale (1968).

RESULTS

The four anaerobic thermophilic semi-continuously fed completely mixed reactors were operated for a period of 350 days. During this time 14 conditions were investigated, and steady state data were collected for each. In addition to the investigation of reactor performance at various conditions, biodegradability studies, nutrient analyses of the effluent, and microbiological studies were performed. Nutrient analyses and microbiological determinations have been presented elsewhere in this report and will not be covered here.

Biodegradability Studies

As described previously, biodegradability studies were initiated to determine the refractory fraction. In order to insure that the refractory fraction (R) was constant for a wide range of manure feed concentrations, biodegradability determinations were conducted on the effluent from the four semi-continuous fermentors operated at steady state, receiving 4, 6, 8, and 10 percent total solids (TS) dairy cow manure. Figure 8 presents the results of biodegradability batch tests conducted on the effluents from semi-continuous reactors operated at 5 and 10-day HRT's while receiving 4, 6, and 8 percent TS feedstock manure and at 2.5, 5, and 10-day HRT's while receiving a 10 percent TS influent. As can be seen from the data presented in Figure 8, the refractory fraction of this material ranged from 0.47 to 0.50, or 47 to 50 percent of the influent total volatile solids. Thus the total biodegradable fraction of the manure at thermophilic temperatures varied from 50 to 53 percent. It is felt that the differences in the respective refractory fractions of the manure at different substrate concentrations is due to experimental error rather than any difference in the refractory fractions of the manure itself. Thus a representative value of 0.49 was chosen for "R" and this value was used throughout this study. The average biodegradable fraction of the manure anaerobically digested at thermophilic temperatures was 51 percent.

Reactor Performance

A summary of all data gathered at steady state for all fourteen conditions is presented in Tables A1 through A14 in Appendix A. These data are the basis for the summary tables which appear in the remainder of this section. An example of daily gas production for one of the 14 reactors is presented in Figure 9.

The criteria used in evaluating the performance of the semi-continuous thermophilic reactors primarily were based upon volatile solids destruction, and methane gas production. Other parameters, such as volatile acids, alkalinity, and pH, were also monitored to serve as supportive indicators of

Figure 8. Graphical determination of the refractory fraction of dairy cow manure using batch decay experiments with effluents from semi-continuous thermophilic fermentors operating at steady state.

Figure 9. Chronological development of gas production in a semi-continuously fed thermophilic fermentor operated at an influent manure feed concentration of 10% TS.

digester malfunction. These data are presented in Tables 6 through 9 for thermophilic fermentors receiving 4, 6, 8, and 10 percent TS manure feed.

Distress signs of anaerobic fermentation typically include depressed gas production and solids destruction efficiency, low alkalinity and pH, and a build-up of intermediate products of fermentation or volatile fatty acids. Since alkalinity was measured in this study through the titration of samples to a methyl orange endpoint (pH 4.0) in accordance with Standard Methods (1971), the total alkalinity data may or may not be a true reflection of the reactor's buffering capacity in the optimum fermentive pH range of 7.0-7.2 (McCarty, 1964). Bicarbonate alkalinity, exerting maximum buffering at pH 6.3 is usually considered to be a more accurate indicator of the reactor's buffering stability at neutral pH values. Volatile acids impart their maximum buffering capacity well below 6.0, and, if present in sufficient quantity, may distort total alkalinity as an indicator of fermentor well-being. A relationship between total alkalinity, volatile acids, and bicarbonate alkalinity is given in the following expression (McCarty, 1971):

$$BA = TA - (0.71)TVA \quad \text{Eq. 18}$$

where: BA = bicarbonate alkalinity, mg/l as CaCO_3
             TA = total alkalinity, mg/l as CaCO_3
             TVA = total volatile acids, mg/l as acetic acid

Using this expression and measurements of total alkalinity and volatile acids, the bicarbonate alkalinity was then computed for all operating conditions of the thermophilic semi-continuous fermentors. Steady state total alkalinity and bicarbonate alkalinity data have been included in Tables 6 through 9 for digesters operated on 4, 6, 8, and 10 percent TS dairy cow manure feedstock. Unstressed reactor systems seem to show that some correlation exists between influent solids concentration or loading and bicarbonate alkalinity in the effluent. As illustrated in the plots in Figure 10 for the 5 and 10-day HRT operating modes, effluent alkalinity rises almost linearly with increased influent volatile solids loading.

TABLE 6. SUMMARY OF STEADY STATE DATA FOR 4% T.S. FEED
THERMOPHILIC REACTOR.

Parameter	Units	HRT = 3-1/3 days	HRT = 5 days	HRT = 10 days
T.S. Loading Rate	gm/ℓ-day	12.0	8.0	4.0
Influent T.S.	gm/day	180.0	120.0	60.0
Effluent T.S.	gm/day	163	99.0	49.2
T.S. Reduction	%	9.22	17.5	18.0
Influent T.V.S.	gm/day	160	106	53.4
Effluent T.V.S.	gm/day	144	86.1	41.6
T.V.S. Reduction	%	10.1	19.2	22.1
Gas Production	ℓ/day*	4.79	17.3	11.3
Gas Composition	CH ₄ %	---**	51.0	61.0
	CO ₂ %	---	49.0	39.0
Gas Production Predicted From T.V.S Destruction	ℓ/day	16.2	20.0	9.67
Influent C.S.T.	sec	1120	1440	---
Effluent C.S.T.	sec	760	907	---
Total Volatile Acids	mg/ℓ	2560	996.0	1700
pH		6.2	6.8	7.3
Alkalinity as CaCO ₃	mg/ℓ	1830	3560.0	5630
Bicarbonate Alkalinity as CaCO ₃	mg/ℓ	20.0	2360	4430

* Corrected for STP (1 ATM and 0°C)

** Assumed 50% for T.V.S. gas production calculations

TABLE 7. SUMMARY OF STEADY STATE FOR 6% T.S. FEED
THERMOPHILIC REACTOR.

Parameter	Units	HRT = 3-1/3 days	HRT = 5 days	HRT = 10 days	HRT = 20 days
T.S. Loading Rate	gm/l-day	18.0	12.0	6.0	3.0
Influent T.S.	gm/cay	270	180	90.0	45.0
Effluent T.S.	gm/day	238	143	72.6	33.8
T.S. Reduction	%	11.7	20.7	19.3	24.9
Influent T.V.S.	gm/day	240	160	80.0	39.8
Effluent T.V.S.	gm/day	209	124	61.2	28.5
T.V.S. Reduction	%	12.9	22.5	23.5	28.4
Gas Production	ℓ/day*	13.1	28.8	16.9	16.1
Gas Composition	CH ₄	%	38.0	51.0	62.0
	CO ₂	%	62.0	49.0	38.0
Gas Production Predicted From T.V.S. Reduction	ℓ/day*	16.9	35.2	15.2	10.3
Influent C.S.T.	sec	2660	2520	---	1390
Effluent C.S.T.	sec	1540	2220	---	966
Total Volatile Acids	mg/l	2770	2160	1680	130
pH		6.4	7.0	7.4	7.0
Alkalinity as CaCO ₃	mg/l	3470	5380	8200	2770
Bicarbonate Alkalinity as CaCO ₃	mg/l	1510	3850	7010	2680

* Corrected for STP (1 ATM and 0°C)

TABLE 8. SUMMARY OF STEADY STATE DATA FOR 8% T.S. FEED
THERMOPHILIC REACTOR.

Parameter	Units	HRT = 2-1/2 days	HRT = 5 days	HRT = 10 days
T.S. Loading Rate	gm/l-day	32.0	16.0	8.0
Influent T.S.	gm/day	480.0	240.0	120.0
Effluent T.S.	gm/day	388	188	97.8
T.S. Reduction	%	19.2	21.5	18.5
Influent T.V.S.	gm/day	427	213	107
Effluent T.V.S.	gm/day	334	163	82.4
T.V.S. Reduction	%	21.8	23.4	22.8
Gas Production	ℓ/day*	45.8	39.4	20.3
Gas Composition CH ₄	%	52.0	52.0	55.0
CO ₂	%	48.0	48.0	45.0
Gas Production Predicted From T.V.S. Destruction	ℓ/day*	89.4	47.9	22.1
Influent C.S.T.	sec	3760	5010	---
Effluent C.S.T.	sec	2490	4410	---
Total Volatile Acids	mg/ℓ	1500	1790	2620.
pH		6.9	7.1	7.4
Alkalinity as CaCO ₃	mg/ℓ	3970	7040.0	10350.0
Bicarbonate Alkalinity as CaCO ₃	mg/ℓ	2910	5770	8490

* Corrected for STP (1 ATM and 0°C)

TABLE 9. SUMMARY OF STEADY STATE DATA FOR 10% T.S. FEED THERMOPHILIC REACTOR.

Parameter	Units	HRT = 2-1/2 days**	HRT = 5 days	HRT = 10 days	HRT = 30 days**
T.S. Loading Rate	gm/l-day	46.0	20.0	10.0	3.3
Influent T.S.	gm/day	300.0	300.0	150.0	50.0
Effluent T.S.	gm/day	236	218	119	31.5
T.S. Reduction	%	21.2	27.2	20.6	37.0
Influent T.V.S.	gm/day	267	266	133	44.4
Effluent T.V.S.	gm/day	206	190	100	29.7
T.V.S. Reduction	%	22.8	28.8	24.9	33.1
Gas Production	l/day*	37.6	49.5	26.6	17.1
Gas Composition CH ₄	%	52.0	52.0	58.0	54.0
CO ₂	%	48.0	48.0	42.0	46.0
Gas Production Predicted From T.V.S. Reduction	l/day*	58.6	73.7	28.6	13.6
Influent C.S.T.	sec	5040	8080	---	4340
Effluent C.S.T.	sec	3120	4640	---	1960
Total Volatile Acids	mg/l	1760	2030	4590	163.0
pH		7.0	7.2	7.5	7.2
Alkalinity as CaCO ₃	mg/l	4170	8200.0	13,500	3930
Bicarbonate Alkalinity as CaCO ₃	mg/l	2920	6760	10,200	3820

* Corrected for STP:(1 ATM and 0°C)

** Reactor Volume = 7.5 liters

Figure 10. Effect of the influent volatile solids concentration on the bicarbonate alkalinity of bench scale thermophilic semi-continuous fermentors operated at 5 and 10 days HRT.

In general, the thermophilic semi-continuous fermentors operating at hydraulic retention times of 5, 10, 20, and 30 days seemed to function with little indication of stress. Upon examination of the steady state data in Tables 6 and 7 it is evident that the only reactors showing gross symptoms of stress were those operated at a 3-1/3-day HRT receiving 4 percent and 6 percent TS manure feed. Both of these units experienced low gas production rates accompanied by depressed effluent pHs, depleted bicarbonate alkalinities, and inordinately high volatile acid concentrations.

In interesting contrast, the reactors receiving 8 percent and 10 percent TS feed and operated at a still shorter HRT of 2.5 days effected the highest rate of gas production observed from any of the thermophilic semi-continuous reactor conditions with little sign of stress due to overloading. One possible explanation for this would be the high buffering capacity offered by the increased ammonia and potassium bicarbonate that was achieved at higher feed concentrations. This pattern of fermentor performance at short solids retention times seems to suggest an interrelationship between influent substrate or mixed liquor substrate concentration and the fermentor's potential for maintaining a high enough biomass growth rate to prevent a situation of overloading at high hydraulic loadings.

The rate of fermentation is commonly described either in terms of substrate (volatile solids) utilization or biogas evolution. As previously shown by the derivation of Equation 17, solids destruction and gas production are related to each other through mass balance considerations. For steady state fermentation, the theoretical gas production calculated from the solids destruction should reasonably compare to the amount of gas actually produced. The magnitude of discrepancy, then, between the theoretical and observed gas production rates serves as a check on the analytical accuracy of the results.

The highest methane gas production rate per unit volume of reactor obtained from the thermophilic control fermentation study was 2.6 l/l-day (STP) observed in a reactor operated at a 2.5 day HRT and 10 percent TS feed. The loading rate for this unit was the highest applied during the study, 2.2 lbs TS/ft³-day (267 kg TS/m³-day). Loading rate data and methane gas production

TABLE 10. LOADING AND METHANE GAS PRODUCTION DATA FROM THE THERMOPHILIC SEMI-CONTINUOUS REACTORS.

HRT days	Influent T.S., g/l	Influent T.V.S., g/l	Loading lbs T.V.S./ft ³ /day	Observed Methane Production l/l reactor/day (STP)
2-1/2	80	71.1	1.77	1.59
2-1/2	100	88.9	2.22	2.60
5	40	35.6	0.44	0.59
5	60	53.3	0.66	0.98
5	80	71.1	0.88	1.37
5	100	88.8	1.11	1.72
10	40	35.6	0.22	0.46
10	60	53.3	0.33	0.70
10	80	71.1	0.44	0.74
10	100	88.8	0.55	1.03
20	60	53.3	0.17	0.59
30	100	88.8	0.18	0.62

rates for the thermophilic reactors showing little or no signs of operational distress, according to pH, alkalinity, and volatile acid data, are listed in Table 10 and shown in Figure 11. As the loading increases from 0.17 lbs/ft³-day to 2.6 lbs/ft³-day, the methane gas production rate gradually rises from a low of 0.46 l/l reactor-day to a maximum of 2.6 l/l reactor-day.

Total volatile solids and biodegradable volatile solids destruction data are tabulated in Table 11. These data compare favorably with earlier studies reported in Table 2. Total volatile solids destruction efficiencies in anaerobic fermentors showing little signs of distress ranged from 19 to 33 percent. Similarly, biodegradable volatile solids reductions for the same reactors varied between 38 and 65 percent. Under thermophilic conditions, total volatile solids and biodegradable volatile solids in reactor effluents were decreased when solids retention times increased, as shown in Figure 12.

Figure 11. Comparison of thermophilic and mesophilic gas production rates at varying solids loading rates.

TABLE 11. SUMMARY OF T.V.S. AND BIODEGRADABLE T.V.S. REDUCTION FOR ALL CONDITIONS AT WHICH THE THERMOPHILIC REACTORS WERE OPERATED.

SRT days	Influent Feed Solids %	Influent T.V.S. S_0 gm/l	Influent Biodegradable T.V.S., S_{bo} gm/l	Effluent T.V.S. S_1 gm/l	Effluent Biodegradable T.V.S., S_{b1} gm/l	T.V.S. Reduction Efficiency %	Biodegradable T.V.S. Reduction Efficiency %
2.5	8.0	71.1	36.3	55.6	20.8	21.8	42.7
2.5	10.0	88.9	45.4	68.6	25.1	22.8	44.7
3.33	4.0	35.6	18.2	32.0	14.6	10.1	19.8
3.33	6.0	53.4	27.2	46.5	20.3	12.9	25.4
5.0	4.0	35.5	18.1	28.7	11.3	19.2	37.6
5.0	6.0	53.2	27.1	41.3	15.2	22.5	44.1
5.0	8.0	71.0	36.2	54.4	19.6	23.4	45.9
5.0	10.0	88.7	45.2	63.2	19.7	28.8	56.5
10.0	4.0	35.6	18.1	27.7	10.3	22.1	43.4
10.0	6.0	53.3	27.2	40.8	14.7	23.5	46.1
10.0	8.0	71.1	36.3	54.9	20.1	22.8	44.7
10.0	10.0	88.9	45.3	66.8	23.2	24.9	48.8
20.0	6.0	53.1	27.1	38.0	12.0	28.4	55.7
30.0	10.0	88.8	45.2	59.4	15.8	33.1	64.9

Figure 12. Effect of SRT on effluent solids characteristics in four completely mixed thermophilic semi-continuous reactors.

Accordingly, solids destruction efficiencies increased as the SRT for each reactor receiving a particular concentration of feed was lengthened, as seen in Figure 13. For most of the influent substrate conditions, the most rapid increase in solids destruction efficiency occurs at SRTs under 10 days; and beyond 10 days such efficiencies continue to rise, but at a much slower rate. It is somewhat disappointing to note that only about 65 percent of the biodegradable fraction is converted to gas even at a retention time of 30 days.

The data plotted in Figure 14 for reactors operating at 5-day and 10-day SRTs suggests a trend of increased solids destruction efficiency with increased solids loading. An increase in influent substrate concentrations also seems to cause a rise in the effluent or mixed liquor substrate levels as seen in Table 11 and in the plot of effluent biodegradable volatile solids versus influent biodegradable volatile solids for reactors operated at a 10-day HRT, shown in Figure 15.

A fairly good correlation between gas production predicted from volatile solids destruction and observed biogas evolution rates was observed among most fermentors operated at SRTs of 5 days or longer. On the other hand, fermentors operated at 3-1/3, 2-1/2, 2-1/2, and 5-day HRTs receiving feed concentrations of 4 percent, 8 percent, 10 percent, and 10 percent TS, respectively, produced gas at rates far lower than that predicted from the solids destruction data. From these four fermentors, the observed gas production rates were only 25 to 60 percent of the theoretical. These wide discrepancies may be due to several sources of error, including gas leaks from the fermentors, the exclusion of large organic particles in the sampling of partially-digested, coarse digester effluents, and the premature volatilization of intermediate organics, such as volatile acids, upon drying (103°C) prior to volatile solids analysis.

Dewaterability

The relative dewaterability of the manure was determined from the CST analysis. Although this procedure is not a direct measure of the specific resistance of a waste as would be obtained from the Buchner Funnel test,

Figure 13. Effect of SRT on volatile solids destruction efficiencies in the bench scale thermophilic semi-continuous fermentors.

Figure 14. Effect of influent volatile solids concentration on volatile solids removal efficiencies observed with bench scale thermophilic semi-continuous fermentors.

Figure 15. Effects of influent substrate concentrations on effluent substrate concentrations in a bench scale thermophilic semi-continuous fermentor operated at a 10 day HRT.

it can be used to show relative ease or difficulty of water separation from an organic slurry. The CST of the raw manure ranged from approximately 1,300 seconds with manure at 4 percent total solids to 7,000 seconds for raw manure at 10 percent total solids. Although thermophilic fermentation did have a beneficial effect of reducing the CST of this material, the resulting values were not indicative of a material which could be easily dewatered. At best, only a 50 percent reduction in CST was noticed, and usually the reductions were more on the order of 25 percent. Since CSTs of 40 to 50 seconds are required for efficient sand bed dewatering of sewage sludge and CSTs of 100 to 200 seconds would be required to dewater this material on mechanical devices, the anaerobically fermented manure will be difficult to dewater. The results of the CST analysis are presented in Tables 6 through 9 for all conditions at which this study was performed.

Ammonia Problems

Historically, ammonia toxicity has been a significant problem in anaerobic fermentors, and was predicted to be a problem particularly in fermentors receiving ammonia-rich wastes from swine and poultry (Jewell and Morris, 1974). It was for this reason that total and free ammonia concentrations in the feed and digester effluents were monitored throughout this study. Free ammonia (NH_3) concentrations were calculated from total ammonia ($\text{NH}_3 + \text{NH}_4^+$), pH, and temperature data, using the following expression (Anthonisen, 1974):

$$\text{FA} = \frac{17}{14} \frac{[\text{NH}_4\text{N}](10^{\text{pH}})}{K_d/K_w + 10^{\text{pH}}} \quad \text{Eq. 19}$$

where: NH_4N = effluent ammonia concentration, mg/l as N

FA = free ammonia concentration, mg/l as NH_3

$K_d/K_w = e^{6344/273 + ^\circ\text{C}}$

A summary of ammonia data and calculated free ammonia concentrations in reactor effluents is presented in Table 12 for all conditions investigated.

TABLE 12. SUMMARY OF EFFLUENT AMMONIA AND FREE AMMONIA
DATA FOR ALL STEADY STATE CONDITIONS.

Influent T.V.S., S_0 gm/l	SRT days	NH_4N mg/l	pH	Temp. °C	Free Ammonia* mg/l
35.5	3-1/3	335	6.2	61	3.6
35.5	5	314	6.8	61	13.1
35.5	10	574	7.3	60	66.9
53.3	3-1/3	393	6.4	61	6.7
53.3	5	410	7.0	62	28.0
53.3	10	863	7.4	59	117.5
53.3	20	364	7.0	60	22.3
71.1	2-1/2	505	6.9	60	24.9
71.1	5	605	7.1	60	46.1
71.1	10	1190	7.4	58	153.9
88.9	2-1/2	557	7.0	59	32.4
88.9	5	675	7.2	60	63.8
88.9	10	1474	7.5	59	245.5
88.9	30	560	7.2	58	47.6

* Calculated From:
$$\text{Free Ammonia} = \frac{17/14 [NH_4N] (10^{pH})}{K_d/K_w + 10^{pH}}$$

Free ammonia levels measured in the reactors at steady state varied widely from 3.6 mg/l to 245.5 mg/l. In general, for reactors operated at a particular HRT, effluent free ammonia concentrations were proportional to the influent solids concentration as seen in Figure 16 for digesters operating at 5 and 10-day HRTs.

The data in Tables 6 through 9 coupled with the free ammonia information in Table 12 yield little evidence of ammonia inhibition in the thermophilic semi-continuous fermentors at steady state. In the two reactors which did show symptoms of distress and failure (the bench units operated at a 3-1/3-day HRT, receiving 4 percent TS and 6 percent TS feed) the free ammonia concentrations were relatively low, below 10 mg/l. Conversely, the unit in which free ammonia was present in concentrations exceeding 245 mg/l (the bench scale unit operated at a 10-day HRT, 10 percent TS feed) had an inordinately high volatile acid content of nearly 4,600 mg/l as acetic. This reactor, however, could not be classified as inhibited, since it exhibited gas production rates and volatile solids destruction efficiencies greater than those obtained from the fermentors operated at a 10-day HRT and 4 percent, 6 percent, and 8 percent TS feed. The high pH (7.5) and bicarbonate alkalinity (greater than 10,000 mg/l as CaCO_3) fail also to corroborate a typical stressed reactor.

The only evidence that ammonia may have had an inhibitory effect was observed during a transition period in which the HRT of each reactor was reduced from 10 to 5 days. These data, including gas production, pH, ammonia, and free ammonia measurements, are tabulated in Tables A15 through A19 in Appendix A for fermentors receiving 4, 6, 8, and 10 percent TS manure feed. Ammonia, free ammonia, and pH measurements in the critical dairy cow manure feedstocks are also included in these tables.

According to the reactor performance data of Tables 6 through 9, steady state gas production rates for the 5-day HRT condition were ultimately higher than those obtained at an HRT of 10 days for reactors operating on the same feed concentrations. However, in a nonsteady state, transitory time frame, between January 20 and January 30, 1977, the semi-continuous reactors receiving 6, 8, and 10 percent TS feed experienced a drop in gas production, accompanied

Figure 16. Correlation between the influent volatile solids concentration and free ammonia in the thermophilic semi-continuous fermentors.

or, perhaps, effected by a dramatic rise in ammonia and free ammonia concentrations. Profiles of gas production, pH, and effluent free ammonia with time are depicted in Figure 17. It can be seen from these profiles that when reactor free ammonia increased to levels far above 150 mg/l, reductions of over 40 percent in the gas production rates were simultaneously observed.

DISCUSSION

The results of this study will be compared with similar studies fermenting dairy cow manure. This comparison will encompass other studies conducted at thermophilic temperatures as well as one study conducted mesophilically (Morris, 1976). In addition, the empirical mathematical model shown previously will be used to analyze the results.

Comparison to Literature

Table 2 presented the results from three studies which thermophilically fermented dairy cow manure at various organic loading rates. The results of these studies plus those from this study are compared graphically in Figure 18. This figure shows the effect the organic loading rate had on the efficiency of TVS destruction. Between the loading rate of 2.0 to 40.0 gm TVS/l-day this relationship was approximated by a straight line function. Initially, it should be noted that these data show a wide degree of scatter especially at the lower organic loading rates. This is substantiated by the fact that the regression coefficient for this plot was found to be only 0.32 (1.0 implies perfect fit). The data scatter can partially be attributed to differences in external factors (i.e. manure characteristics, feed rations, stage of lactation, and the time of the year at which the study was conducted) inherent between different studies. The poor regression coefficient obtained is due obviously in part to the data scatter and also to the fact that a straight line fit is being applied to a non linear function.

The data does show that the results of this study were found to compare favorably with those of other studies. This study was found to achieve comparable or greater TVS destruction for reactors operated at high organic

Figure 17. Profiles of gas production, effluent total and free ammonia, and pH during a period of suspected ammonia inhibition in four bench scale thermophilic semi-continuous fermentors.

Figure 18. Comparison of the effect of organic loading rate on TVS reduction efficiency in the thermophilic anaerobic fermentation of dairy cow manure.

loading rates (20 gm TVS/l-day) and equivalent or slightly less TVS destruction for those reactors operated at low organic loading rates. In addition, this study did achieve methane gas production rates at the high loadings which were comparable to data developed by Bryant et al., (1976).

Comparison Between Thermophilic and Mesophilic Fermentation

This thermophilic data and the mesophilic data from an earlier study (Morris, 1976) can be directly compared since all procedures and materials were similar. Both studies were operated at similar HRTs and influent substrate concentrations, and both studies obtained dairy cow manure from Cornell University's dairy research facility. The methods of manure collection and storage were identical for both studies. Many of the animals used and the environmental conditions were the same for both studies. As many operational practices as possible were standardized with the materials and methods so that differences in reactor performance between these two studies would be indicative of the effects of the temperature of the reaction rather than other external influences. However, the animals were being fed a different ration during the thermophilic study, and its composition was varied several times.

A summary of performance data from the thermophilic fermentations conducted in this study and from the mesophilic reactors operated by Morris are presented in Table 13. These data, including removal rates and removal efficiencies of biodegradable volatile solids (BVS), methane production per liter of reactor, and methane produced per gram of BVS added, will serve as a basis of comparison of fermentations conducted at mesophilic and thermophilic temperatures. A comparison of these two studies is shown in Figure 11.

Before comparing solids destruction rates and efficiencies between the mesophilic and thermophilic studies, it should be pointed out that the animal manure fed to the thermophilic fermentors had a higher biodegradable fraction (51 percent) than the degradable fraction determined for the manure feedstock used in the mesophilic study (42 percent). Caution should be exercised in comparing mesophilic and thermophilic reactors of virtually equal influent

TABLE 13. COMPARISON OF STEADY STATE DATA BETWEEN THIS STUDY OPERATED AT THERMOPHILIC TEMPERATURES AND DATA DEVELOPED BY MORRIS (1976) AT MESOPHILIC TEMPERATURES.

HRT days	This Study						Morris (1976)					
	S_o g BVS/l	S_{bo} g BVS/l	E_b^* %	$(S_{bo}-S_{bl})/t$ g BVS/l-day	λ CH ₄ l reactor-day	λ CH ₄ gm BVS added	S_o g BVS/l	S_{bo} g BVS/l	E_b^* %	$(S_{bo}-S_{bl})/t$ g BVS/l-day	λ CH ₄ l reactor-day	λ CH ₄ gm BVS added
2.5	71.1	36.3	42.7	15.5	1.588	0.109	69.8	29.7	20.2	6.0	0.182	0.015
2.5	88.9	45.3	44.7	20.3	2.603	0.143	--	--	--	--	--	--
3.33	35.6	18.2	19.8	3.6	0.160	0.029	--	--	--	--	--	--
3.33	53.4	27.2	25.4	6.9	0.333	0.041	--	--	--	--	--	--
5.0	35.5	18.1	37.6	6.8	0.589	0.162	34.9	14.8	54.7	8.1	0.385	0.130
5.0	53.2	27.1	44.1	11.9	0.978	0.180	52.3	22.3	49.1	10.9	0.488	0.110
5.0	71.0	36.2	45.9	16.6	1.367	0.189	69.8	29.7	45.1	13.4	0.813	0.137
5.0	88.7	45.2	56.5	25.5	1.715	0.190	87.2	37.1	44.2	16.4	1.100	0.148
10.0	35.6	18.6	43.4	7.9	0.460	0.254	34.9	14.9	82.4	12.2	0.616	0.416
10.0	53.3	27.2	45.1	12.5	0.698	0.257	52.3	22.3	79.3	17.6	0.827	0.372
10.0	71.1	36.3	44.7	16.2	0.744	0.205	69.8	29.7	75.1	22.3	0.933	0.314
10.0	88.9	45.3	48.8	22.1	1.027	0.227	87.2	37.1	68.2	25.3	1.233	0.332
20.0	--	--	--	--	--	--	34.9	14.9	89.2	13.2	0.405	0.547
20.0	53.1	27.1	55.7	15.1	0.589	0.435	52.3	22.3	87.2	19.4	0.546	0.492
20.0	--	--	--	--	--	--	69.8	29.7	85.2	25.3	0.673	0.453
20.0	--	--	--	--	--	--	87.2	37.1	82.5	30.6	0.773	0.417
30.0	--	--	--	--	--	--	34.9	14.9	93.9	13.9	0.273	0.554
30.0	--	--	--	--	--	--	52.3	22.3	88.7	19.7	0.379	0.513
30.0	--	--	--	--	--	--	69.8	29.7	84.5	25.1	0.458	0.463
30.0	88.9	45.3	64.9	29.4	1.234	0.410	87.2	37.1	88.4	32.8	0.575	0.465

* E_b = Biodegradable Volatile Solids (BVS) Destruction Efficiency

total volatile solids concentrations but of dissimilar biodegradable volatile solids loadings.

The data in Table 13 show no clear advantage of one fermentation temperature over another. Under completely unstressed, low-loading conditions at HRTs of 10 to 30 days, mesophilic digestion seems to result in higher BVS removal efficiencies and in a more stabilized effluent. With these retention times, mesophilic fermentors receiving feedstocks containing 14.9 to 37.1 g/l BVS experienced biodegradable substrate removal efficiencies between 68 and 94 percent, while the thermophilic reactors, receiving 20 percent higher biodegradable substrate loadings, performed at BVS removals of only 43.4 to 64.9 percent. Further comparisons of the daily amounts of BVS destroyed between mesophilic and thermophilic reactors of the same HRT and of comparable influent BVS concentrations also tend to support the hypothesis that at the longer retention times mesophilic temperatures have an advantage over thermophilic temperatures in producing a stable effluent.

At a 5-day HRT, however, the solids destruction advantage of mesophilic over thermophilic reactions begin to decrease, as shown in Table 13. Moreover, at the 2-1/2-day HRT conditions, two thermophilic reactors operating on 8 and 10 percent TS manure feed showed BVS removal efficiencies of about 43 percent, and BVS removal rates of 15.5 to 20.3 g/l/day, far exceeding the performance of the mesophilic reactor, operated at a 2-1/2-day HRT, 8 percent TS feed, which effected a BVS removal efficiency of only 20.2 percent and a BVS removal rate of 6.0 g/l/day.

When the gas production rates, expressed as liters of methane gas produced per liter of reactor per day, are compared for fermentors of equal solids loading between this study and the mesophilic experiments (Table 13), it is apparent that, at the short HRTs (2-1/2 and 5 days), the gas production rates are substantially higher at thermophilic temperatures. At HRTs of 10 days or more, methane production rates were, for the most part, equivalent or slightly higher for the mesophilic fermentations.

In summary, it would appear that at 10-days HRT or more, mesophilic fermentation is capable of a higher degree of BVS reduction than that observed with thermophilic fermentations. However, units operated at a short HRT of 2-1/2 days which received high influent concentrations of dairy cow manure feedstock, 8 to 10 percent TS, seemed to function with greater stability at thermophilic temperatures, thus performing at greater daily BVS removals and gas production rates than those observed in an equivalent mesophilic unit.

SUMMARY AND CONCLUSIONS

The thermophilic anaerobic semi-continuous reactors were operated for a total of 14 conditions. The majority of these conditions resulted in stable fermentations, consistent TVS reductions, and reliable methane gas production rates. For the range of hydraulic and organic loading conditions employed in this study, the following conclusions concerning the conventional thermophilic fermentation of dairy cow manure seem appropriate:

1. Effective stabilization of dairy cow manure can be accomplished by thermophilic anaerobic fermentation. Operated at influent total solids concentrations of 4 to 10 percent and at hydraulic retention times (HRT) of five days or greater, the thermophilic reactors of this study were capable of total volatile solids (TVS) destruction efficiencies of 19.1 to 33.1 percent and methane gas production rates of up to 1.72 ℓ/ℓ reactor/day.
2. The biodegradability studies showed that the refractory fraction of the dairy cow manure used in this study as a feedstock for the thermophilic was about 0.49.
3. Under special conditions, a single stage thermophilic anaerobic fermentor, receiving dairy cow manure as a feedstock, may operate at hydraulic retention times as low as 2-1/2 days while showing few signs of stress due to overloading. In this study, thermophilic reactors operated at a 2-1/2-day HRT and 8 and 10 percent TS were characterized by an accelerated but stable performance with methane gas production rates as high as 1.59 and 2.60 ℓ/ℓ reactor/day, respectively,

and with volatile solids destruction efficiencies of both reactors close to 22.0 percent.

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CHAPTER III

SIMPLIFIED FERMENTOR DEVELOPMENT

Earlier analyses of the feasibility of implementing anaerobic fermentation of agricultural residues indicated that the economics were not favorable for small scale agricultural operations (Jewell et al., 1976). This analysis attempted to utilize the lowest costs alternative materials and technology which were available in 1975. Because of the large potential for applications to small scale farming as described in the introduction section of this report, this negative conclusion should be accepted only if the critical controlling parameters cannot be modified to decrease the cost of small installations. About 45 percent of the total manpower of this study was devoted to examination of the potential of developing small, low-cost agricultural residue fermentors.

Some of the basic assumptions which guided the development of the simplified fermentor and the rationale used to develop these are summarized as follows:

1. The unit should be simple and self erectable, if possible--most of the units will have a total capital investment of less than \$50,000. Thus, large businesses will have difficulty in justifying activity in these areas. Skilled manpower is available at odd times of the years in many agricultural operations.
2. The unit must be constructed of low cost materials that are widely available.
3. The major restrictions identified as being required for "successful" anaerobic fermentation needed to be examined in light of a simple uniform agricultural residue feedstock such as dairy cow manure. Anaerobic digestion of sewage sludge requires careful control of pH, temperature, feed rates, mixing and solids concentration, among other important variables. This study focused on the need for external mixing and low solids concentration. Due to an abbreviated time-schedule temperature variations were examined in only a few instances.

Two basic design concepts were examined in order to define the potential of developing a simplified reactor: an unmixed horizontal gravity flow fermentor, similar to a "plug flow" design and semi-solid fermentation (i.e. fermentation with dry solids concentration significantly higher than 20 percent of the total weight).

UNMIXED HORIZONTAL DISPLACEMENT FERMENTATION--PLUG FLOW CONCEPT

The lowest cost reactor could be imagined as a trench using the soil for structural support and a low cost flexible cover. Omission of external mixing and the use of gravity to produce an overflow would result in a low cost design. Such a design would not be supported by the bulk of the background information available on anaerobic fermentation. For example, the lack of mixing with sewage sludge would encourage operational problems such as float formation or sedimentation of solids, and would possibly promote inefficient contact of the various groups of bacteria needed to interact to produce methane. Full scale displacement fermentors using a similar theory of design, have been operated in the early European facilities treating animal manures, but no operating data is available from these systems. The only full scale system located in the U.S. is being operated with beef manure at Custer, Michigan (Jewell et al., 1976).

The term "plug flow" in this study is used to describe an elongated reactor which is not mixed. Although it is recognized that gas bubble formation, feeding, and temperature stratification will cause some mixing, the unmixed horizontal displacement reactor will be referred to as the "plug flow" design.

In order to demonstrate the feasibility of using a plug flow design, certain system parameters must be defined, such as the relation of detention time to gas production and solids destruction, and the effect of minimal mixing on solids separation and total system operation. The specific objectives of this portion of the study were to:

1. Determine the feasibility of using a plug flow concept for dairy animal manure fermentation in small bench-scale reactors. With a

- larger pilot plant design to treat the residue produced by 5 cows, document the feasibility on a large scale;
2. Examine the lower limits of mixing and maximum solids concentrations that can be tolerated in a stable and efficient gas-producing fermentation system.

BACKGROUND

Two types of unmixed displacement fermentors have been used in the past in full scale operations. These are the fibrous material fermentor developed by Buswell and Boroff (1933) and the Darmstadt System (or Fermentation Canal Process) (Jewell et al., 1976). These systems were similar in that both operated with relatively dry solids concentrations and with very little mixing. Unfortunately, no operational data is available for either system.

Other full scale displacement reactors have been successfully operated with agricultural residues. Hutchinson (1972) operated a displacement reactor and several reactors in series with cattle wastes. The products of the systems, compost and liquid manure, were used to fertilize coffee fields in Kenya. The gas produced was utilized for cooking, water heating, and refrigeration.

A type of displacement reactor which was used for digestion of dairy residues was developed by Bell (1973). This was a vertical displacement anaerobic fermentor followed by a primary oxidation tank. It was observed that dewaterability increased and permanganate value (oxygen demand) decreased, while useful amounts of gas were obtained. The conclusion of this study was that dairy farm wastes could be efficiently digested under certain conditions.

A study by Rangasamy, et al., (1973) reported that the optimum detention time for an unmixed reactor operating on milk wastes was 6 days; the loading rate was 0.036 lb BOD/ft³/day. The gas production was measured at 12.3 to 13 cf/lb BOD destroyed. A parallel mixing study showed that a completely-mixed digester could accommodate nearly twice the maximum organic loading tolerated without stress by the unmixed fermentor.

There have been two full-scale demonstrations of anaerobic digestion in a horizontal plug flow mode. Fry (1974) reports operating such a unit on swine manure in South Africa. There is little quantitative data associated with this project. In the second case, a horizontal flow anaerobic digester has been constructed on a 350-head beef feedlot at Custer, Michigan (Jewell et al., 1976). This is the only full scale operation in the U.S. approximating a plug flow design of which the authors are aware. No performance data has been reported for this operation.

To maintain an anaerobic treatment system that will stabilize wastes efficiently, the acid formers and methane formers must be in a kinetically balanced state. Methane formers must metabolize volatile acids as fast as they are formed to avoid the toxic effects of high concentrations of the intermediate substrates and lowered pH levels. Unfortunately, the methane bacteria are more sensitive to modest changes in the environment, such as variations in temperature, pH, and the presence of toxic substances, than the acid formers (McCarty, 1964). If the methane formers alone are inhibited by unfavorable environmental conditions, a build-up of volatile acids and a change in alkalinity and pH will eventually occur, causing further inhibition of methanogenesis. In this manner, environmental conditions in a high rate single-compartment anaerobic digestion may deteriorate to the extent that fermentation is completely halted.

A two-stage (two-compartment) anaerobic fermentation process has been proposed (Ghosh, 1975; Gaddy, 1974; Pohland and Ghosh, 1971) in order to promote rapid reactions and to avoid the problems of uncontrolled acid intermediate exposure to methane formers and to maintain optimum environments for both acid and methane forming populations. In this process fermentable residues would first be introduced into a compartment where liquefaction and acid formation would predominate; the product intermediates of the first compartment would then be passed to a second compartment where conversion of the acids to methane would occur. Ghosh (1975) stated some of the benefits of two-stage digestion as: 1) providing an optimum environment for both acid forming and methanogenic bacterial populations; 2) allowing a reduction in reactor volume through increased process efficiency; and 3) yielding higher

fractions of methane in the digester gas. No full scale application of this two stage concept has been operated on agricultural residues.

A similar separation of environmental conditions as proposed in the two-stage reactors may also be achieved in plug flow reactors. In theory, a plug flow reactor operates in such a way that particles in suspension leave the reactor in the same sequence that they entered. In theory, axially uniform fluid segments flow through the system with no longitudinal back mixing. Each fluid segment may conceivably undergo rapid liquefaction and acid production predominantly in the front section of the reactor followed by methane formation throughout the reactor. Such a plugflow condition may be approximated by a number of completely mixed reactors arranged in series or by a long unmixed tubular reactor. Therefore, data developed in a unit consisting of a number of completely mixed reactors in series is often used to approximate plug flow operations that always exhibit some degree of mixing in practice.

Because of this simplicity, several plug flow fermentors have been built on a full scale, but little controlled data appears to be available. A study conducted previously at Cornell University provided a small amount of information regarding process performance with dairy manure (Jewell et al., 1976). In this study, two 20-liter reactors consisting of horizontal plexiglass tubes, were operated at 22.5°C, 8 percent total solids feed, and detention times of 15, 30, and 60 days. The preliminary data suggested that the absence of external mixing and reactor temperatures as low as 22.5°C could compete with the more conventional complex designs.

Theoretical Considerations

An idealized plug flow reactor is one in which the individual fluid particles of feed pass through the reaction vessel in the same sequence in which they entered (Metcalf and Eddy, Inc., 1972). This definition of the ideal plug flow condition implies the following characteristics of a "plug flow" reactor:

1. there is no intermixing or interaction between particles;

2. each increment of fluid passes through as an independent cross section or plug;
3. the velocity of the fluid throughout the reaction vessel, and at all points in any cross section of the stream, is constant;
4. each fluid particle is retained in the reaction vessel for a time equal to the theoretical retention time;
5. microbial activity, substrate concentration, temperature, or any other physical, chemical, or biological property must be constant for all points in any cross section of the stream.

Since no longitudinal mixing occurs in an ideal plug flow reactor each element of the mixture in such a system is analogous to a batch culture of microorganisms which is merely traveling through the reactor with time. Thus the substrate concentration in each segment decreases along the length of the reactor from the influent to the effluent, while the microorganism concentration theoretically increases as a result of substrate assimilation.

In practice, it is difficult to achieve an ideal plug flow condition, particularly in a liquid fermentor (Lawrence and McCarty, 1969). From the characteristics of the ideal plug flow reactor, it can be seen that any back mixing, settling, float formation, or short-circuiting observed to any extent in a biological reactor would preclude the establishment of idealized plug flow conditions.

Flow patterns through a reactor can be adequately described by an F curve. Suppose a given treatment system is given a one-time dosage of suspendable particles, such as a chemical tracer or a dye. If the cumulative percentage of tracer particles leaving the reactor is monitored and graphed over time, the resultant plot would be an F curve. Figure 19 illustrates F curves for completely mixed plug flow, and typical dispersed flow reactors (Gloyne and Eckenfelder, 1966). The F curves suggest patterns of substrate and microorganism movement through a liquid fermentor which may indicate the manner in which kinetic models should be applied in describing a non-ideal plug flow reactor. These theoretical patterns can be compared to the results of tracer studies developed from the pilot plug flow unit examined in this study.

Figure 19. Typical F-curves for completely mixed, ideal plug flow, and nonideal plug flow fermentors when a constant feed is initiated at zero time.

Kinetic Considerations

The basic equations commonly used in describing substrate removal in microbial culture may also be manipulated to model plug flow simulation systems. Lawrence and McCarty (1969), for instance, have developed a kinetic model for a plug flow system with recycle from Monod (1949) which relates the rate of substrate utilization both to the concentration of microorganisms in the reactor and to the concentration of substrate surrounding the organisms. An adaptation of this approach has been shown to be applicable to anaerobic systems operating on dairy cow manure (Jewell et al., 1976). The basic equation for the first order removal of substrate in a batch reactor is as follows:

$$\frac{dS_b}{dt} = -kS_{b1} \quad \text{Eq. 20}$$

where: S_{b1} = biodegradable substrate (BVS) concentration
 $\frac{dS_b}{dt}$ = rate of biodegradable substrate (BVS) removal
 k = first order decay rate coefficient
 t = time

In integrated form Equation [20] becomes:

$$S_{b1} = S_{b0}e^{-kt} \quad \text{Eq. 21}$$

where: S_{b0} = the initial concentration of biodegradable substrate.

As previously mentioned, each element of fluid in an idealized plug flow reactor is analogous to a batch unit traveling through the system with time. It follows that Equation 21 can also be used to describe substrate removal in an idealized plug flow system conceptually shown in Figure 20.

The performance of plug flow reactors can be approximated by several completely stirred tank reactors (CSTRs) connected in series as depicted in

A. PLUG FLOW

B. A SERIES OF FOUR CSTRs

Figure 20. Schematic comparison of an idealized plug flow reactor and a series of completely stirred tank reactors (CSTRs).

Figure 20. The equation for CSTR performance when a first order decay function involved is:

$$S_{b1} = \frac{S_{bo}}{1 + k(\text{HRT})} \quad \text{Eq. 22}$$

If several (n) CSTRs are arranged in series, as illustrated in Figure 20, the effluent BVS concentration, S_{bn} may be predicted by the equation:

$$S_{bn} = \frac{S_{bo}}{\left[1 + \frac{k(\text{HRT})}{n}\right]^n} \quad \text{Eq. 23}$$

The HRT term in Equation 23 refers to the hydraulic retention time of the combined reactor series.

If a fermentor system is comprised of an infinite number of CSTRs placed in series, the following simplification in Equation 23 may be made:

$$\lim_{n \rightarrow \infty} \left[1 + \frac{k(\text{HRT})}{n}\right]^n = e^{k(\text{HRT})} \quad \text{Eq. 24}$$

and

$$S_{bn} = S_{bo} e^{-k(\text{HRT})} \quad \text{Eq. 25}$$

This is the same result obtained for a batch reactor Equation 21, or for an idealized plug flow system. From this development it can be seen that with respect to substrate removal, an infinite series of CSTRs is an exact mathematical equivalent to idealized plug flow. A series of three or four CSTRs may be an approximation to a plug flow reactor.

The first task in the kinetic evaluation of the plug flow reactor is to calculate the first order substrate removal coefficient, k . This parameter

is most easily obtained from the semi-continuously fed fermentors which were operated concurrently as controls for the plug flow fermentors. If a number of these control reactors have been operated under varied HRT or influent substrate conditions, k may be calculated in the manner described in Chapter 6 by measuring the slope of a plot of $(S_{bo} - S_{bl})/\text{HRT}$ versus the effluent BVS concentration, S_{bl} . However, if not enough data are available for a graphical determination, k may be approximated directly from continuous, semi-continuous, or batch control reactor data using equations derived from first order kinetics and mass balance considerations. These equations have been developed in the appendix and may be stated as follows:

$$\text{Continuous (CSTR)} \quad k = \frac{S_{bo} - S_{bl}}{S_{bl}(\text{HRT})} \quad \text{Eq. 26}$$

$$\text{Semi-continuous} \quad k = t_i \ln \left[1 + \frac{t_i(S_{bo} - S_{bl})}{S_{bl}(\text{HRT})} \right] \quad \text{Eq. 27}$$

$$\text{Batch} \quad k = (\text{HRT})^{-1} \ln \frac{S_{bo}}{S_{bl}} \quad \text{Eq. 28}$$

where: t_i = incremental time period in semi-continuous mode (days)

Once the first order substrate removal coefficient has been determined, Equation 23 may be used to determine the number of theoretical CSTRs in series required to produce an effluent BVS concentration comparable to that observed for a given plug flow reactor. In theory, the higher the number of CSTRs in series required for comparable performance, the closer the plug flow reactor may be considered to simulate the plug flow condition.

MATERIALS AND METHODS

All bench scale experiments described in this section were conducted at the Agricultural Waste Management Laboratory at Cornell University, Ithaca,

New York. The pilot scale study was carried out at the full scale 500 cow dairy Teaching and Research Center, a Cornell University facility located at Harford, New York.

The bench-scale plug flow experiments were conducted using two different types of reactor models. The initial experiments were run with three unmixed horizontal displacement (UHD) reactors; these were 5 foot long, 6 inch diameter plexiglass columns allowing uniform flow down the entire length. The second type of model consisted of four completely-mixed reactors operating in series. Most of the bench scale study was completed with the use of the second type of model. Larger-scale plug flow experiments were carried out with a pilot UHD fermentor.

All dairy animal manure feed was obtained from the same source as described in Chapter 2.

Bench-Scale Unmixed Horizontal Displacement Reactors

The size of the small scale reactors was 20 liters with gas collection space above the liquid level. The reactors were inclined at an approximate 10 degree angle to facilitate the flow of substrate through the reactor. A schematic diagram of the reactor system may be seen in Figure 21. Feeding was accomplished by opening a port near the top of the upper inclined side and pouring the prepared feed mixture into the hole. To remove the effluent, a one-inch diameter port, located on the bottom of the cylinder, was opened. The hydraulic detention time of the system was determined by the volume of manure wasted and fed each day.

The gas produced within the reactor was drawn off to a wet test meter for volumetric measurement. Located between the meter and reactor was a water vapor trap to prevent condensation inside the meter. After passing through the meter, the digester gas was vented to the atmosphere.

Temperature for the bench-scale plug flow fermentors was maintained at $35 \pm 1^{\circ}\text{C}$ in a constant temperature room. Temperature was maintained by a Fenwal brand temperature controller and sensor probes.

Figure 21. Schematic diagram of the unmixed horizontal displacement or plug flow bench scale apparatus.

Bench-Scale Completely Mixed Reactors In Series

Due to operating problems with the small plug flow reactors, which will be discussed in the results section, it was necessary to design an alternative, yet comparable, experimental model. The design chosen was a reactor system consisting of four four-liter glass bottles connected to each other by tubing as shown in Figure 22. Each reactor bottle had a reactor liquid volume of three liters and was connected in series to the adjacent units by a one-inch diameter tube fitted onto glass nipples. The first reactor chamber in the series had a one-inch diameter tube near the top of the bottle for the purpose of feeding the system. Figures 23 and 24 show the 20°C and 35°C apparatus, respectively.

Wasting and feeding was accomplished in a stepwise fashion. First, the one-inch diameter tube serving as the effluent port for reactor four (Figure 22) was opened and the proper volume of manure was withdrawn. After wasting reactor four, the interconnecting line from reactor three to four was opened. Reactor three was then pressurized by manipulation of the gas collection bottles and the manure flowed into reactor four until the proper level was regained. This procedure was repeated until reactor one had been wasted into reactor two. Then the feed port was opened and the feed mixture was added. Because of the interconnecting tubes, the system was never exposed to oxygen except when the first reactor was opened for feeding. The units were manually shaken twice per day to simulate four CSTRs in series.

Volumes of gas produced in each of the CSTRs in the reactor series were individually collected and measured for these units by a system of liquid displacement bottles. Inside each displacement bottle was a brine consisting of salt-saturated 15 percent H_2SO_4 . The purpose of the acid brine was to prevent carbon dioxide from dissolving into the liquid and causing an error in gas measurement. The gas displaced the methyl orange-colored brine into a 20-liter reservoir by siphon action, and the corresponding volume of displaced brine could be easily measured. Each day the measuring bottles would be re-zeroed by reverse siphoning the brine out of the reservoir.

Constant temperature was maintained for three sets of four CSTRs in series. One system was placed on a laboratory bench exposed to the ambient room

Figure 22. Schematic of a system of four CSTR units in series with associated gas collection system.

Figure 23. Four CSTR fermentors in series used to simulate plug flow conditions at 22°C.

Figure 24. Four CSTR fermentors in series used to simulate plug flow conditions at 35°C.

temperature (22°C), one system was located in the mesophilic constant temperature room (35 ± 1°C) and the high temperature system was incubated in a small thermophilic constant temperature chamber (60 ± 1°C).

Pilot Scale Unmixed Horizontal Displacement Reactor - Plug Flow Reactor

The pilot-scale UHD reactor consisted of an enclosed, impermeable membrane (Hypalon brand) lined, plywood tank with dimensions of 20 feet long by 3 feet wide by 4 feet high (3 foot liquid depth). This unit was designed to handle manure from three dairy cows when operated at a 30-day hydraulic retention time (HRT) and 8% TS feed; the reactor had a liquid volume capacity of 1500 gallons. A schematic of the pilot fermentor is shown in Figure 25. Figures 26 to 30 show photographs of the unit under construction and the finished unit. The baffles at the influent and effluent ends of the digester, were positioned one foot inside the end plates and extended two feet into the liquid. A 1.0-inch layer of styrofoam was used to provide insulation. The basin's walls and floor were then lined with one solid piece of Hypalon (Dupont) liner, extending the entire length of the digester. The top section of the liner was sealed to the side edges of the lower liner section and folded over and attached to each of the end baffles below liquid level, thus forming a gas-tight bag above the reactor basin.

Three one-foot wide by three feet high, one-inch thick plexiglass windows were located on one side, each providing access to a third of the fermentor. The windows into the pilot unit were fitted with discharge lines, hot water lines, and temperature control sensors. Also, placed in each window were three sampling tubes and a four inch capped hole. The sampling tubes consisted of two-inch PVC pipes, located at three different heights, mounted in each window, which could be slid in or out through a plexiglass sleeve with a maximum reach of 2 1/2 feet into the digester, thus providing the ability to take samples at three different points lengthwise, at three different heights, and at various points across the width of the reactor. When not in use, the tubes were drawn outside of the reactor as far as possible, to avoid interference with the flow of the manure. Variations in the thickness of the floating crust layer were observed through the windows.(see Figure 27).

Figure 25. Schematic of the 5000 liter pilot scale plug flow reactor.

Figure 25. Photograph of the pilot scale plug flow fermentor located at the Research Laboratory of the Cornell University Agricultural Sciences Teaching and Research Farm, Harford, New York.

Figure 27. Photograph of one of three sampling windows located on the pilot scale plug flow fermentor.

Figure 28. Photograph of the interior structure of the pilot scale plug flow fermentor.

Figure 29. Photograph of the pilot scale plug flow fermentor feeding system.

Figure 30. Photograph of the effluent end of the pilot scale plug flow fermentor.

The plug flow reactor was maintained at a temperature of 35°C by circulating hot water through three independent networks of 1/2 inch PVC pipe, each servicing a third of the reactor volume. Each network of pipes consisted of four five-foot long pipes, spaced 6 inches apart, six inches above the reactor floor. Water flow through the networks was controlled by solenoid switches, activated by temperature sensors located above each of the three sets of heating pipes. A standard 30 gallon water heater was used to heat the water. Dial thermometers at each window indicated the temperature of the reactor contents at various locations. The room and effluent temperatures were recorded daily. Electric consumption by the water heater was metered.

The flow of gas generated by the UHD reactor was measured with two wet test meters. The gas passing through these meters was piped directly outside the building and released to the atmosphere.

The pilot-scale fermentor required handling such large volumes of manure that the feeding procedure had to be mechanized as much as possible. A 360-gallon, trailer-mounted, feed tank was loaded with the manure feedstock and was transported to the reactor. The manure was diluted to the desired total solids concentration and mechanically mixed for 15 minutes prior to feeding. The manure slurry was then transferred via a Moyno pump to the influent end of the reactor on a once-a-day basis. As the manure feed was pumped into the baffled influent end of the UHD reactor an equal volume of effluent was displaced and overflowed from the effluent end into a 400-gallon holding tank which was periodically emptied. The pilot-scale reactor, when operated at a 30-day HRT, required 50 gallons of feed per day. At this operating condition the feed and effluent holding tanks had a 7-day holding capacity.

Semi-Continuous Completely-Mixed Control Reactors

These reactors were designed to be operated at identical conditions as the plug flow reactor and the reactors in series. In this way, they would serve as controls for comparison purposes. These reactors consisted of four liter bottles with a three-liter liquid volume.

A one-inch diameter glass tube was located near the top of the bottle for feeding purposes. This feed port was sealed when not in use. Wasting of these reactors was accomplished through a one-inch diameter glass tube located at the bottom of the reactor bottle. Each reactor was manually agitated twice per day. The gas collection system consisted of liquid displacement bottles as described previously.

Operating Modes of Bench Scale Reactors

The operating parameters chosen for investigation in the bench scale studies were intended to provide a range of performance data useful for kinetic analysis and full scale design. The first set of experiments were conducted with the three plexiglass column reactors. These were operated at mesophilic (35°C) temperatures at hydraulic retention times of 10, 20, and 30 days, respectively. All bench scale experiments were run at a feed concentration of 8 percent total solids.

After it was determined that the reactor design used was inappropriate for short detention time studies (i.e., below 20 days) due to solids accumulation problems, the model was changed to a system of completely mixed reactors in series to complete the investigation. Unless stated otherwise, the HRT referred to with the series reactors will be the total retention period of all 4 units. The 35°C phase of the study was finished by a sequence of tests at low detention times, specifically 12 and 10 days HRT. The effect of recycle was investigated by operating the series reactors at 10 and 4 days HRT (1 day in each of the 4 reactors) with an approximate 20 percent recycle of the effluent into reactor one.

Since plug flow experiments had been performed at 22°C in a previous Cornell study (Jewell, 1976), a series of CSTRs was also operated at this temperature. The detention times studied were 12 and 30 days.

Operating Modes of the Pilot Scale Plug Flow Reactor

The first condition investigated in the pilot scale unit was intended to duplicate the same set of operating conditions as examined in a bench scale

experiment, to observe scale-up differences. The operating mode which appeared to offer the highest potential for stable operation was 35°C, 8 percent total solids feed, and a 30-day HRT.

Following this condition, the feed concentration was increased to an "as received" 10 to 13 percent total solids, while the temperature and detention time were unchanged. The purpose of this was to develop data at the solids concentration of animal waste without using dilution. The different reactor designs examined and the length of the study is shown in Table 14.

A summary of all operating conditions investigated in the plug flow study is presented in Table 15. This portion of the study is being continued and expanded to large scale 65 cow experimental units.

Sampling and Testing Program: Bench Scale Reactors

All of the reactor systems used in this study were wasted and fed on a once-daily basis. Raw manure was diluted on a volumetric basis to achieve the desired percent of dry solids concentration. The effluent pH from each system was measured daily. Also recorded were daily gas production volumes; for the series reactor experiments, individual gas productions were recorded for each stage. Prior to attainment of steady state conditions, weekly analyses were performed for total and volatile solids. When steady state conditions were observed, the following analyses were performed on the effluent: total solids, total volatile solids, pH, alkalinity, volatile acids, total kjeldahl nitrogen, ammonia nitrogen, and capillary suction time (a measure of dewaterability). When steady-state analyses were performed on the series systems, each stage was separately analyzed for total solids, total volatile solids, pH, and volatile acids.

The criteria used in this study for identifying the steady state condition were identical to those discussed in Chapter 2 for the operation of the conventional control reactors. In the plug flow simulation reactors, the steady state condition was evaluated primarily on the basis of stabilized gas production and volatile solids destruction efficiency. The understressed steady

TABLE 14. SUMMARY OF NUMBER AND TYPES OF REACTORS EXAMINED IN PLUG FLOW STUDIES AND THE DURATION OF EXPERIMENTS.

Reactor Type	Reactor Size (liters)	Numbers of Units Studied	Duration of Experiment (days)	Major Parameters Examined
Bench Scale Plug Flow	20	3	130	HRT
Semi-Completely Mixed* Reactors in Series	12	12	150	Temp., HRT, recycle
Semi-Completely Mixed Control Reactor for Bench Scale Studies	3	2	150	Temp., HRT
Completely Mixed Control for Pilot Scale Plug Flow	3	1	240	Solids, HRT
Pilot Scale Plug Flow	5565	1	240	Solids, HRT

TABLE 15. SUMMARY OF OPERATING CONDITIONS IN THE UNMIXED HORIZONTAL FLOW OR PLUG FLOW REACTOR.

Reactor Type	Temperature °C	Influent Total Solids S_o gm/l	HRT days	Total Solid Loading Rate $\frac{lb}{ft^3 \cdot day}$ $kg/m^3 \cdot day$	
Bench-scale plug flow	35	80.0	10	0.50	8.00
	35	80.0	20	0.25	4.00
	35	80.0	30	0.17	2.67
Bench-scale, 4 CSTRs in Series	22	80.0	12	0.42	6.67
	22	80.0	30	0.17	2.67
	35	80.0	12	0.42	6.67
	35	80.0	10	0.50	8.00
	35	80.0	10*	0.50	8.00 (w/o Recycled Solids)
	35	80.0	4**	1.25	20.00 "
	60	80.0	12	0.42	6.67
	60	80.0	4	1.25	20.00
Pilot-scale plug flow	35	80.0	30	0.17	2.67
	35	112.0	30	0.23	2.67

* 40% Volumetric Recycle Applied

** 20% Volumetric Recycle Applied

state operation of these fermentors was typically characterized by near neutral pH's (6.9 - 7.2), high alkalinities (3.0-6.0 g/l as CaCO_3), a high methane fraction in the gas (50%) and non-inhibitory levels of volatile acids (<5000 mg/l as acetic). A steady state condition in a bench reactor was identified only after the system had been operated for two or three hydraulic retention times. These units were then operated for at least one additional HRT at steady state pending sampling and analysis.

Each new feed batch was subjected to a complete analysis at the time of collection immediately after obtaining the feed material. A summary of the complete testing program is presented in Table 16.

Sampling and Testing Program: Pilot Scale Reactor

This system was analyzed according to the program indicated in Table 17. Free ammonia nitrogen was calculated, and the elapsed time between new feed batches was recorded. The manure in the feed tank was never more than six days old.

Twice weekly, the gas was analyzed for methane and carbon dioxide content. The gas was collected in a closed gas sample tube, and analyzed on a Gow Mac Model 550 gas chromatograph. During steady state conditions, the only additional test performed was for volatile acids.

In general, at least one or two hydraulic retention time (HRT) periods of operation were required to bring the large pilot-scale reactor to steady state conditions. Data collection continued through the steady state for one or more HRT periods before changing over to a new set of operating conditions.

In addition to effluent sampling, core samples were taken from various points in the unit using the movable, two-inch sampling tubes. The tubes were positioned approximately half way into the reactor and flushed to remove residual manure from the tube; samples were then taken from each of the tubes. The samples were analyzed for total solids, volatile solids and pH to establish a profile in solids destruction and pH as the material moved down the length

TABLE 16. ANALYSIS PROGRAM AND TESTING SCHEDULE FOR BENCH SCALE PLUG FLOW STUDY.

Parameter	Influent	Effluent		
		Daily	Weekly	Steady state
Total Solids	X		X	X
Total Volatile Solids	X		X	X
pH	X	X	X	X
NH ₄ -N	X			X
TKN	X			X
Volatile Acids				X
Alkalinity	X			X
CST	X			X
Temperature		X	X	X
Total Gas Production		X	X	X
Gas Composition				X

TABLE 17. ANALYSIS PROGRAM AND TESTING SCHEDULE FOR PILOT SCALE PLUG FLOW STUDY.

Parameter	Influent	Effluent			
		Daily	Twice/Week	Once/Week	Steady State
Total Solids	X		X	X	X
Total Volatile Solids	X		X	X	X
pH	X	X	X	X	X
NH ₄ -N	X		X		X
TKN	X		X		X
Volatile Acids				X	X
Alkalinity	X		X		X
CST	X				X
Temperature Profile		X	X		X
Ambient Temperature		X	X		X
Crust Layer Thickness		X	X		X
Electric Consumption		X	X		X
Total Gas Production		X	X		X
Gas Composition			X		X
Gas Pressure		X	X		X
Internal Profile Analysis				X	X

of the reactor. While the pilot-scale unit was operating on 8 percent TS, manure samples were taken from only the bottom two sampling tubes of each window, since the top most tubes extended into a thick float material which would not flow through the two inch tube. During the first loading condition, the caps were removed from the four inch sampling ports and float samples were taken. Total and volatile solids, and pH of the float samples were determined. At the 10 to 12 percent total solids loading, the crust was gradually broken apart and pushed out; liquid samples could then be drawn from all of the sampling tubes.

Tracer Studies--Liquid and Solids Retention Time Measurements

To determine the retention times of the solids and liquids in the pilot scale reactor chemical tracers were introduced and monitored in the effluent samples. The detailed procedure is described in the Appendix. To follow the path of solid-phase materials through the digester, a chromium-tagged fiber was used. This was prepared by soaking washed beef manure fiber in a saturated dichromate solution under heat, as described by Uden (1977). After the dichromate treatment, the fiber was washed and soaked in an ascorbic acid reducing solution and washed again. The fiber was then air dried and assayed for chromium. Subsequent to pulse feeding the reactor with a certain amount of the chromium-tagged fiber, chromium was monitored in the effluent with time.

The liquid retention time was determined with soluble salt tracers of cobalt lithium ethylenediaminetetraacetate ($\text{CoLiEDTA-3H}_2\text{O}$) and sodium chloride (NaCl). Effluent samples were then analyzed for cobalt and chloride.

Hydraulic and solids retention times were then estimated on the basis of concentration profiles of the effluent. Curves for chromium indicated a solids retention time and curves for cobalt and chloride yielded an effective liquid retention time.

Biodegradability of Dairy Cow Manure Feedstock

The biodegradable volatile solids (BVS) fraction of the manure feedstocks

used in this study was determined by the same general procedure described in the Methods and Materials section of Chapter 2, Conventional Controlled Fermentation Section.

Analytical Techniques for Fermentor Effluent Analysis

The procedures and techniques used for determining total solids, total volatile solids, pH, and alkalinity were taken from Standard Methods (1971). Ammonia nitrogen and total kjeldahl nitrogen were determined by the methods of MacKenzie and Wallace (1954) and the capillary suction times of the effluents were measured by a technique described by Baskerville and Gale (1968).

Gas composition was measured using Gow Mac Model 550 thermal conductivity gas chromatograph with Porapak Q column packing. Volatile acid concentrations were also determined by gas chromatography on a Gow Mac Model 750 flame ionization detector using Chromasorb 101 column packing.

Chlorides, monitored in the pilot-scale plug flow reactor as a tracer, were analyzed in accordance with the argentometric described in Standard Methods (1971). Chromium and cobalt, also introduced into the plug flow reactor as tracers, were measured by atomic absorption spectroscopy in effluent samples previously refluxed in nitric acid (Hayes and Theis, 1976) as outlined in the appendix.

RESULTS

Biodegradability of Dairy Cow Manure

The total fraction of organics which are biodegradable indicates the maximum methane generation possible. Organics which are resistant to degradation are referred to as refractory organics. The refractory volatile solids concentration was determined from data generated from the batch digestion of 8 percent TS dairy cow manure which yielded a plot of S_1/S_0 versus $(S_0 - \text{SRT})^{-1}$ where S_1 was the volatile solids concentration of effluent samples taken from the batch reactor at various solids retention times (SRT). The intercept

of this plot, shown in Figure 31, is a measure of the refractory fraction, R. In this study, the value of R was determined as 0.56.

Thus, the biodegradable fraction of the manure was observed to be 44 percent of the influent volatile solids concentration. The high refractory fraction is most likely related to the high percentage of indigestible plant cell wall constituents present in the manure; also of importance is the presence of intricate polysaccharide-lignin complexes which degrade more slowly than many other organics (Goehring and VanSoest, 1970).

The total biodegradable fraction can be converted to maximum methane yield potential since it is known that about 8 ft³ of methane is formed for every pound of complex organics converted to methane. Theoretically, the maximum gas production that could be produced from this dairy cow manure at 35°C would be about 7 ft³ per pound of volatile solids fed or about 3.5 ft³ of methane per pound fed.

The value for R of 0.56 is in good agreement with other values for anaerobic studies on dairy manure conducted at temperatures around 35°C (Jewell et al., 1976; Bryant et al., 1976). The ranges of values reported varied from 0.430 to 0.630 as the refractory fraction of the dairy manure.

Unmixed Horizontal Displacement Reactors

Three 20-liter plexiglass columns were operated at mesophilic (35°C) temperatures with detention times of 10, 20, and 30 days. The influent total solids concentration was maintained at about 80 gm/l by the appropriate dilution of the raw dairy manure.

The start-up procedure for these fermentors consisted of placing manure and digested sewage sludge (as seed) in the reactors, initially feeding the reactors at a reduced solids loading and gradually increasing the hydraulic and solids loading during the start-up period. The 20-day and 30-day reactors experienced unstable conditions during start-up, but a careful recovery program allowed them to regain stability. The specific start-up programs for each reactor are listed in Appendix 3.

Figure 31. Determination of the refractory fraction of dairy cow manure with 8 percent TS feed in a completely mixed batch reactor at 35°C.

Gas production and pH data taken from the 20 and 30 day bench-scale plug flow fermentors are presented in Figures 32 and 33, respectively. At the end of the first HRT in the 20-day and 30-day reactors, each was producing about 0.20 $\times$ gas/ $\times$ reactor/day. This low gas production rate indicated that the microbial system had not been established. At the 30 to 40 day period both units began producing gas at much higher rates reaching maximum values of nearly 2 volumes of gas per reactor volume per day, for the 20 and 30 day reactors, respectively.

Although operation of these small six inch diameter units at the longer retention periods resulted in stable operation, solids accumulation presented considerable operational problems with detention time around 10 days. The reactor receiving the highest solids loading, the 10 day HRT reactor (8% TS feed), exhibited plugging at the effluent port within 30 days of start-up. This made it difficult to obtain accurate data on solids destruction efficiencies. Consequently, the 10-day HRT unit was terminated, after 75 days of operation, before steady state data could be obtained. Similarly, an increasing float solids layer observed in the 20 day and 30 day HRT reactors led to suspicions that solids were also accumulating in these units, though at a slower rate. In an attempt to measure the solids retained by the reactors and to allow removal of some of the excess solids through effluent wastage, the reactors were manually shaken on the 94th day of operation, resulting in a resuspension of the float and settled solids. A sharp decrease in gas production was observed in both reactors within 48 hours following remixing, as seen in Figures 32 and 33. After more than 120 days of operation, the gas production rate had not regained the higher rate observed prior to day 94. Approximation to steady state after 120 days of operation resulted in gas yields of 1.3 volumes per reactor volume per day and 1.1 vol/vol-day for the 20 and 30 day HRT reactors, respectively. These values correspond to gas production of 5.4 ft³/lb TS fed and 6.5 ft³/lb. TS added for the 20 and 30 day HRT reactors respectively. It appeared that in both cases good conversion efficiency of the biodegradable fraction was being achieved.

The effluent characteristics of the bench scale plug flow reactors during plateau periods of elevated gas production are summarized in Table 18. Also

Figure 32. Gas production and pH variations from a 20 day HRT bench-scale plug flow reactor.

Figure 33. Gas production and pH variations from a 30 day HRT bench scale plug flow reactor.

TABLE 18. COMPARISON OF DATA FROM THE BENCH SCALE PLUG FLOW (35°C) FERMENTORS OF THE STUDY TO THE COMPLETELY MIXED FERMENTORS OPERATED IN AN EARLIER STUDY ON DAIRY COW MANURE (Jewell *et al.*, 1976).

	HRT = 20 days		HRT = 30 days		
	Completely Mixed *	Plug Flow**	Completely Mixed*	Completely Mixed**	Plug Flow**
Influent Total Solids, g/l	80.0	80.0	80.0	86.1	80.0
Influent Total Volatile Solids, g/l	69.7	70.7	69.8	74.6	70.7
Effluent Total Solids gm/l	54.1	52.0	53.9	51.7	58.0
% red	32.4	35.0	32.6	40.0	27.5
Effluent Total Volatile Solids gm/l	44.5	43.1	44.7	41.4	48.4
% red	36.2	39.2	36.0	44.5	31.7
Alkalinity (gm/l)	10.4	10.7	11.5	8.6	10.8
TKN (gm/l)	3.31	2.86	3.53	-----	2.91
NH ₃ -N (gm/l)	1.54	1.40	1.76	2.1	1.55
Org. N (gm/l)	1.77	1.46	1.77	-----	1.36
Gas Production ***					
l/l reactor/day	1.20	1.56	0.81	.84	1.38
l/gm VS _A	0.344	0.441	0.350	.338	.585
l/gm VS _D	0.950	1.13	0.972	.759	1.23
Gas Composition					
CH ₄ /CO ₂	62/67	57/43	62/37	54/46	62/37
lCH ₄ /gmVS _D	0.589	0.644	0.603	.410	0.701
lCH ₄ /gmVS _A	0.213	0.251	0.217	.183	0.334

*Jewell *et al.*, (1976)

**This study using tube reactors

***All gas production data corrected to STP

included in this table, for comparison, are data from the bench scale, completely mixed reactors, operated under similar conditions (8% TS; 20 and 30 day HRT; 32.5°C) obtained from a previous study with dairy cow manure (Jewell, et al., 1976). The data presented in Table 18 indicate gas production capability of plug flow reactors to exceed that of completely mixed reactors. This data is not conclusive since the two studies were not conducted in parallel with the same substrate. Also, solids accumulation occurred in the plug flow models as is reflected by the values of gas produced per gram of VS destroyed. The theoretical value would be about 0.9 liters of gas (60% methane) per gm of VS destroyed, but the measured values for the plug flow units were higher indicating that solids were accumulating within reactors.

Completely Mixed Reactors in Series

The experiments with a bench scale system of four CSTRs in series were conducted at 22°C and 35°C. Some data was also obtained at 60°C. These reactor systems operated without problems of solids accumulation which was evident with the small plug flow reactor models. Start-up problems were not encountered with the series reactors even though reactor retention periods of 2 or 3 days were used. Table 19 summarizes the effluent characteristics and total system performance for all conditions investigated in this phase of the study. These data show that successful operation of series reactors is feasible even at individual reactor retention periods as short as one or two days. Of course effluent recycle must be utilized to seed the effluent with an established microbial populations at these short HRT's.

Appendix Tables B5 to B9 contain a complete summary of analysis for each condition investigated. Also included in each summary table is the corresponding control data for each specific operating condition. The effluent ammonia concentration in all series experiments was approximately twice the influent concentration. At no point did the units exhibit inhibitory characteristics due to ammonia or free ammonia concentrations. It may also be noted that fermentation did not significantly improve the dewaterability of dairy manure as seen in comparing CST data for effluent with that determined for raw feed.

TABLE 19. SUMMARY OF 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 —— (recycle) ———	HRT = 10 days	HRT = 12 days
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
gm/l reduced	10.13	20.91	20.40	20.89	20.66
% 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	50.53
gm/l reduced	9.83	20.17	19.63	20.92	20.49
% reduced	13.7	28.3	27.9	29.1	28.9
Total Gas Production l/day	3.21	4.41	15.58	15.82	16.40
l/l reactor-day	0.268	0.368	1.298	1.318	1.367
Methane Production Rates l/day	-----	-----	6.51	8.17	8.53
l/l reactor-day	-----	-----	0.543	0.681	0.711
l gas/gm VS_A	0.044	0.155	0.185	0.184	0.231
l gas/gm VS_D	0.327	0.547	0.661	0.631	0.800
l CH_4 /gm VS_d	-----	-----	0.276	0.325	0.416

Individual results for steady-state gas production and gas composition by reactor stage are summarized in Table 20; Appendix B10 shows the steady-state solids concentrations observed in each stage. Figures 34 through 38 illustrate the gas production history of all the CSTRs-in-series experiments. The effects of the various operating modes on the 35°C experiments are illustrated in gas production profiles for the entire reactor and in each of the four reactor stages as shown in Figures 34 and 35. These data indicate a slight decrease in gas production in the course of a change-over from a 12-day to a 10-day HRT at 35°C. There was no significant increase in gas production when recycling was initiated at a rate of 500 ml per day (0.4 Q per day) during the 10-day condition. Table 20 shows that recycling effected a small increase in gas production in the first two stages which was countered by a drop in gas production in the final two stages; the quality of the gas did improve noticeably, however. The table also indicates a general trend of increased methane percentage in each successive stage.

Volatile acids data are presented in Table 21. This information represents the total volatile acid concentration in each of the four stages of the series experiments, expressed in mg/l as acetic acid. Appendix B14 lists the breakdown of volatile acid data into each of the four individually measured acids, i.e., acetic, propionic, butyric, and valeric.

Gas production and solids destruction in comparable plug flow experiments are provided in Figures 39 and 40 for the 22°C bench scale experiments. It may be observed that there was good agreement between the performance characteristics of each model.

Similar illustrations are provided in Figures 41 and 42 for the 35°C bench-scale experiments. Also included in these graphs are the gas production and solids destruction performance data of the pilot scale reactor.

Pilot Scale Unmixed Horizontal Displacement Reactor

The pilot-scale plug flow reactor at Harford was initially fed 8.0 percent total solids dairy cow manure at a 30-day hydraulic retention time basis without

TABLE 20. PROFILE OF GAS PRODUCTION AND GAS COMPOSITION FROM REACTOR SYSTEMS OF FOUR CSTRs IN SERIES.

Temperature	HRT days	Reactor Stage	Gas (ℓ /day)	(ℓ / ℓ reactor volume-day)	% CH ₄ / % CO ₂
22°C	12	1	0.41	0.137	
		2	0.76	0.253	
		3	0.88	0.293	
		4	1.16	0.387	
	30	1	1.19	0.400	
		2	1.32	0.440	
		3	1.00	0.333	
		4	0.90	0.300	
35°C	12	1	2.36	0.787	50/50
		2	4.98	1.660	49/51
		3	4.64	1.547	51/49
		4	4.42	1.473	57/43
	10	1	3.13	1.043	32/68
		2	4.80	1.600	38/62
		3	4.78	1.593	46/54
		4	2.87	0.957	52/48
	10 40% Recycle	1	4.04	1.347	50/50
		2	4.97	1.657	53/47
		3	3.78	1.260	53/47
		4	3.03	1.010	51/49
	4 20% Recycle	1	3.52	1.173	43/57
		2	5.08	1.693	42/58
		3	5.73	1.910	48/54
		4	4.06	1.353	48/52
60°C	12	1	4.06	1.353	
		2	5.59	1.863	
		3	5.50	1.833	
		4	3.59	1.197	

Figure 34. Total gas production and pH profile for a system of four CSTRs in series: experiments conducted at 35°C.

Figure 35. Profile of gas production by stages in a system of four CSTRs in series simulating the plug flow design at 35°C.

Figure 36. Total gas production and pH profile for a system of four CSTR units in series: experiments conducted at 22°C.

Figure 37. Profile of gas production by stages in a system of four CSTRs in series at 22°C.

Figure 38. Total gas production and pH profile for a system of four CSTRs in series: experiments conducted at 60°C.

TABLE 21. PROFILE OF VOLATILE ACIDS IN REACTOR SYSTEMS OF FOUR CSTRs IN SERIES.

Temperature	HRT, days	Total Volatile Acids, mg/l as Acetic Acid			
		Reactor Stage			
		1	2	3	4
22°C	12	3489	5127	4971	3423
	30	1729	2050	2224	1608
35°C	10	3884	5974	4538	3021
	10 40% Recycle	3340	1896	1878	2135
	12	3075	5246	4338	2939
	4 20% Recycle	4030	3595	4067	3708
60°C	12	6523	7696	5696	5904
	4	2311	4856	5403	6157

Figure 39. TVS destruction efficiency relation to HRT at 22°C for the bench-scale plug flow simulating reactors (20 liters).

Figure 40. Gas production in relation to HRT at 22°C for the bench-scale plug flow simulating reactors (20 liters).

Figure 41. TVS destruction efficiency relation to HRT at 35°C for the bench-scale plug flow simulating reactors.

Figure 42. Gas production relation to HRT at 35°C for the bench-scale plug flow simulating reactors.

recycle. After more than 150 days of operation at this condition, the feed concentration was raised to 11 to 12 percent solids. Operational data for this large pilot-plant are shown in Figure 43. Complete data from these conditions are presented in Appendix Tables B11 and B12. Data from analyses conducted on the manure feedstocks used in this study are also tabulated in Appendix Table B13 and Table 22.

During the 62-day start-up phase of operation, gas production remained below 2500 liters per day (at standard temperature and pressure), gradually increasing with some variation. As indicated in Figure 43, gas production then rapidly increased during the next 12 days to 5000 liters per day, or one volume per reactor volume per day.

The pH of the effluent during start-up varied between 6.8 and 7.4. The pH increased with increased gas production to 7.9, at which point it varied only slightly during steady state operations.

Steady state data from the first phase of operation are tabulated in Table 22. A summary of the range of values observed in all of the influent and effluent parameter measurements is also presented in Table 22. Steady state gas production, from day 73 to day 149, averaged about 5150 liters per day, varying in a range from 3400 to 6200 liters per day. Some of the large variations in the gas production data may be attributed to such things as blocked gas lines, and jammed meters. On a number of occasions, the large 6 inch sampling caps were opened to take floating solids samples, and repair gas effluent lines. At these times the bag would be deflated and the gas lost. At steady state the gas composition consisted of 55 percent methane and 45 percent carbon dioxide.

During the 8.0 percent total solids feed condition, the pH of the feed varied between 7.1 to 8.8, averaging 7.8. Ammonia nitrogen and free ammonia nitrogen varied from 1450 to 2360 mg/l and 34 to 840 mg/l respectively. Total Kjeldahl Nitrogen (TKN) of the feed had a range of 2960 to 4770 mg/l and alkalinity was 9346 to 14,000 mg/l. The total solids averaged 82.1 gm/l,

Figure 43. Chronological development of gas production and effluent pH from the pilot-scale plug flow reactor.

TABLE 22. A SUMMARY OF ANALYSES CONDUCTED ON MANURE FEEDSTOCKS AND EFFLUENTS FROM PLUG FLOW PILOT REACTOR.

Parameter	8% TS Feed Condition				10-12% TS Feed Condition			
	Influent Range	Avg.	Effluent Range	Avg.	Influent Range	Avg.	Effluent Range	Avg.
pH	7.1-8.8	7.8	7.0-8.1	7.7	6.5-7.5	6.7	7.3-7.9	7.7
Alkalinity, g/l CaCO_3	9.3-14.1		11.2-15.6		9.7-14.7		15.0-17.4	
TKN, mg/l	2958-4768		3240-5412		3600-5652		3926-5316	
$\text{NH}_4\text{-N}$, mg/l	1451-2362		1947-2498		1698-2820		2460-2961	
Free Ammonia, mg/l	34-843		38-333		7-86		68-243	
Total Solids, g/l	63-91	82	47-77	59	103-120	112	75-96	88
% Destroyed				28.0				21.4
Total Volatile Solids, g/l	54-79	71	36-65	48	90-104	97	62-80	73
% Destroyed				32.4				24.7

or 8.2 percent, ranging between 62.9 to 90.9 gm/l. Volatile solids of the feed varied 53.8 to 78.9 gm/l, averaging 70.9 gm/l. Average effluent total solids were 59.4 gm/l, or 5.9 percent total solids. Volatile solids averaged 48.4 gm/l. These figures indicated a total solids reduction of 28 percent and volatile solids reduction of 32 percent for the 5 months of operation at a 30 day HRT.

Part of the observed reduction in solids was due to the accumulation of a floating solids layer. Figure 44 is a graphical representation of the increasing thickness of the crust with time, consequently, reducing the effective volume of the digester. This float or crust was composed of fibrous material, with the appearance of undigested silage and hay. The total and volatile solid concentrations of the floating solids are presented in Table 23. Within the float layer, a sample was taken from the top and bottom of the layer. Zone 1 represents the influent portion of the reactor, Zone 2 the middle of the unit and Zone 3 the effluent. Total and volatile solids were highest at the influent end, and within each zone, solids concentration was higher at the top of the layer than the bottom, as would be expected. The float layer was thickest at the input and had a tendency to decrease in thickness as the manure moved through the reactor.

A profile of the distribution of total and volatile solids reduction, and pH change as the manure moved through the reactor, is presented in Table 24. In each set of analysis, both total and volatile solids decrease in each successive zone, indicating the solids destruction with time, going from the influent end (Zone 1) of the reactor to the effluent end (Zone 3). Within each zone, the solids from the center and bottom sampling tubes were of similar concentration. No samples were taken from the top sampling tubes due to the floating solids layer. Figure 45 summarizes the internal solids concentration changes. The decreasing concentration of the parameters shown in Table 24 and Figure 45 indicate an exponential decrease along the length of the reactor that would be expected in a plug flow reactor. More than three quarters of the total organic solids destruction occurred in the first three or four feet of the plug flow reactor, or most of the conversion occurred in the first four or 5 days of retention time.

Figure 44. Chronological development of the float layer thickness by zone in the pilot scale plug flow reactor.

TABLE 23. HARFORD FLUG FLOW PILOT REACTOR FLOAT SOLID CHARACTERISTICS (ALL VALUES IN gm/l).

	Zone 1		Zone 2		Zone 3	
	Top of Float	Float Liquid Interface	Top of Float	Float Liquid Interface	Top of Float	Float Liquid Interface
4/28/77						
Total Solids	142.2	122.9	127.8	119.7	131.8	111.2
Volatile Solids	122.6	106.6	107.5	101.6	107.4	95.1

TABLE 24. DISTRIBUTION OF SOLIDS AND pH IN THE PILOT-SCALE PLUG FLOW REACTOR RECEIVING 8% and 10-12% TS FEED (30-DAY HRT)

Date	Average Influent TS gm/l VS gm/l	Parameter	Zone 1 (Influent end)		Zone 2		Zone 3 (Effluent End)		Effluent
			Center	Bottom	Center	Bottom	Center	Bottom	
1/25/77	78.8	pH	7.2	7.3	7.3	7.3	7.4	7.4	7.6
to		Total Solids	73.3	73.9	71.6	71.2	70.7	72.0	71.0
2/3/77	68.2	Volatile Solids	61.9	62.5	60.3	60.0	58.2	60.3	59.6
2/3/77	87.6	pH	7.3	7.4	7.6	7.7	7.6	7.7	7.8
to		Total Solids	75.5	75.1	70.4	70.8	66.5	66.2	64.0
2/8/77	76.0	Volatile Solids	64.0	63.8	59.3	59.6	55.9	55.3	53.1
2/8/77	87.6	pH	7.5	7.5	7.8	7.6	7.7	7.7	7.7
to		Total Solids	72.8	74.1	69.0	69.2	64.2	65.7	63.1
2/15/77	76.0	Volatile Solids	61.2	62.1	57.7	57.8	53.2	54.9	52.3
2/22/77	84.6	pH	7.6	7.6	7.7	7.7	7.7	7.8	7.8
to		Total Solids	75.8	78.5	65.0	61.8	59.6	59.8	58.6
3/1/77	73.4	Volatile Solids	63.8	66.5	53.8	50.9	48.8	42.9	47.6
3/1/77	86.1	pH	7.8	7.9	7.9	7.9	7.8	7.9	7.9
to		Total Solids	69.5	67.4	61.4	60.2	58.4	58.3	57.0
3/8/77	74.2	Volatile Solids	58.1	56.1	50.3	49.0	47.4	47.3	46.0
3/8/77	81.3	pH	7.9	7.9	7.9	7.9	7.8	7.9	7.9
to		Total Solids	69.3	66.8	58.8	59.2	56.0	55.3	54.7
3/15/77	69.8	Volatile Solids	57.5	55.2	47.7	48.1	44.8	44.2	43.6
3/15/77	78.4	pH	7.8	7.8	7.8	7.8	7.8	7.9	7.9
to		Total Solids	64.4	68.0	60.0	60.0	53.6	54.3	50.6
3/22/77	66.4								

TABLE 24 CONT'D.

Date	Average Influent TS gm/l VS gm/l	Parameter	Zone 1 (Influent End)		Zone 2		Zone 3 (Effluent End)		Effluent
			Center	Bottom	Center	Bottom	Center	Bottom	
		Volatile Solids	53.6	56.6	49.2	48.8	42.8	43.7	39.4
3/31/77	75.9	pH	7.9	7.9	7.9	7.9	7.9	7.9	7.9
to		Total Solids	58.3	59.7	52.1	53.4	50.1	51.5	46.9
4/7/77	64.6	Volatile Solids	46.9	43.3	41.3	42.3	39.3	40.5	36.4
4/22/77	86.4	pH	7.9	7.8	7.8	7.9	7.8	7.9	7.9
to		Total Solids	50.8	51.4	56.8	57.3	53.8	55.0	50.8
4/29/77	74.8	Volatile Solids	50.1	51.0	46.4	46.7	43.6	44.2	40.3
6/30/77	110.8	pH	7.7	7.7	7.7	7.7	7.6	7.7	7.6
to		Total Solids	92.6	91.6	91.5	91.6	90.1	92.6	90.0
7/7/77	95.3	Volatile Solids	76.7	76.6	75.2	75.5	73.9	74.9	73.2
7/7/77	104.9	pH	7.8	7.8	7.8	7.8	7.7	7.7	7.7
to		Total Solids	92.6	90.9	86.6	90.4	89.6	91.1	88.3
7/12/77	89.9	Volatile Solids	76.0	74.8	70.9	74.3	73.6	74.7	72.2
7.12/66	108.3	pH	7.7	7.7	7.8	7.8	7.7	7.7	7.7
to		Total Solids	88.6	89.2	89.8	88.9	89.2	89.6	88.4
7/19/77	94.0	Volatile Solids	72.8	73.2	73.7	73.0	72.9	73.4	70.7

Figure 45. Internal total solids concentration in the plug flow pilot reactor at feed solids of 8 and 11 percent, 30 day HRT and 35°C. Samples were taken from the center of the reactor and each point represents an average of 5 steady state samples.

During the latter stages of operation, the gradual increase in pH reflected a decreased volatile acid concentration. During a steady-state, the pH was relatively the same throughout the length of the fermentor.

The feed for the second solids loading condition was 10 to 12 percent total solids, as it was removed from the barn floor without dilution. The heavier solids loading was started after 156 days of operation at 8 percent total solids. Average gas production at the higher loading conditions was 5700 liters per day (STP) with a high of 8300 liters per day and a low of 3870 liters per day (Figure 43). Gas composition averaged 61 percent methane, and 39 percent carbon dioxide, which was of higher quality than the composition observed in the first operating condition.

In comparing ammonia, total kjedahl nitrogen and alkalinity of the feed, both feed conditions were similar in composition, as indicated in Table 22. The ammonia and alkalinity concentration were higher in the effluent at 10-12 percent solids loading than at the 8 percent solids loading. Total kjedahl nitrogen was about the same.

Total and volatile solids of the feed ranged between 102.9 to 119.2 gm/l and 89.5 to 103.7 gm/l, averaging 111.6 and 96.7 gm/l, respectively for the two conditions. The effluent solids averaged 88.5 gm/l total solids and 72.0 gm/l volatile solids at steady state, indicating 22 percent total solids destruction and 25 percent volatile solids destruction, respectively.

The floating solids in the pilot scale unit were forced out by higher influent total solids concentration during the initial operating phases with higher feed solid concentrations. No significant new layer formed since this change thereafter.

DISCUSSION

Bench Scale Plug Flow Simulation Studies

The following topics will be used to discuss the degree of success achieved with the small scale units: methane production to solids

destruction balance; influence of detention time on system performance; volatile acids data interpretation; the effect of recycling; and a comparison with complete-mixed control reactors. Also, included is a kinetic analysis of the results of the bench scale study.

Methane Production and VS Destruction. Theoretically, one gram of complex organic TVS utilized in anaerobic fermentation should result on 0.5 liters of methane synthesized. The equation relating solids destruction with the theoretical methane gas production at STP was expressed in Equation 17:

$$G_t = 0.5(S_0 - S_1)/HRT \quad \text{Eq. 17}$$

Where: G_t = methane production at STP (volume CH_4 /volume reactor/day),
 S_0 = Influent Volatile Solids Concentration
 S_1 = Effluent Volatile Solids Concentration

The predicted and observed values of methane production for all plug flow units are given in Table 25. The ratio of measured to predicted methane products is lower than expected. One possible explanation is that at low detention times, i.e. higher stress conditions, the volatile solids are readily stabilized to volatile acids which are not completely fermented by the methane formers. This incomplete fermentation would be directly related to the physical and environmental factors which promote the stress situation. As shown in the results, the volatile acids concentration in the reactors generally did increase with successively shorter detention times.

High concentrations of volatile acids may introduce an error into the measurement of volatile solids since the acids will volatilize during drying at 103°C (Lawrence and McCarty, 1967). This introduces an error into the volatile solids destruction data and into the methane yield predictions based on VS reduction. Using the volatile acid data for each condition the volatile solids data was corrected to account for this error into the volatile acid concentrations to the solids measurements. Table 26 shows the corrected methane yield predictions. A significant increase in agreement between predicted and measured values of methane production was observed. The level of agreement decreased as the system approached higher stress conditions. A ratio of observed to theoretical methane of 0.98

TABLE 25. OBSERVED METHANE AND THEORETICAL METHANE PRODUCTION CALCULATED FROM TVS DESTRUCTION DATA: BENCH-SCALE AND PILOT-SCALE PLUG FLOW SIMULATION FERMENTORS.

Reactor 35°C	HRT days	TVS g/l day	Destruction g/day	Theoretical CH ₄ l/day STP	Observed CH ₄ l/day STP	Observed CH ₄ Predicted CH ₄
Pilot Scale Plug Flow Reactor	30	36.6	6954	3407	2828	0.83
"	12	20.5	20.5	10.0	8.5	0.85
Systems of Four CSTRs in Series	10 40% Recycle	20.9	25.1	12.3	8.1	0.66
"	10	19.6	23.6	11.5	6.4	0.56
CSTRs Control Reactors	10	19.6	5.9	2.8	1.6	0.57
	12	25.6	6.4	3.0	2.6	0.87

TABLE 26. OBSERVED METHANE AND THEORETICAL METHANE PRODUCTION CALCULATED FROM TVS DESTRUCTION DATA CORRECTED FOR VOLATILE ACIDS LOSS DURING DRYING BENCH-SCALE AND PILOT-SCALE PLUG FLOW SIMULATION FERMENTORS

Reactor	HRT days	TVS Destroyed gm/day	Volatile Acids gm/day	Corrected TVS destroyed gm/day	Predicted CH ₄ ℓ/day	Observed CH ₄ Predicted CH ₄
Pilot-Scale UHD Reactor	30	6954	89.3	6865	3364	0.84
Reactor Systems of Four CSTRs	12	20.5	2.94	17.8	8.70	0.98
In Series	10R	25.1	2.56	22.5	11.0	0.75
	10	23.6	3.62	19.9	9.77	0.67
Completely Mixed Control Reactors	10	5.9	.22	5.7	2.7	0.59
	12	6.4	.11	6.3	3.0	0.87

corresponded to the 12-day HRT, and this decreased to 0.74 at the 10-day HRT with recycle and 0.67 at the 10-day HRT without recycle. The Pilot plug flow reactor showed a 0.84 observed to theoretical methane production ratio at a 30-day HRT. However, this is most likely explained by solids accumulation in the float within the reactor.

Effect of HRT on System Performance. Figures 40 and 42 show the relationship between TVS destruction and HRT at 22° and 35°C, respectively. The destruction efficiencies of TVS increased with increasing HRT, but leveled off at the higher HRTs. This is related to the extent of biodegradability of dairy manure. At the longer HRTs, the majority of the biodegradable fraction of the manure had been assimilated.

The unit gas production per gram of VS added is depicted in Figures 39 and 41 for the 22° and 35°C experiments, respectively. Unit gas production (l/g VS added) at 22°C increased throughout the study, but since the loading (gVS/l reactor/day) was reduced at high HRTs, the total gas production (unit gas production times loading) declined. Therefore, an optimized design at 22°C, in terms of total gas production and reduction of BVS should specify and HRT equal to or greater than 30 days.

It is significant to note the relative success of operating an anaerobic fermentor at 22°C versus 35°C. Low temperature operation in terms of reactor design and construction costs may increase, but this may be offset by increased net energy production. However, the added incentive of diminished energy input to maintain the system makes this an alternative which will be more closely evaluated in future experimental programs.

Unit gas production at 35°C increases through 20 days and drops off slightly at 30 days. Consideration of designing the reactor to achieve maximum gas production and BVS destruction would result in an HRT of 15 to 20 days.

Interpretation of Volatile Acids Data. The breakdown of volatile acid data into specific components of acetic, propionic, butyric, and valeric acids is presented in Appendix Table B14. It is interesting to note the concentration trends of each organic acid through each stage of the series experiments, and

how these trends compare with theoretical predictions. Next to formic acid, acetic acid is the most rapidly converted acid in methane fermentation, and is responsible for 70% of the total methane production (McCarty, 1964; Kugelman and Chin, 1971). Only 20% of the acetic acid is directly converted from organics, however. The other 50% is derived from propionic and other intermediate acids (McCarty, 1964). Few species of methane bacteria utilize propionic, valeric or butyric acids (Toerien and Hattingh, 1969). These acids may be converted to acetic acid or utilized by a hydrogen-forming bacteria in the reactor. Therefore, it would be reasonable to expect an increase in the acetic and propionic acid concentrations in the acid-forming stages of the four-stage series, and a decrease in propionic acid in the latter stages as it is converted to acetic acid. The acetic acid should also be converted to methane, but this utilization should be moderated by the addition of acetic acid as a second-step end product. Intermediates such as butyric and valeric acids will also contribute to the concentration of propionic acid as well as to acetic acid production.

In nearly all of the series experiments, the total volatile acid concentration of stages one and four were nearly equal. There was generally a large increase from stage one to stage two, except in the recycle studies. The total acid concentration then began a slight decline in stage three, continuing this trend into stage four. This removal of volatile acids was accompanied by an increase in methane percentage of the gas in each successive reactor.

A decline in acetic acid concentration was observed throughout each reactor system, as would be expected. The minor intermediates, butyric and valeric acids, also showed a reduction as they progressed through the stages. Since this decrease was not accompanied by an increase in total acetic acid, it may be interpreted that this additional source of acetic acid was readily converted to methane in the latter stages. One interesting observation was that the propionic acid concentration increased or stayed the same, in all but two of the test conditions, throughout the entire system. The two exceptions, 30 days at 22°C and 4 days with recycle at 35°C, did show an increase through three stages with only a slight decline in the fourth stage.

This phenomena might be explained when one considers the supposition that methane-forming bacteria do not directly convert propionic acid to methane (Toerien and Hattingh, 1969); the propionic acid would essentially be an intermediate step in the formation of acetic acid which is converted to methane. The concentrations observed in the systems, generally ranging from 1.0 to 2.5 gm/l (as acetic acid), indicate a significant amount of propionate in the reactor. The inhibitory and toxic effects of propionate on bacteria have been previously observed (Stadtman and Barker, 1951; Kugelman and Chin, 1971). Kugelman and Chin (1971) reported on a study which indicated that propionate inhibited the conversion of propionic acid to acetic acid. This study concluded that propionate had little effect on methane production but could severely affect acid production. Thus the absence of a decline in the propionic acid concentration across the stages in the series reactors may be related to the inhibitory effect of propionate on conversion of propionic to acetic acid. This incomplete fermentation would also be reflected in the observed methane production, another possible explanation for the low ratios of observed/predicted methane production.

A similar observation of propionic and butyric acids not being assimilated at low detention times was reported by Pohland and Ghosh (1971). This was attributed to slower conversion rates by fewer species as well as the preference for acetic acid by methane formers. However, at those conditions of low detention times (higher acid production, thus higher concentrations of propionate) it is possible that some inhibition of acid formers was also taking place to restrict conversion of the intermediate acids to acetic acid.

Effects of Recycling. The effect of recycling was investigated in the bench-scale series experiments at 35°C. The main objectives in the recycling study were to attempt to: 1) increase overall methane production; and 2) maintain stable conditions at very low detention times (i.e., one day HRT or less).

The first condition included operation of the four stage system at an overall volume equivalent to 40 percent of the raw feed. Effluent from the fourth reactor in series was returned to reactor one with the raw feed. This was run concurrently with a system at a 10-day HRT without recycle. A summary

of the results is included in Table 19. Solids destruction and total gas production increased slightly. A significant increase in methane production was observed, from 0.276 to 0.325 liters $\text{CH}_4/\text{gm VS}_D\text{-day}$. The increase in methane percentage by stage is indicated in Table 20. Whereas the non-recycle experiment only achieved 50% methane in the final stage of the series, the recycle study yielded near or above 50% methane in all four stages. The added stimulus of seeding a healthy methane population into the first stage appeared to have a significant effect on establishing an efficient methane fermentation process early in the system. The effect of seeding methane formers into the first stage is also evident in the concentration of volatile acids observed in each stage as shown in Table B14. The conversion of organic acids to methane is much more efficient in the second stage of the recycle system.

The second recycle condition examined a very short detention time of one day per stage, or a four day system detention time. The recycled effluent was maintained at the same volume as in the first condition, with a relative decrease to 20% of the raw feed. Stable conditions were achieved but total system efficiency was reduced. Total gas production was maintained at a higher level than the 10-day recycle study, but TVS reduction dropped from 26 to 22 percent. Table 20 indicates a decline in gas quality to less than 50% methane in all four stages. Unit methane production dropped to 0.212 liters $\text{CH}_4/\text{gm VS}_D\text{-day}$. Therefore, while it was possible to establish a stable system at a detention time of one day per stage, the decline in efficiency is a disadvantage of this high rate design. These data illustrate the variables which could be considered to increase the rate of methane production by taking advantage of mixing theory and the fundamentals of microbial reproduction. Application of these principles to a more biodegradable substrate would result in more dramatic improvement of the process. Careful process control and buffering would be required in high rate series reactors with or without recycle.

Comparison of the CSTR Control Units With Plug Flow Reactors. The plug flow model data and the conventional control reactors are compared in terms of several parameters in Table 27. The results of the low temperature (22°C) or short detention time experiments indicated that the plug flow models were operating under greater stress than the conventional reactors. At the higher

TABLE 27. COMPARISON OF CONTROL UNITS TO PLUG FLOW MODEL, (expressed as plug flow/control ratio).

Reactor	Temp.	HRT	Performance Parameters					
			TVS red.	GAS		METHANE		
				$\ell/\ell-d$	ℓ/VS_{A-d}	ℓ/VS_{D-d}	$\ell/\ell-d$	ℓ/VS_{D-d}
Four CSTRs*	22	12	0.66	0.41	0.40	0.65		
"	"	22	30	1.07	0.83	0.83	0.77	
"	"	35	10	1.00	0.85	0.86	0.85	
"	"	35	12	0.81	0.89	0.89	1.11	0.83 1.03
Plug flow- Bench scale	35	20	1.08	1.30	1.28	1.19	1.18	1.08
Plug flow- Pilot unit	35	30	1.10	1.09	1.06	0.96	1.12	0.99
Plug flow- Pilot unit	35	30	0.80	1.19	1.35	1.44	1.14	1.38

*Four completely mixed reactors in series

temperature, longer detention time studies, the plug flow models exhibited a more stable, efficient operation. The actual differences in performance characteristics, however, were not very significant as reflected by the comparative parameter ratios in Table 27.

Thus, as would be expected it was observed that mixing might offer an advantage towards efficient system operation at high loading rates. It may, nevertheless, be concluded that plug flow models were nearly as efficient or more efficient in most cases than the conventional completely-mixed control reactors.

Pilot Scale Plug Flow Reactor

Feasibility of the Plug Flow Design. The large scale pilot was successfully operated without mechanical mixing and with solids feed up to 12 percent. These are the first data to assist in explaining the lower limits of this unique design. It is valuable because it provides evidence that indicates that the strict requirements for efficient operation of an anaerobic fermentor with sewage sludge does not apply to substrates such as dairy cow manure. Additional evidence also indicated that temperatures lower than 35°C would have little effect on the total potential gas production. Additional studies need to identify the optimum temperature based on maximum gas production at minimum reactor cost.

Although there is insufficient data to support a complete kinetic comparison between the conventional completely mixed design and the plug flow design, some qualitative observations can be made. The data for the decrease of solids concentration in the plug flow reactor at both the 8 percent and 12 percent solids feed concentration shown in Figures 45a and 45b illustrate that most of the solids are converted in the first 10 percent of the volume of the reactor, or in about 3 to 5 days retention time. Data from the mesophilic completely mixed reactors show that this retention time results in an unstable reactor with little solids reduction. Even increasing the temperature to 60°C to obtain higher microbial reproduction rates resulted in failure at solids concentration below 6 percent and less than 3 days HRT (see Chapter 2).

However, multiple completely mixed reactors connected in series can be successfully operated at retention times down to 1 day HRT in each unit at 35°C, provided that about 20 percent of the effluent is recycled.

Although the information obtained in this study regarding reasons for the success of the simplified design is not sufficient to completely clarify this concept, there are several areas which can assist in this explanation. First, the homogeneous substrate and the presence of a significant number of methanogens in the feed simplify the reactor requirements. Secondly, the high buffering capacity combined with the high viscosity results in lower concern for mixing and minimum adverse impact due to such variables as volatile acid accumulation.

Finally, this design information opens the possibility of implementing low cost reactors in many different installations. Simplistically, this can be illustrated by comparing reactor costs to the potential energy value. The design gas production rates will be about one volume per reactor per day, and this will result in an annual gross energy value of about \$0.50 per cubic foot of reactor. Rigid tanks with floating gas covers costing \$3 to \$10 per cubic foot result in severe economic limitations. However, it appears that the use of soil supported structures and tanks and covers made of flexible materials, such as Hypalon, will result in total reactor costs of less than \$0.50 per cubic foot. This enables the design to be much more competitive. Furthermore, since the liner material itself costs less than \$0.10 per cubic foot of reactor it can be replaced more often than can rigid, more costly materials.

Two conditions were investigated with the large pilot-scale reactor. Operation of the unit with "as produced" or undiluted animal residues produced several important observations regarding the operation of a plug flow fermentor. The influence of each condition on various parameters of operation will be discussed in the following subsections: float solids management; solids destruction and gas production; tracer study results; and volatile acids and pH relationships.

Float Solids Management. One of the primary reasons for changing the influent total solids from 8 percent to an "as received" 10-12 percent concentration was to test the hypothesis that high solids concentration would force the float solids which had been accumulating, out of the reactor. During the diluted feed condition, dry fibrous material had accumulated and occupied about 40% of the reactor volume. It was hypothesized that the main mechanism of float formation was attachment of gas bubbles to the fiber particles, thus reducing the apparent density of the float solids which would subsequently cause them to rise and concentrate at the surface.

Additional advantages to increasing the feed concentration would include: 1) fewer manure handling problems as would be incurred if dilution were necessary; and 2) decreased reactor volume required to handle the same number of animals. One way to inhibit this flotation process would be to increase the viscosity of the reactor contents, thus making gas bubble attachment and bubble-particle movement through the liquid regime more difficult. The viscosity could have been increased by: 1) decreasing the digester temperature; or 2) increasing the total solids concentration in the feed. Previous small-scale studies had indicated that a decrease in temperature brought about a significant decline in gas production. The increase in solids concentration from 8 percent to 10 percent completely eliminated the float problem.

The effect of increased solids concentration on viscosity may be observed in Table 28. It is apparent that a significant increase in viscosity occurs at an increase from 8 to 11 percent total solids in dairy manure at elevated temperatures. This increased viscosity also enhances another aspect of the unmixed plug flow design in that it reduces the necessity for mixing to keep the solids evenly suspended throughout the digester. An interesting comparison may be made with the viscosity values for digested sewage sludge, which appear to be an order of magnitude lower than dairy manure. An unmixed reactor design will be less successful in operation with municipal sewage sludge feeds due to the possible partitioning or settling of different density material within the digester. The stratification of layers in unmixed municipal digesters has been typically observed and well documented in the literature (Pohland and Ghosh, 1971).

TABLE 28. A COMPARISON OF VISCOSITIES OF VARIOUS RESIDUES.

Type of Waste	T.S. (%)	Temp. (°C)	Shear Rate (1/sec)	Viscosity ₂ dync-sec/cm ²	References
Fresh Dairy Manure	11.02	24	30	19.97	Kumar, <u>et al.</u> , (1972)
	8.6	7	30	16.80	
	8.6	42	30	8.71	
Beef Manure					
	8.0	25	4.41	125	Bashford, <u>et al</u>
	10.0	25	4.41	345	
	12.0	25	4.41	815	
	8.0	25	11.02	85	
	10.0	25	11.02	200	
	12.0	25	11.02	520	
	8.0	25	22.04	65	
	10.0	25	22.04	135	
	12.0	25	22.04	355	
Digested Sludge					
	10.0	17	--	0.92	Weber (1972)
	12.0	17	--	0.98	
	14.0	17	--	1.01	
	16.0	17	--	1.13	
	18.0	17	--	1.16	

Solids Destruction and Gas Production. The two loading rates investigated at a 30-day detention time were 0.17 and 0.23 lb/ft³/day, as reported earlier for the 8% TS and 10-12% TS manure feed conditions, respectively. Table 29 compares some of the performance characteristics of each condition. The reduction of volatile solids at the second loading rate (10-12% TS feed) was measured as 25%. This represented a decrease from the extremely high reduction value of 49% observed at the first condition. However, in light of the accumulation of float material during the 8% TS feed condition, it was concluded that the high solids destruction efficiency observed was undoubtedly inflated by the separation of non-degraded particles from the active digester zone of the reactor. Conversely, the TVS reduction reported for the 10-12% TS feed condition, under which no solids accumulation was observed, should be considered to be the more reliable estimate of the solids-destruction capability of the plug flow reactor.

The unit gas production per gram of VS added to the system was observed to be 0.357 and 0.333 in the conditions of 8% TS and 10-12% TS feed, respectively.

The rates of gas production throughout the reactor were examined in grab samples taken from the center of each of the three reactor zones. The serum bottle technique fitted with gas collection syringes for measurements of initial gas production rates and gas composition were utilized to obtain a measure of the activity. Gas production rates and composition analyses which were performed on these internal samples produced results shown in Table 30. The influent third of the reactor was seen to produce the greatest quantity of high quality digester gas in the entire reactor. The methane percentage declines through the reactor, offering the lowest yield in the effluent zone of the fermentor. These data support the fact that the unit was approximating a plug flow reactor.

The decreasing percentage methane in the gas with increasing retention time was observed in many instances. The constituents of dairy manure and the relative amount of time required for complete waste stabilization assist in explaining the varying gas composition in stable reactors. Dairy animals fed a higher roughage diet will excrete a high percentage of slowly degraded material.

TABLE 29. COMPARISON OF PERFORMANCE CHARACTERISTICS IN THE PILOT-SCALE PLUG FLOW REACTOR RECEIVING 8% AND 10-12% TS MANURE FEED. THE HRT IN BOTH CASES WAS 30 DAY HRT.

Parameter	8% Feed	10-12% Feed
Influent Total Solids gm/l	86.1	112.4
Influent Total Volatile Solids gm/l	74.6	97.0
Effluent Total Solids gm/l	48.7	88.5
gm/l reduced	37.4	23.9
% reduced	43.4	21.3
Effluent Total Volatile Solids gm/l	38.0	72.0
gm/l reduced	36.6	25.0
% reduced	49.1	25.7
Gas Production l/day	5065	6145
Gas Composition %CH ₄ /%CO ₂	56/44	61/39
Methane Production l/day	2821	3750
l gas/gm VS _A	0.357	0.333
l gas/gm VS _D	0.728	1.290
l CH ₄ /gm VS _D	0.406	0.789

TABLE 30. PROFILES OF GAS PRODUCTION AND GAS COMPOSITION EFFECTED BY THE CONTINUED FERMENTATION IN SERUM BOTTLES (35°C) OF ALIQUOT SAMPLES TAKEN FROM THE PILOT-SCALE PLUG FLOW FERMENTOR OPERATED AT 10-12% TS FEED, 30 DAY HRT, 35° C.

Zone	Total Gas Production ℓ/ℓ	%CH ₄	%CO ₂
1	1.38	65	35
2	0.78	56	44
3	0.69	47	53

TS = 12%

HRT = 30 days

Smith, et al., (1969) reported, for silage-fed heifers, that nearly 66 percent of the fecal dry matter was composed of plant cell wall content. The individual constituents were expressed as percent dry weight of the cell wall: hemicellulose, 30%; cellulose, 40%; and lignin, 30%. These biologically resistant particles account for the low measure of biodegradability of dairy manure. The hemicellulose and the cellulose comprise a large fraction of the total carbohydrates in dairy manure. The remaining percentage of the manure is made up of fats, proteins, urea, and other soluble matter. The proteins will exist mainly in the form of dead microorganisms.

The digestability of each of these fractions was summarized for the fermentation of domestic sludges by Chynoweth and Mah (1971). They reported that lipids were 76 percent digestible, followed by proteins (36%) and carbohydrates (13%). While the nature of these elements will be different in animal wastes, the fact that carbohydrates are slowly degraded remains valid.

Therefore, as the manure moves through the reactor, the easily degradable components such as urea, fats, and proteins may be assimilated more rapidly. This would leave the more resistant materials, cellulose, hemicellulose, and lignin, to be degraded at later stages in the fermentation process.

It is possible to determine the percent of methane which should be produced in the gas resulting from the specific degradation of urea, fats, proteins, and carbohydrates using the equation developed by McCarty (1964) for predicting gas composition based on the chemical structure of the waste substrate. The generalized equation may be expressed as:

The percentage of methane yield per specific organic substrate is summarized in Table 31. Typical chemical compositions which were used in the calculations were: 1) $CO(NH_2)_2$ for urea; 2) $C_{15} H_{31} COOH$ for fats; 3) $C_4 H_6 ON$ for protein; and 4) $C_6 H_{12} O_6$ for carbohydrates (McCarty, 1971). Readily degradable substrates (urea, fats, and proteins) yield the highest percentages of methane. These theoretical considerations support the observation that

the influent third of the plug flow fermentor could be expected to produce the highest quality gas if the proteinaceous material, and lipids were decayed first. As the manure reaches the effluent end, microbial degradation occurs more slowly, because the majority of the organics being assimilated are the biologically resistant celluloses and lignin complexes. As noted in Table 31 these produce a lower quality gas, which corresponds to the gas composition reported in the latter stages of the pilot scale digester.

TABLE 31. THEORETICAL METHANE YIELD OF VARIOUS SUBSTRATES.

Substrate	Chemical Composition	Methane Yield (% Total Gas)
Urea	$\text{CO}(\text{NH}_2)_2$	75
Fat	$\text{C}_{15}\text{H}_{31}\text{COOH}$	72
Protein	$\text{C}_4\text{H}_6\text{ON}$	63
Carbohydrate	$\text{C}_6\text{H}_{12}\text{O}_6$	50

Therefore, the internal gas production and composition values reported indicated that each zone is responsible for a specific type of microbial assimilation in the overall fermentation of complex organic wastes.

Tracer Study. The experiment to measure the degree of mixing using a Cr tagged fiber and a CoLi-EDTA liquid tracer, provided an independent measure of solid and liquid detention times. A sample chloride tracer was also used to serve as a verification for a cobalt tracer.

The observed results of the tracer study during the 8% TS feed condition are illustrated in Figure 46. Profiles of effluent chromium and cobalt concentrations show that the 30-day design detention time was not fully attained. Effluent concentrations of both chromium and cobalt reached maximum values around 17 days after tracer introduction. The F-curves for the chromium and cobalt constructed from the graphical integration of the concentration versus

Figure 46. Effluent chromium and cobalt tracer concentrations from a pilot scale plug flow fermentor operating at 8% TS feed, 30 day HRT, with float material present.

Figure 47. F-curves for cobalt and chromium tracers passing through a pilot scale plug flow reactor operated at 8% TS feed, 30 day HRT, with float material present.

time curves are shown in Figure 47. The maximum rates of chromium and cobalt being expelled from the UHD reactor, the inflection points of the tracer F curves, occurred around 20 days. The mean effective hydraulic and solids retention times of the plug flow reactor during the first operating condition could both, therefore, be considered to fall between 17 and 20 days. The difference between observed and theoretical values of the solids and hydraulic retention times may be explained, in part, by the significant amount of float solids accumulation which filled nearly forty percent of the reactor volume in the testing with 8 percent solids feed. This would have reduced the effective detention time of the active digester zone to 17 days (considering a 5585 liter digester with 190 liters fed daily).

The three curves describing the tracer study for the 10-12% TS feed condition are presented in Figure 48. Once again the observed peak values for effluent chromium and cobalt concentrations were observed to occur around 17 days. During this operating condition, the highest rate of soluble and insoluble tracer expulsion from the UHD reactor occurred at 17 to 18 days after the tracer introduction as shown in the F-curves for cobalt and chromium in Figure 49.

The effluent concentration profile curves representing the results of each tracer study are similar in general shape to the standard curves used to describe a dispersed plug flow system with a moderate to large amount of dispersion (Weber, 1972; Levenspiel, 1972). This should correspond to a dimensionless dispersion coefficient between 0.025 and 0.2; the dispersion coefficient assigned to complete-mixed flow is infinity. Furthermore, the shapes of the F-curves for the soluble and insoluble tracers resemble the general F-curve shape characteristics typically obtained with tracers passed through dispersed plug flow systems (Gloyne and Eckenfelder, 1968).

Volatile Acids. The volatile acid concentrations measured in the pilot-scale digester are summarized in Table 32. The total volatile acid concentrations also decrease in a manner representing of a plug flow design. One observation which was contrary to the results noted in the bench-scale CSTRs-in-series

Figure 48. Effluent chromium, cobalt and chloride tracer concentrations from a pilot scale plug flow fermentor operated at 10 to 12% TS feed, 30 day HRT, with no float material present.

Figure 49. F-curves for cobalt and chromium tracers passing through a pilot scale plug flow reactor operated at 10 to 12% TS feed, 30 day HRT, with no float material present.

experiments was the nonaccumulation of propionic acid along the length of the reactor. As previously reported, the high concentrations of propionate were thought to act in an inhibitory manner towards the acid-forming bacteria responsible for conversion of propionic acid to acetic acid. However, the nearly ten-fold decrease in the propionic acid concentration between the bench and pilot-scale units may have placed the system below the inhibitory concentration. In the study reported by Kugleman and Chin (1971), propionic acid assimilation was inhibited only with additions of high concentrations (>6 gm/l) of propionic acid which would then be oxidized to the propionate form. This value of 6 gm/l propionic acid (expressed as acetic acid) had been nearly attained in several of the CSTRs-in-series experiments.

TABLE 32. PROFILE OF VOLATILE ACID CONCENTRATIONS IN PILOT SCALE REACTOR.

Zone	Sample Location	Acetic	Propionic	Butyric	Valeric	Total as (mg/l acetic)
1	C	570	76	68	-	714
	B	580	76	68	-	724
2	C	340	80	--	-	420
	B	340	80	--	-	420
3	C	245	40	--	-	285
	B	195	40	--	-	235

Notes: C = 10 inches from the bottom

B = 23 inches from the bottom

Ammonia and pH. The pH of the digester was consistently in the range of 7.0 to 8.0 throughout the entire period of operation averaging 7.7. The internal pH profile showed little variation. This is indicative of three important features of the system. First, the natural buffering capacity of dairy manure

is apparently sufficient to maintain a high pH during the high rate of volatile acid formation in the first zone. Second, the populations of acid formers and methane formers were in proper balance to prevent the excessive accumulation of organic acids in any zone of the digester. Third, the mixing provided by the thermal gradients, feeding technique, and gas formation cause a significant departure from ideal plug flow and this resulted in minimum pH variation.

The greatest variation in pH occurred between the raw feed and the influent third of the digester. This may be attributed to the rapid conversion of organic nitrogen to ammonia nitrogen; this reaction removes H^+ ions from solution, thereby increasing the pH. At no point in the study was ammonia toxicity a problem, except possibly for one short period (around day 50) where free ammonia of the feed was greater than 800 mg/l (pH of feed was 8.8). The change in feed characteristics was attributed to the fact that the manure had frozen in the barn and could not be collected for several days. This resulted in an increase of effluent free ammonia to 200-300 mg/l, accompanied by a temporary decline in gas production, in the period of days 90-100. Figures 50a and 50b illustrate the changes observed in influent and effluent ammonia concentrations during this entire period. Although 800 mg/l free ammonia is generally considered toxic for digestion of municipal sewage sludge, anaerobic systems using swine wastes have been successfully operated under much higher influent ammonia concentration (Kroeker, et al., 1976).

The pilot-scale reactor had no difficulty in regaining its previous stability after being briefly pulsed with 800-900 mg/l influent free ammonia. Thus it was observed that this plug flow design could maintain steady operation under a "shock load" of influent free ammonia levels exceeding 800 mg/l and could recover from a temporary period of limited inhibition. This is an important conclusion, since one proposed disadvantage of plug flow reactors is that they are unable to efficiently handle shock loads of toxic materials. Though limited, the dispersion characteristics of the pilot-scale UHD reactor would tend to modify the effects of such pulses of toxic substances. It appears reasonable that a proper period of acclimation could enable such a system to operate efficiently at high ammonia concentrations, as suggested by other studies (Kroeker, et al., 1976; Melbiner and Donnellon, 1971).

Figure 50a. Ammonia and pH characteristics of the dairy cow manure feedstocks diluted to 8% TS used for the operation of the pilot scale plug flow fermentor at 30 days HRT.

Figure 53b. Ammonia and pH characteristics of effluent from the pilot-scale plug flow fermentor operated at 30 days HRT, 8% TS manure feed.

Figure 50c. Influent and effluent free ammonia concentrations in the pilot scale plug flow fermentor operated at 30 days HRT, 8% TS manure.

SUMMARY AND CONCLUSIONS

All of the reactor systems in the plug flow investigation were successfully operated at each condition with little evidence of process distress or failure. The bench scale study provided valuable information on the performance capabilities of plug flow simulating reactors operated at high solids and short hydraulic loadings. The pilot-scale plug flow fermentation experiments demonstrated that efficient solids destruction and gas production performance may be attained in the anaerobic fermentation of high concentration dairy manure residues (up to 10-12 percent TS) with simplified reactors. This phase of the program indicated that several engineering considerations, including float solids management techniques should be taken into account in a full scale design.

Specifically, the following conclusions seemed appropriate:

1. The refractory fraction of dairy cow manure at mesophilic temperatures was observed to be 56 percent, or 44 percent was biodegradable and available for conversion to methane.
2. The efficiency of dairy cow manure solids destruction which can be achieved in a plug flow simulating reactor is at least 25 percent in a stable operating unit at a 30 day HRT, and possibly much shorter.
3. Plug flow simulating reactors were found to be capable of solids destruction and gas production rates comparable to conventional completely mixed reactors at all conditions investigated, showing greater relative solids destruction efficiency at equal detention times;
4. The percent of methane measured in the gas from all reactors varied between 55 and 61 percent, with the highest percentage observed in the pilot scale reactor;
5. The bench scale experiments, operating at low detention times, appeared to accumulate propionic acid.
6. Recycling in the bench scale series reactors used to simulate plug flow conditions produced a significant increase in methane production. At a 10 day detention time (35°C - 2.5 days in each reactor) the

methane yield increased from 0.543 to 0.681 liters per liter of reactor by recycling a volume of effluent equal to 40 percent of the total feed volume.

7. Four series reactors with recycle can be operated at retention times as low as 1 day in each unit.
8. Float solids accumulated in the pilot scale reactor at the first loading condition and occupied 40 percent of the volume, when the feed total solids was 8 percent. When the feed concentration was increased to an "as received" 10 to 12 percent concentration, the float solids were sheared off and displaced out of the digester.
9. The solid and liquid detention times within the pilot scale reactor were equivalent as measured by tracer studies. The average retention time was about two thirds the calculated value thus indicating that a significant amount of mixing and dispersion occurred in this reactor.
10. Total gas production in the pilot scale reactor was 0.357 (0.91) and 0.333 (1.10) $\ell/\text{gm VS}_A$ (ℓ/ℓ reactor day) during the loading rates of 0.17 and 0.23 lb/ft^3 day, respectively.
11. The influent third of the pilot scale unit appears to be responsible for most of the methane production through the rapid assimilation of fats, proteins, and urea; slowly degraded materials, such as cellulose and lignin are degraded in the latter stages of digestion.
12. Because of the successful operation of the large pilot plug flow unit, its potential impact on the economics of methane production from agricultural residues and its ease of operation, this design concept is presently being tested for a 65 cow unit at Cornell University.

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CHAPTER IV

ANAEROBIC FERMENTATION OF SEMI-SOLID MATERIALS

INTRODUCTION

Anaerobic fermentation of municipal and agricultural residues has been used for over forty years (Buswell, 1936; Loehr, 1974; Metcalf and Eddy, Inc., 1972). This operating experience has led to several guidelines, including one that recommends that the total solids concentration of the fermentor be kept under 10 percent (Buswell and Hatfield, 1936). Sewage sludge studies usually use less than 7 percent. However, since most agricultural residues are produced at solids concentrations greater than 10 percent dry solids, it is important to examine the limitation to determine whether it applies to agricultural residue fermentation. The goal of this portion of the study was to determine the feasibility of anaerobically fermenting organics at solids concentrations significantly greater than 10 percent dry weight.

Successful semi-solid fermentation (i.e., use of organics at 10 percent or greater concentrations) would permit higher organic loadings and the design of smaller volume fermentors when compared to conventional high-rate fermentation (digestion) systems. It would also significantly change the water requirements and the final material handling problems. Presently most manure handling operations on farms are solid systems. The equipment and experience with such systems can be easily transferred to the anaerobic semi-solid fermentation process. Also if it is demonstrated that chemical additions will speed up or enhance the fermentation process, the chemical requirement for a high solids system might be less than the amount required for a 10 percent total dry solids system.

Another advantage is that the process yields a product which can easily be utilized as a constituent in animal feeds. This is in contrast to liquid

fermentation systems where the products must be harvested from the liquid effluent by centrifugation or other costly processes (Han et al. 1976).

LITERATURE REVIEW

The development of anaerobic fermentations that occurred in Europe during the 1940's and 1950's were recently reported to use several designs that included semi-solid materials (Jewell, et al., 1976). The fermentation systems included several which were designed to operate with mixtures of animal wastes and dry bedding at total solids contents in excess of 15 percent. Unfortunately, the operating data available for these systems are incomplete.

Nelson (1939) conducted a study on fibrous farm wastes to ascertain the effect of temperature, chemical composition and particle size on the rate and quantity of methane produced by anaerobic fermentation. Boshoff (1963, 1967) considered methods of producing methane on tropical farms using high solids concentration.

There have been several small-scale studies in this country on the feasibility of high total solids concentration anaerobic fermentations, but none have led to a continued study of the process on a larger scale. Schulze (1958) worked with a mixture of air dried sewage sludge and actively digesting sludge placed in batch and batch-fed units. His results indicate that methane is produced at total solids concentrations up to 40 percent. The major problem encountered in this study was the buildup of high volatile acid concentrations at the higher solids concentration.

Keefer's (1947) batch study with sewage sludge was different in that chemical coagulants and a Buchner funnel were used for dewatering. The results indicated that while the sludge containing 25 percent solids produced 8 percent more gas per gram of volatile solids, there was a 10 percent smaller loss in volatile solids content with the drier system. Wong-Chong (1975) has conducted batch and batch-fed studies on anaerobic fermentation of dairy cow manure with a total solids content of 20 percent. His results

indicate that the process produced gas at a rate comparable to what has been reported for conventional (5% TS) systems, 3.3 ft.³ biogas/lb. VS added.

EXPERIMENTAL PROCEDURE

The literature review revealed several successful attempts at batch and batch-fed anaerobic fermentation systems at total solids concentrations greater than 20 percent. An experimental procedure utilizing semi-continuous reactors was devised to further evaluate the process as it applied to dairy cow manure mixed with straw bedding.

Description of the Reactors

The reactors were modeled after the "tip-top" digesters described by Buswell (1936). That design allows the reactors to be placed in different positions to increase mixing (Figure 51). Two cylindrical reactors were used, one with a 4.5 l volume and the second with a volume of 4.0 l. Gas production was measured by a liquid displacement system.

Operation of the Reactors

Both reactors were placed in an environmental chamber that maintained a constant 35°C temperature. Initially the units were started with a 10 percent total solids (TS) mixture of one-half raw dairy cow manure and one-half seed from operating anaerobic fermentors. Over a 20 day period the units were brought to a total solids concentration of 22 percent dry solids by two methods. Reactor 1 was fed a 25 percent total solids mixture of wheat, straw and manure pictured in Figure 52. Reactor 2 was fed a 25 percent total solids mixture of plastic balls (an inert material) and manure. Both reactors were initially operated at a solids retention time (SRT) of 20 days.

Feeding for the semi-solid fermentor in this study was carried out manually in a daily fill and draw manner. The feeding procedure consisted of removing the top of the reactor, mixing the reactor contents by hand, removing a predetermined aliquot from the reactor, replacing the material

Figure 51. Photograph of a bench scale semi-solid fermentor (35°C) with gas collection system.

Figure 52. Close-up photograph of the stabilized effluent material from the bench scale semi-solid fermentor.

with an equal volume of feedstock, manually mixing the feed material into the reactor contents, and resealing the reactor by replacing the plug. Following each feeding the reactor was purged with nitrogen gas to remove oxygen. A similar fill and draw feeding procedure was employed for the liquid control reactor.

Reactor 1 was operated at a 20 day SRT and fed with a 25 percent total solids feedstock for a period of 40 days. The SRT was then reduced to 10 days and the effects on the system were observed. After 25 days of operation at the 10 day retention time with gas production declining the reactor was again operated at a 20 day RT with a 25 percent total solids feed. This mode of operation was continued for 35 days and the reactor returned to the former level of gas production. The reactor was then fed a 35 percent total solids feed mixture and within 25 days gas production dropped over 90 percent and the pH was below 6.0. This study was terminated at this time.

In reactor 2 the blend of plastic balls and manure never became a homogeneous mixture and, after 40 days, with gas production declining, its operation was halted. Reactor 2 was then converted to a control unit with a liquid mixture of wheat straw, dairy cow manure and tap water. Initial total solids concentration was 2.5 percent. Start up was carried out with a mixture containing one half volume of the liquid substrate and one-half seed from the semi-solid reactor diluted to 2.5 percent. The hydraulic retention time (HRT) was 20 days. After 23 days of operation the feed concentration was increased to 5 percent and the reactor was operated in this manner for a 24 day period. In the final stage of operation, the feed total solids concentration was held at 5 percent but the HRT was decreased to 10 days. The reactor was operated for 15 days at these conditions and then terminated.

Data collection included daily gas production values and pH readings. An analysis of the gas composition produced by the semi-solid reactor was performed weekly and total solids content of the semi-solid reactor was monitored.

RESULTS AND DISCUSSION

The data indicate that anaerobic fermentation of animal manure and bedding mixture are possible at high total solids concentration. Table 33 lists the important values for fermentation parameters measured in the reactors used in this study. The semi-solid reactor when operated at a residence time of 20 days with a 25 percent total solids feed concentration, produced 1.60 ℓ gas/ ℓ reactor per day. This converts to 200 ml gas/g VS added (3.25 ft³/lb VS added). The gas composition had 52 percent methane and 48 percent carbon dioxide.

As previously discussed, the bench-scale (4.5 ℓ) semi-solid fermentor (35°C) was operated at a feed concentration of 25 percent TS and at SRT's of 20 and 10 days. Figure 53 shows the variation of pH and gas production during these operating modes. In the first period of operation, lasting 40 days, when the reactor was operated at a 20 day SRT, gas production rates of 6 to 8 liters per day were typically observed. However, during the next 20 days the loading rate was doubled through the reduction of the SRT to 10 days, the gas production rate rapidly deteriorated to around 4 ℓ per day. In the next period of operation, the reactor was returned to a 20 day SRT at 25 percent TS feed, resulting in a recovery of gas production to the 6 to 8 ℓ per day range within 20 days after the change to the lower loading.

After 20 days of stable reactor gas production (20 day SRT, 25% TS feed), the total solids content of the feed was increased to 35 percent. Within 25 days after the feed change, gas production fell from 1.5 ℓ/ℓ reactor/day to 0.1 ℓ/ℓ reactor/day and the pH dropped from 8.1 to 5.8. The gas composition data also indicated a failing reactor. Seven days after initiating the higher total solids feed the CH₄/CO₂ ratio was 49/51. Accurate gas composition measurements were not possible soon after that due to low gas production rates.

An interesting comparison can be drawn between the two separate periods when reactor 1 was subjected to high loading rates. During the period when the loading rate was doubled by decreasing the SRT to 10 days while main-

Table 33. COMPARISON OF DATA FROM LIQUID FERMENTORS AND SEMI-SOLID FERMENTORS.

Parameter	Liquid Reactors			Semi-solid Reactors	
Total Solids Content percent	2.5	5.0	5.0	22.0	22.0
Solids Retention Time days	20	20	10	20	10
pH Range	6.4-6.7	6.6-6.7	6.7-6.8	7.4-8.0	7.2-7.5
Gas Production liters/day	1.45	2.64	3.46	7.20	4.35
Gas Per Unit Volume Per Day liters gas/liter reactors-day	0.36	0.66	0.86	1.60	0.96
Loading Rate grams VS/liter reactor-day	1.10	2.20	4.40	7.86	15.70
1b VS/Ft ³ -day	0.07	0.14	0.28	0.49	0.98
Gas Production Per Substrate Added liters gas/g VS added	0.32	0.30	0.20	0.20	0.06
ft ³ gas/1b VS added	5.12	4.80	3.13	3.25	0.96
Gas Composition CH ₄ /CO ₂	55/45			52/48	47/53

Notes:

1. All gas production data are corrected to STP
2. VS = volatile solids
3. All reactors operated at 35°C

Figure 53. Chronological development of gas production and pH variation from a semi-solid bench scale completely mixed reactor.

taining the feed at 25 percent total solids concentration the gas production rate fell to 0.9 ℓ/ℓ reactor /day. The pH remained above 7.0. However, when the loading rate was raised to a level less than twice the original rate by increasing the total solids content of the feed, within 20 days gas production fell to less than 0.25 ℓ/ℓ reactor/day and the pH dropped below 6.0. Although the data indicate certain loading limitations for the semi-solid fermentor, the moisture content of the reactor material also seems to be a critical factor to performance, especially at reactor solids concentrations in the range of 25-35 percent.

The control reactor was first operated as a semi-solid fermentor with a feed of dairy cow manure and plastic balls. The blend of plastic balls and manure never became a homogeneous mixture and after 40 days, with gas production declining, its operation was halted as described by Figure 54.

The reactor was converted to a control unit (4.0 ℓ volume) with a mixture of wheat straw, dairy cow manure and tap water at a total solids concentration of 2.5 percent with a 20 day SRT. Gas production stabilized at 1.5 ℓ biogas/day with pH ranging from 6.4 to 6.6. The feed mixture was then increased to 5 percent total solids with the SRT remaining at 20 days. Figure 55 indicates the gas production increased to 2.6 ℓ and the pH range increased to 6.6-6.7. Comparable increases were also observed when the loading rate was increased by decreasing the SRT to 10 days, while maintaining the feed mixture at 5 percent total solids. As the loading rate increased the volume of biogas produced/weight VS added decreased and approached the value measured for the semi-solid fermentor. This is illustrated in Table 33.

CONCLUSIONS

Several preliminary conclusions can be drawn from the initial data collected in this short study.

1. Anaerobic semi-solid fermentation of dairy manure and bedding appears to be feasible. Dry solid concentrations in excess of 20 percent are useable.

Figure 54. Gas production and pH variation for a semi-solid bench scale completely mixed reactor with feed composed of manure and plastic balls.

Figure 55. Gas production and pH variation from completely mixed reactor used as a continuous reactor.

2. Gas production rates for semi-solid reactors can be comparable to values observed for liquid systems fermenting the same substrate.
3. An anaerobic semi-solid fermentor when fed a 35 percent total solids substrate displayed a rapid deterioration in gas production rates and pH.

The potential advantages previously discussed in the introduction coupled with the data in this chapter indicate a potential role for semi-solid fermentation as a viable process in agricultural residue management. Much additional studies are needed to define the fundamental and practical limitations of high solids anaerobic fermentation.

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CHAPTER V

MULTIPURPOSE FERMENTOR DEVELOPMENT

INTRODUCTION

The need for an anaerobic fermentor which would be capable of separating more than methane as separated by-products was illustrated in the introductory chapter. Fermentation of materials such as freshly produced dairy wastes results in a certain quantity of energy, protein, nutrients, fiber, and water that have a present combined value of about \$16 per ton of wet (10 percent solids) animal residue processed. This value, of course, assumes a readily available market for such products. The most readily usable product is methane which has a value of about \$3 per ton of animal manure processed. If it were possible to effectively separate these by-products it would significantly improve the economics of agricultural residue fermentation.

A review of the feasibility of recovery of multiple products during methane generation would identify many technical problems. First, the operation involving multiple product recovery would need to be able to handle large volumes of organic residues in order to result in large enough quantities of by-products to make their recovery feasible and economically attractive. For example, small farms may be capable of producing up to one ton of stabilized organic fiber per day suitable for paper production. However, paper mills normally have design capacities to process quantities up to several thousand tons per day. Thus, the first conclusion that becomes obvious in this analysis is that multipurpose reactors would be most valuable in large scale operations where up to several hundred tons per day of organics are produced. This processing rate is more representative of the proposed energy plantations than the existing animal raising operations.

Development of a small fermentor capable of: a) rapidly converting biodegradable organics to methane and carbon dioxide, and b) assisting in the separation of bacteria, fiber and soluble nutrients; were the general objectives of the multipurpose reactor phase of this study. It was felt that these units would be complex and would be applicable to large installations.

The most obvious problem that recovery of fermentor by-products have in common is the liquid-solid separation of the fermenting slurry. The use of conventional technologies such as centrifuging effluents is costly in terms of capital and energy. Thus, the question was raised as to possibilities of using the fermentor to enhance or to actually cause liquid solid separation. One possibility would be to utilize the gas formation phenomenon to create a solids flotation process. This concept formed the basis of one approach to the question of multipurpose reactor development.

The application of the conventional sewage sludge anaerobic digestion technology to anaerobic fermentation of large quantities of organics would result in extremely large and costly installations. Reactors covering 10 acres of land area would be required for beef feedlots of 100,000 head. Thus, the development of high rate fermentors would be essential in demonstrating that the technology could be optimized in these large scale installations. A concept on which the senior author has worked for some years formed the basis of the studies to develop a high rate reactor which could greatly reduce the reactor volume and possibly at the same time separate the single cell protein. This high rate reactor took advantage of the phenomenon of microbial attachment to a surface. By providing a large surface on small particles and using an upflow column, an anaerobic attached microbial film fermentor was developed. Theoretically, this attached microbial film would contain a large fraction of the bacterial protein, and it would be easily separated from the slurry and harvested. This was the second concept examined in this study of the development of multipurpose reactors.

FLOTATION ANAEROBIC FERMENTOR - THERMOPHILIC INTERNAL LIQUID-SOLID SEPARATION CONCEPT

The concept of an anaerobic fermentor which serves also as a flotation unit has been developed and analyzed in this study. This reactor was designed to have the capability of anaerobically fermenting dairy cow manure and accomplishing a partial phase separation of solids from the liquid due to a flotation phenomenon that occurs when gaseous bubbles attach to solid particles within the reactor. This unit was operated in the thermophilic temperature range (50°C). It was hypothesized that at thermophilic temperatures, liquid viscosity would be low and gas production rates would be high enough to promote efficient solids separation through the flotation phenomenon.

The concept of liquid solids separation in anaerobic digestion is not new. For years scum layers have been observed to form in municipal digesters which are unmixed. Other studies have also shown that reactors fermenting both dairy cow manure (Jewell *et al.*, 1977; Morris, 1976) and beef cattle manure (Jewell, 1975) exhibit the same phenomenon. Long fibrous material has been found to float to the top of these digesters and form a thick relatively dry crust layer. In most cases this is considered to be a major operating problem. An example of the small scale flotation problem that can occur is shown in Figures 56 and 57.

Flotation of solids in anaerobic fermentation is due to the production of gas within the reactor. This process is analogous to dissolved air flotation (DAF) which is used in municipal sludge thickening and other liquid-solids separation processes. Parameters such as the weight ratio of air to solids (A/S) and solids loading rates are critical to the DAF process (Eckenfelder *et al.*, 1958) and thus similar parameters should be evaluated when analyzing the ability of the internal liquid-solid separation reactor to produce a float layer.

The specific objectives of this portion of the study were to:

1. determine the feasibility of separating liquids and solids in a high rate thermophilic anaerobic fermentor;

Figure 56. Photograph of a bench scale thermophilic (60°C) control fermentor which has experienced solids separation.

Figure 57. Close-up photograph of the particulate fiber float layer of the bench scale thermophilic control fermentor.

2. determine the effect of various float handling methods on the ability of this reactor to separate liquids and solids;
3. evaluate the effect of gas production rate on liquid-solid separation;
4. determine the operational modes which provide near-optimal liquid-solids separation while maintaining acceptable solids destruction and gas production efficiencies;
5. examine the effect of liquid supernatant recycle on both fermentation and liquid-solid separation.

LITERATURE REVIEW

Separation of solids from the liquid phase is a common occurrence in most poorly mixed anaerobic fermentation systems. The mechanism of anaerobic flotation involves the attachment of evolved gas bubbles to solid particles which provides a buoyant force to bring about flotation through a reduction of the apparent density of the solid phase (Rich, 1961). It is by this mechanism that layers, consisting of greases, fats, oils and organic matter, are formed in unmixed municipal anaerobic digesters.

Mixing of municipal anaerobic digesters was intended, among other reasons, to break up the scum layer which was formed. This was accomplished in two ways: a) the turbulent action of the mixers physically break up the scum layers and keeps this material in suspension, and b) because of the turbulence created, gas bubbles are sheared from the solids and they are no longer able to float.

A float layer analogous to this scum layer has been found to form especially rapidly in thermophilic anaerobic fermentation of dairy cow (Jewell, et al., 1976; Morris, 1976) and beef cattle (Jewell, 1975) manures. The float layer which was formed in these reactors was fibrous in nature and thus is much different than the scum layer found in municipal systems.

The rheological characteristics of non-Newtonian fluids such as domestic sludges and animal manures should be examined prior to a review of mixing requirements within anaerobic digesters fermenting these substrates. The viscosity of liquid slurries is dependent upon temperature, total solids

concentration and the molecular structure of the slurry constituents (Bashford et al., 1969; Kumar et al., 1972). Table 34 presents typical viscosities determined for dairy cow manure (Bashford et al., 1969), beef cattle manure (Kumar et al., 1972), and domestic sludges (Babbitt and Caldwell, 1939; Weber, 1972) at various total solids concentration and temperatures. Similar in behavior to most fluids, as temperature increases the viscosity of these slurries decrease and as the percent total solids concentration of these slurries increases so does the viscosity. It is, however, interesting to note that at similar solids concentrations and temperature dairy cow manure (6.17% TS at 24°C) has a viscosity of 2.68 dynes-sec/cm² while domestic sludge at even a lower temperature (6.7% TS at 12°C) only has a viscosity of 0.28 dynes sec/cm², which is nearly a 10-fold difference. This comparison indicates that although sewage sludge and dairy cow manure may have the same biological characteristics, they will not react the same to physical agitation. Domestic sludge would be a much easier fluid to mix, requiring much less energy to keep the material in suspension. However, once the mixing system is shut off the majority of the particulate solids in the slurry will settle out rapidly. In contrast a homogenous dairy cow manure mixture in a quiescent vessel will require a much greater length of time to settle and unless the slurry is very dilute (<1% TS) no distinct interface between liquids and solids will be formed. Morris (1976) made mention of the fact that mesophilic anaerobic reactors operated at 4 and 6 percent total solids and mixed only once per day, did develop a small layer of settled solids on the bottom of the reactor daily but that units operated at total solids concentrations of greater than 6 percent did not exhibit this phenomena.

The mixing of municipal sewage sludge anaerobic digesters is a recommended practice in nearly all reviews on this subject (EPA, 1974; WPCF, 1968). However, design criteria for adequate and feasible mixing have yet to be established for agricultural residues. The purpose of mixing an anaerobic digester is to: a) break up the scum layer, b) evenly distribute the heat, c) promote distribution of the influent wastes with the digester contents, d) prevent deposition of grit and settleable solids, and e) disperse potentially toxic metabolic products such as volatile acids (Loehr, 1974;

TABLE 34. RHEOLOGICAL PROPERTIES OF ANIMAL MANURES AND DOMESTIC SEWAGE SLUDGES.

Type	T.S. (%)	Temp (°C)	Shear Rate (sec ⁻¹)	Viscosity ₂ (Dyne-sec/cm ²)	Reference
Dairy Cow Manure (Fresh)	7.69	24	30	6.50	Bashford, <u>et al.</u> , 1969
	6.17	24	30	2.68	
	5.15	7	30	3.98	
	4.08	7	30	1.18	
	5.15	42	30	1.82	
	4.08	42	30	0.50	
Beef Cattle Manure	4.0	25	4.41	31.0	Kumar, <u>et al.</u> , 1972
	6.0	25	4.41	62.0	
	4.0	25	11.02	8.3	
	6.0	25	11.02	33.0	
	4.0	25	22.04	8.7	
	6.0	25	22.04	21.0	
Raw Domestic Sewage	6.7	12	-	0.28	Babbitt and Caldwell, 1937
Digested Domestic Sludge	10.0	17	-	0.92	Weber, 1972
	12.0	17	-	0.98	
	14.0	17	-	1.01	
	16.0	17	-	1.13	
	18.0	17	-	1.16	

Wilcox, 1953; Sperry and Pfeiffer, 1953). In unmixed municipal anaerobic digesters the problems of scum formation and grit and solids settling create unmixed or dead spaces within the digester (Enloe, 1959), thus reducing the effective volume of the digester and reducing the hydraulic retention times. These studies have shown that when the units are continuously mixed, higher solids loadings can be achieved (Morgan, 1954) and greater total volatile solids destruction becomes achievable at equivalent retention times (Wilcox, 1953; Rudolfs and Trubnick, 1945). Some studies have indicated that continuous overmixing may be detrimental to the digestion process and have recommended that intermittent mixing be employed (WPCF Manual of Practice No. 16, 1968; Sperry and Pfeiffer, 1953).

In anaerobic fermentation of dairy cow manure used as a flotation process continuous mixing would not be utilized. If the reactor was operated on a semi-continuous basis, wasted and fed once daily, a great deal of mixing might not be necessary to provide good conditions for fermentation and flotation. Removal of float solids and liquid from the reactor could be done separately. Prior to wasting, the liquid portion of the reactor would be mixed to insure a homogeneous effluent and solids could be removed from the float at the same time. This would minimize the occurrence of dead spots within the reactor and the effective volume would not be appreciably different from the total reactor volume. After introduction of the fresh wastes, mixing of the digester contents would also have to be employed to insure proper distribution of the feed throughout the system. Because of the high viscosity of this material, settling of solids should not be of particular concern with this system. Therefore, it is felt that if a system were operated in this manner, much of the mixing requirements would be eliminated and the dual process of efficient anaerobic fermentation and liquid-solid separation could be achieved.

Many factors influence the ability of the DAF units to efficiently separate liquids and solids. The parameters of importance include: a) gas bubble size, b) recycle ratio, c) feed solids concentration, d) detention period, e) air to solids rates, f) type and quality of sludge, g) solids and hydraulic loading rates, and h) use of chemical aids (EPA Process Design Manual, 1974).

The air to solids ratio (A/S) is critical to the operation of the flotation process. It is defined as the ratio of the weight of gas released in the waste to the weight of the solids in the influent. In DAF treatment of sewage sludges, A/S ratios of 0.02 are recommended (Eckenfelder et al., 1958). The higher the A/S ratio the more effective separation becomes (Ettelt, 1964).

The type of sludge and chemical dewatering conditioning aids that are required are related. Sludges which are capable of floating are obviously the ones that can be thickened by this process. The degree to which they can be floated determines the type and quantity of conditioning aids required.

Gas bubble size is important because a higher degree of gas-solid attachment is assured if the bubbles have a high surface to volume ratio. Accordingly, smaller bubbles would theoretically produce more efficient flotation (Katz and Gernopolos, 1967). Formation and coalescence of small bubbles should be easily achieved in anaerobic fermentors since the gas is formed in-situ.

The solids and hydraulic loadings which are recommended for the dissolved air flotation system are 10 to 20 lbs/day/ft² for activated sludge (even higher for primary) and 0.5 to 1.5 gpm/ft² respectively (Katz and Gernopolos, 1967). It is generally felt that as the solids loading rate increases the efficiency of thickening decreases. A 15 foot deep fermentor receiving animal residues at 5 percent solids and operating on a 5-day HRT results in solids and liquid loading rates of 11.7 lbs/ft² day and 0.019 gpm/ft², respectively. Thus, the thermophilic fermentor is operated within the two parameter ranges specified for flotation.

The weight ratio of gas to solids added are significant for anaerobic liquid-solid separation reactors and in this study will be compared to values determined as representative operating criteria for DAF systems. It should be emphasized that unlike the DAF process, the anaerobic liquid-solid separation reactor not only involves the physical operation of gas-bubble interaction but also depends upon biological reaction to produce gas at rates necessary for flotation. For this reason, factors affecting flotation in a natural system in many instances must be considered to be a sum total response from physical and biological operations.

MATERIALS AND METHODS

The laboratory model of the internal liquid-solid separation flotation reactor was designed to represent a vertical section of a full scale unit. The unit was a one foot diameter PVC pipe 9 feet tall (with a liquid height of approximately eight feet) and a liquid volume of 180 liters. Plexiglas plates were bolted to the top and bottom of the unit. In addition, three plexiglas windows were installed at different heights on the unit to allow for visual inspection of the reactor contents. The unit was tested for gas and liquid leaks and then placed into operation. The piping and valving system used on this reactor provided flexibility in mixing and feeding. A Moyno 2L3 progressive cavity pump was used for both purposes.

The entire reactor was placed in a constant temperature environmental chamber. A photo of the chamber and the reactor is shown in Figure 58. Heating of the chamber was accomplished by using twelve infrared heat lamps. These lamps were positioned along the sides of the chamber and each heat lamp had a sheet metal heat reflector in front of it to more evenly distribute the heat within the chamber. Two temperature control units were installed in the reactor itself, one at the top and one at the bottom. The top probe controlled the top six heat lamps and likewise the bottom probe controlled the bottom six lamps independently. Two fans located at the top of the chamber were connected to air ducts and continuously circulated air in the chamber.

Raw dairy cow manure used as feed for this unit was stored in a refrigerator at 4°C to minimize degradation. The required amount of manure was removed from the refrigerator each day and mixed with hot tap water, for dilution to the proper percent solids, before being fed. Prior to feeding the unit was completely mixed by recirculation of the liquid within the system. The pump was allowed to run until approximately 5 volumes of the reactor had passed through the pump.

A prescribed amount of liquids and solids were then wasted from the unit each day, separately from the locations shown in Figure 58. The liquid effluent was allowed to flow by gravity out of the unit through a valve at the

Figure 58. Photograph of the internal liquid-solid separation flotation reactor.

bottom. A plexiglas plate was removed from the top inspection window and float solids were removed manually. After wasting, the unit was fed the raw diluted manure and again mixed for a sufficient length of time to insure distribution of the feed. Photos of the unit and the effluents are shown in Figures 59 to 62.

In one condition liquid effluent was used to dilute the manure rather than hot tap water. Liquid effluent was saved for one day while total solids analysis was performed. The liquid effluent was then used for the next day's dilution. The quantity of raw manure fed was adjusted so that the concentration of the feed to the unit (percent total solids) was the same as in previous test conditions.

Analysis of the influent, liquid effluent, float solids and gas was performed for different parameters and at various times during the course of this study. An analysis and testing schedule for the influent and effluent from this unit is presented in Table 35. Analyses of solids, pH, alkalinity, volatile acids, temperature and gas measurement and composition were all performed according to Standard Methods (1971). Nitrogen analysis was conducted according to MacKenzie and Wallace (1954) and the capillary suction time (CST) test was from Baskerville and Gale (1968).

Completely mixed thermophilic fermentors were used as control reactors for this portion of the study. A full description of these reactors is presented in Chapter 2 of this report.

RESULTS

The internal liquid solids separation flotation anaerobic reactor was operated for a total of 156 days, during which time five conditions were examined until steady state was achieved. Table 36 presents a description of the five conditions at which this system was operated. This reactor was operated with and without recycle. Conditions 1 through 4 were operated on a simple fill and draw basis. The HRT for all of the nonrecycle conditions was 5 days, and the unit was fed a 5 percent total solids manure slurry.

Figure 59. Photograph of the wasting and feeding systems on the internal liquid-solid separation flotation reactor.

Figure 60. Wasting float solids from the top of the internal liquid-solid separation flotation reactor.

Figure 61. Close-up photograph of drained stabilized float solids from the internal liquid-solid separation flotation reactor.

Figure 62. Photograph of stabilized liquid effluent obtained from the internal liquid-solid separation reactor.

TABLE 35. ANALYSIS PROGRAM AND TESTING SCHEDULE FOR INTERNAL LIQUID-SOLID SEPARATION REACTOR.

Analysis Parameter	Influent Feed	Effluent Daily			Effluent Bi-Weekly			Effluent Steady-State		
		Liquid	Solids	Gas	Liquid	Solids	Gas	Liquid	Solids	Gas
Total Solids	x	x	x		x	x		x	x	
Total Volatile Solids	x	x	x		x	x		x	x	
TKN	x							x		
NH ₄ N	x				x			x		
pH	x	x	x		x	x		x	x	
Alkalinity	x				x			x		
Volatile Acids					x			x		
CST	x							x		
Temperature		x	x		x	x		x	x	
Total Gas Measurement				x			x			x
Gas Composition										x

TABLE 36. METHOD OF OPERATION OF INTERNAL LIQUID-SOLIDS SEPARATION FLOTATION ANAEROBIC FERMENTOR.

Condition #	Operating Conditions	HRT Days	SRT Days
1	Operate unit at a 5 day HRT. Remove all float solids formed each day.	5	4.62
2	Operate unit at a 5 day HRT. Remove float solids at a 5 day retention time. Remix float into unit after wasting.	5	5
3	Operate unit at a 5 day HRT. Remove float solids at a 5 day retention time. No remixing of float.	5	5
4	Operate unit at a 5 day HRT. Remove float solids at a 10 day retention time. No remixing of float.	5	7.6
5	Operate unit at a 5 day HRT. Remove float solids at a 10 day retention time. Use effluent for dilution.	5	22.5

The only difference between the test conditions was the length of time the float solids were allowed to remain in the digester. This resulted in different SRT and HRT's. Condition 5, however, was operated in a different manner. The liquid effluent removed from the reactor was used for dilution of the raw manure. When the effluent was removed, total solids analyses were performed. Since one day was required to determine total solids, the liquid effluent removed was saved and used as the next day's dilution. The effluent was then mixed with raw manure to make a 5 percent total solids feedstock and refed to the reactor on a 5-day HRT basis. Because the liquid effluent contained 2.0 to 3.5 percent total solids, the amount of biodegradable solids fed to the reactor in condition 5 was much less than what was fed to other conditions.

When modes of operation in terms of float handling and recycle were altered for each condition, the SRT of the reactor in some cases was altered so that it was no longer equal to the HRT for the system. Table 36 tabulates SRT and HRT values for all five test conditions. The largest difference occurring between the HRT and SRT values was observed during the condition with extensive effluent recycle (condition 5).

All steady state data for the five conditions studied are presented in Tables D1 through D5 of the Appendix. Figure 63 presents a summary of all gas product data developed during this portion of the study.

Flotation Reactor Performance With Respect To Fermentation

The effects of different methods of float solids handling on fermentation can be seen by comparing the results between condition 1 through 4. Table 37 contains a summary of the results of these conditions as well as condition 5 which employed recycle. As seen in Table 37, the influent solids loadings for conditions 1 through 4 were equal, and each of these conditions employed an HRT of 5 days. Since the SRT's for the first three reactor runs were close in magnitude, it would be expected that any differences in volatile solids destruction efficiencies observed between these conditions would be minimal. Accordingly, these three conditions averaged approximately 25% total volatile

Figure 63. Chronological development of gas production in a laboratory-scale thermophilic liquid-solid separation fermentor operated on 5% TS manure feed, 50°C.

TABLE 37. PERFORMANCE OF THE INTERNAL LIQUID-SOLID SEPARATION REACTOR WITH RESPECT TO FERMENTATION PARAMETERS.

Parameter	Units	Condition #1	Condition #2	Condition #3	Condition #4	Condition #5
Influent T.S.	gm/day	1800	1800	1800	1800	400
Reactor loading Rate	gm/l-day	10.0	10.0	10.0	10.0	2.2
Effluent T.S.	gm/day	1398	1455	1398	1347	150
T.S. Reduction	%	22.3	19.2	22.3	25.2	62.5
Influent T.V.S.	gm/day	1590	1616	1597	1598	359
Effluent T.V.S.	gm/day	1170	1236	1172	1115	127
T.V.S. Reduction	%	26.4	23.5	26.6	30.2	65.2
Gas Production*	l/day	192.6	299.2	387.7	433.8	142.5
	l/l reactor day	1.07	1.66	2.15	2.41	0.79
	l/gm VS added	.121	.185	.243	.271	.397
Gas CH ₄	%	54.5 ^x	54.5 ^x	55.6	55.8	52
Comp. CO ₂	%	45.5 ^x	45.5 ^x	44.4	44.2	48
Gas Production* Predicted From T.V.S. Reduction	l/day	385.3	348.6	382.2	432.8	223.1
Effluent T.V.A.	mg/l ⁺	-----	-----	229	458	391

*Corrected to STP - 0°C and 1 ATM
⁺As acetic acid

^xComposition determined as average of other conditions since no data available
¹Gas leaks observed during this stage

solids destruction. However, when the float solids retention time (FRT) was increased to 10 days, thus increasing the SRT to 7.6 days, the total volatile solids destruction increased to 30 percent. This fact is also reflected in the observed gas production data. Condition 4 (10-day FRT) produced the greatest volume of gas of any condition.

The observed gas production values for condition 1 (total daily float removal) are most likely inaccurate since gas leaks were encountered during this condition. This is reflected in the fact that for this condition the gas production calculated from total volatile solid destruction is much higher than the observed gas production. For conditions 2 through 4 these values are within 20 percent of each other. These data tend to indicate that although the float remains separate from the reactor contents, some total volatile solids degradation does occur within the float portion.

Condition 5 employed liquid effluent recycle. Since the recycled liquid had a significant solids content and the total solids loading was maintained constant, the quantity of raw manure feedstock introduced into the reactor was much less for this condition than all others (reduced from 1800 gm/day for conditions 1 through 4 to 400 gm/day for condition 5). Therefore the solids retention time at this condition was increased to 22.5 days. This had the effect of increasing the total volatile solids destruction to 65.3 percent which is more than double that of any condition. An estimate of the total volatile solids destruction based on measured gas production data shows that 41.2 percent of the total volatile converted to methane and carbon dioxide. This is more likely according to the biodegradability fraction measured in the control units (see Chapter 2).

Gas composition for all conditions ranged from 52 to 56 percent methane. This compares favorably with data gathered for the control thermophilic reactors (Chapter 2). The total volatile acid analysis (TVA) showed that this unit operated between 300 and 500 mg/l as acetic acid for all conditions. This is considerably less than what was observed in the control reactors.

Flotation Reactor Performance with Respect to Liquid-Solid Separation

Table 38 summarizes data gathered at steady state on the liquid-solids separation properties of this reactor. Increasing the retention period of the float in the reactor 5 days (conditions 2 and 3) produced a much drier float than removing all float formed daily (condition 1). An increase from 11.5 percent to 13.7 percent total solids was observed. No significant difference in the float solids concentration was found when it was retained for greater than 5 days or when it was remixed with the rest of the reactor contents. At the five day FRT approximately 9 percent of the reactor volume was occupied by this material whereas at the 10 day FRT the amount of float increased proportionally to approximately 18 percent of the total volume. When recycle was employed (condition 5) both the volume of float and the percent total solids of the float decreased dramatically. The decrease in volume was apparently

TABLE 38. ABILITY OF THE LIQUID-SOLIDS SEPARATION FLOTATION REACTOR TO SEPARATE AND CONCENTRATE SOLIDS.

Condition #	% of Digester Occupied by Float	Quantity of Float Removed Daily (ℓ)	Solids Concentration	
			% TS Float	TVS Float,% TS
1	2.50	4.5	11.50	92.07
2	9.79	3.5	13.70	91.87
3	8.33	3.0	13.61	90.13
4	18.22	3.5	13.47	90.10
5	3.29	0.5	11.79	87.89

due to the decrease in rate of gas production due to the low biodegradable solids feed rate. Since the volume of the float was decreased, the depth of the float was only a few inches (approximately 3.0 inches) during this condition.

The float material was observed to be odorless after it was stored for long periods of time either in the open air or in closed containers. In addition, the liquid did not drain from the float when placed on screens, but it dried due to evaporation.

DISCUSSION

Reactor Performance with Respect to Flotation

A number of factors seem to influence the degree of floating material which can be developed by the thermophilic flotation reactor, and the percent solids of the float material. These factors include the method of mixing of the reactor and the depth of the float material, the rate of gas production, the gas production to solids loading ratio, and both solids and hydraulic loadings.

Three different methods of float handling were employed in this study, variations of the FRT, remixing the float with the remainder of the reactor, and recycling of liquid effluent for feed dilution. In general, when the total float volume was removed on a daily basis the total solids concentration was found to be lower (11.5 percent total solids) than when the FRT was maintained at 5 and 10 days (approximately 13.5 percent total solids). However, no distinct advantage was found in improved float dryness when the FRT was increased from 5 to 10 days since the float remained at approximately 13.5 percent total solids for both conditions.

Conditions 2 and 3 were operated in a similar manner with FRT and an HRT of 5 days. In condition 2 the float solids remaining in the unit each day were remixing. As can be seen there is no difference in the percent total solids of the float between these two conditions. Thus, remixing of the float daily had no effect on the process.

In condition 5 recycling of the effluent was employed. Because of the way this condition was handled the amount of fresh solids introduced to the unit was much lower than that for other conditions. Therefore, the amount of float within the unit was small. As can be seen from the data, the percent total solids in the float decreased substantially in this condition as compared to conditions 2 through 4.

A significant parameter in controlling the percent total solids in the float appeared to be the depth of float (Figure 64). In the conditions where the float was maintained at a depth of 8 inches or greater the float was found to be drier than with a smaller amount of float. It is felt that the float acts somewhat like a sponge since it is much drier than the liquid below it. A moisture gradient is established and thus the lower regions of the float have a lower percent total solids concentration than the upper fractions. In conditions 1 and 5 the float depth was small (in the order of 2 and 3 inches). Thus the float removed contained more moisture than the float removed at the other conditions. In these conditions only the top layer of the float was removed and the lower, wetter, portions of the float were allowed to remain in the digester. Increasing the float depth from 8 inches to 12 inches did not create drier float solids. This is possibly due to the fact that the moisture in the float was in equilibrium with the moisture in the gas above the float.

The effect of gas production on percent total solids in the float is depicted in Figure 65. It would be logical to expect that as more gas is produced more solids would be floated. None of our data however, shows this to be the case. The quantity of float present seems to be more a function of the solids management techniques employed. However, the float produced at the higher gas production rates was found to be drier than the float produced at the lower gas production rates. Gas production rates could definitely affect the dryness of the float due to moisture removed by evaporation. As more gas is produced, more moisture can be carried out of the reactor by the gas. There was no clear indication of this being the case. Therefore, with the present data available, it is difficult to determine whether float depth or gas production was the dominant factor in producing a dry float, although both factors were probably involved.

Figure 64. Relationship between the depth of the float solids and the percent total solids within the float of a laboratory-scale thermophilic liquid-solid separation fermentor operated on 5% TS manure feed, 50°C.

Figure 65. Relationship between gas production and total solids concentration in the float of a laboratory-scale thermophilic liquid-solid separation fermentor operated on 5% TS manure feed, 50°C.

Figure 66 presents the effect of the G/S ratio on the percent total solids in the float. The G/S ratio is analogous to the A/S ratio used in the DAF sludge thickening process. G/S ratios in the order of ten times greater than normal A/S ratios were found to occur in this unit. It seems that a relationship between G/S ratios and percent total solids in the float might exist, except for the value for condition 5. This point shows that any relationship between these two parameters is probably due to either the depth of the float or moisture removal due to the gas production rate rather than a dependence upon the G/S ratio. If this were not the case a percent total solids of at least 13.5 to 14.0 should have been found in the float for condition 5. Because the G/S ratios determined for all conditions were so much greater than conventional A/S ratios, an increase or decrease in the G/S ratio at these high gas production rates was probably an insignificant factor in influencing the quantity or dryness of the float produced. In other words, all the material that could be floated was floated and only G/S ratios considerably lower than those which occurred would have an effect on the float material. However, because gas production within the anaerobic reactors serves the dual purpose of floating material and withdrawing moisture from the reactor, the gas production rate rather than the G/S rates could have a significant effect on the dryness of the float due to moisture removal.

Solids loadings of $5.04 \text{ lbs/day/ft}^2$ (condition 1 through 4) and $1.12 \text{ lbs/day/ft}^2$ (condition 5) were employed in this study. These values are well below the maximum loadings recommended for DAF thickening of difficult sludges ($10\text{-}20 \text{ lbs/day/ft}^2$). Likewise the hydraulic loadings recommended for DAF thickening ($0.5 - 1.5 \text{ gpm/ft}^2$) were much lower in this study. The fact that both these loadings are well below what is recommended for DAF thickening processes is beneficial to flotation by this system.

Table 39 presents a summary of data on the depth of the float, G/S ratios and solids and hydraulic loading rates.

Figure 66. Effect of gas to solids ratio on the float solids concentration in a laboratory-scale thermophilic liquid-solid separation fermentor operated at 5% TS manure feed, 50°C.

TABLE 39. SUMMARY OF DATA AFFECTING SOLIDS FLOTATION IN THE INTERNAL LIQUID SOLIDS SEPARATION REACTOR.

Condition	HRT days	SRT days	FRT days	Depth of Float (inches)	G/S Ratio (lbs/lbs)	Solids Loading lbs/day/ft ²	Hydraulic Loading gpm/ft ²
1	5	4.62	1	2.4	0.137	5.04	0.0084
2	5	5	5	8.0	0.213	5.04	0.0084
3	5	5	5	9.4	0.273	5.04	0.0084
4	5	7.6	10	17.5	0.306	5.04	0.0084
5	5	22.5	10	3.18	0.468	1.12	0.0084

Reactor Performance With Respect to Gas Production

The internal liquid solids separation reactor achieved approximately 25 percent total volatile solids destruction while operating at a 5-day HRT and approximately a 5-day SRT (conditions 1 through 3). Although this unit was operated 10°C cooler than the thermophilic control reactor. This compares favorably to the range of total volatile solids destructions observed in the completely mixed reactors (see Chapter 2). As the SRT of the liquid-solid separation reactor was increased from 5 days to 7.60 days, the total volatile solids destruction increased to approximately 30 percent. This would indicate that although the float solids exist in a semi-solid state, fermentation proceeds within this zone. In condition 5 where recycle was employed the total volatile solids destruction increased to 42 percent (based on gas production correlation). The effect of SRT on total volatile solids destruction is shown in Figure 67. Table 40 summarizes the total biodegradable volatile solids (TBVS) destruction efficiencies assuming that the manure was only 50 percent biodegradable.

No studies were initiated to determine the effect which minimal mixing played upon the reactor. The reactor was mixed twice daily, once prior to wasting and once after feeding. It is possible that more frequency mixing of the reactor may have improved the solids destruction efficiency. However, after initial start-up of the reactor no solids deposition was noticed. Solids destruction evidently did occur within the float zone of the reactor as indicated by the improvement in total volatile solids destruction in condition 4 over conditions 1 through 3.

SUMMARY AND CONCLUSIONS--FLOTATION REACTOR

The thermophilic flotation reactor concept was shown to be capable of achieving stable fermentation and solids flotation in the same reactor. The following conclusions were drawn from this study:

1. The reactor was able to achieve a flotation solids concentration of between 13.0 and 14.0 percent total solids when fed total solids of 5 percent dry weight and at float retention times of 5 and 10 days.

Figure 67. Effect of solids retention time on total volatile solids destruction achieved by laboratory-scale thermophilic liquid-solid separation fermentor operated on 5% TS manure feed, 50°C.

TABLE 40. SUMMARY OF TOTAL BIODEGRADABLE VOLATILE SOLIDS DESTRUCTION EFFICIENCY FOR THE THERMOPHILIC INTERNAL LIQUID SOLIDS SEPARATION REACTOR.

Condition #	HRT days	SRT days	TBVS Destruction (%)
1	5.0	4.62	52.8
2	5.0	5.00	47.0
3	5.0	5.00	53.2
4	5.0	7.60	60.4
5	5.0	22.50	82.4*

*Calculated from observed gas production.

2. The solids conversion to gas efficiency with respect to total volatile solids destruction was found to be 25 and 30 percent for the conditions where the float confirmed 13.0 to 14.0 percent total solids. Daily float removal did not significantly reduce gas production.
3. Operation of the liquid portion of the reactor at a 5-day HRT and the float retention time at 10 days (condition 4) achieved comparable or better liquid-solids separation efficiencies, as expressed in percent TS in the float, than all other conditions examined. Of all conditions operated without recycle it exhibited the highest TVS destruction efficiencies, and thus it is considered the best mode of operation at this time.
4. Because recycling the effluent and long FRT's (condition 5) do achieve substantially higher TVS destruction efficiencies than the above condition, but exhibited poorer liquid-solid separation capabilities, it is felt that the use of recycle should be re-examined more carefully to determine if this mode of operation could be utilized to achieve optimal efficiencies with respect to both parameters.
5. The percent total solids within the float was most affected by the depth of the float layer and the gas production rates at which the reactor was operating.
6. Gas to solid ratios within the reactor are found to be greater than the A/S ratios used in dissolved air flotation (DAF) thickening by a factor of 10. Therefore, fluctuations in the G/S ratios within this reactor probably had little effect on the performance of this unit.
7. Both hydraulic and solid loadings in this unit were much lower than those recommended for DAF thickening which had a beneficial effect on this process.
8. When effluent recycle was employed solids destruction efficiency increased significantly. However, the percent solids in the float were decreased. It is believed that this was due to the low solids loadings employed and the resultant small float layer produced.

9. More research is needed to determine if high solids loadings and liquid recycle can be combined to produce better flotation characteristics and equivalent reactor performance.

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ANAEROBIC ATTACHED FILM EXPANDED BED REACTOR

The second attempt to illustrate the potential for improvement of anaerobic fermentation emphasized the development of a high rate reactor capable of separating the viable microbial community to enable high rates of liquid through-put.

The attached film expanded bed (AFEB) process consists of an upflow filter bed composed of small inert particles that are coated with a film of microorganisms capable of metabolizing organic materials as the feedstock is passed through the reactor. The AFEB reactor liquid contents are recirculated to achieve the upflow velocities necessary for slight expansion of the media. The principle upon which the success of this system is based is that functional microorganisms may attach themselves to inert particles much larger and heavier than they are, thereby allowing the reactor to be operated at high volumetric loading rates without danger of biomass washout.

In comparison with the media used in trickling filters, 1 to 3 inch gravel or slag, the material used in AFEB systems is small (1.0 mm in diameter), making a high surface area available for active biological film formation.

Attached film systems, such as the anaerobic packed column and the AFEB, have been used in the treatment of various types of wastes in the past. Packed column technology has been applied to the anaerobic treatment of sewage (Coulter et al., 1956; Pretorius, 1971, 1972), industrial strength soluble wastes (McCarty, 1966; Young and McCarty, 1969), and in the aerobic nitrification process (McHarness et al., 1975). AFEB systems have been used in the aerobic and anaerobic treatment of sewage strength wastes (Jewell, 1974; Atkinson and Knights, 1975; Leuschner, 1976), anaerobic biological denitrification (Jeris et al., 1974; Jeris and Owens, 1975), and in the chemical

engineering industry (Atkinson and Davies, 1972). However, at no time has this technology been applied to the anaerobic fermentation of a high-strength particulate residue such as dairy cow manure as is reported here. This concept has been under development by the senior author since 1973.

The specific objectives of this portion of the study are to:

1. Define the feasibility of the anaerobic AFEB reactor for the high rate fermentation of highly concentrated, particulate organic residues;
2. Establish the accuracy of the assumption of separation of a viable microbial population on inert media surfaces to accomplish long SRTs with short HRTs;
3. Operate the system over a range of influent substrate concentrations and HRTs to determine its capabilities.

LITERATURE REVIEW

The widespread use of anaerobic digestion of sewage sludge over the past 50 years has allowed researchers to become familiar with the behavior and limitations of this process. The most notable limitation is the slow growth rate of anaerobic microorganisms which makes it difficult to operate a single-pass high rate fermentor at hydraulic retention times lower than 10 days. One way of preventing washout is to attach the viable microorganisms to surfaces that remain in the reactor. Research has been conducted on a number of anaerobic fixed film reactors, the earliest of which was the submerged anaerobic filter with static surfaces.

As first tested, the anaerobic filter was a submerged column packed with rocks which provided a support surface for the attachment of microorganisms. A schematic of the anaerobic filter is shown in Figure 68 (McCarty, 1966). Coulter et al., (1956) developed a two-stage system for the treatment of sewage. This system consisted of a sludge contact chamber and a packed rock contact chamber. The total HRT within the system was approximately 35 hours with only a 2.5-hour HRT required for the filter portion of the system.

Figure 68. The anaerobic submerged packed column.

This unit was not supplied with any external heat, but rather was operated at simulated winter and summer ambient temperature conditions. Results from this study showed that during winter conditions (4°C) approximately 67 percent of the BOD₅ and greater than 90 percent of the suspended solids were removed. Operation at summer conditions (20°C) showed a significant increase in BOD₅ removal efficiency of 81 percent, while suspended solids removal remained at greater than 90 percent.

A very similar system to the one proposed by Coulter et al., was developed by Pretorius (1971, 1972) for the treatment of raw sewage. This system was also a two-stage process consisting of a sludge contact chamber and a packed anaerobic filter which was termed a biophysical filter. This unit was operated at 20°C and the HRT of the entire system was varied between one and two days. Sewage used in this study had a soluble COD of 500 mg/l and total treatment efficiencies of 80 to 90 percent were observed (Pretorius, 1971). The sludge contact chamber provided for conversion of organics to volatile fatty acids while in the biophysical filter the acids were converted to carbon dioxide and methane gas.

McCarty (1966) and Young and McCarty (1969) have investigated the use of the anaerobic packed columns (Figure 68) for the treatment of slightly more concentrated soluble wastes in the range of 1500 to 3000 mg/l COD. These packed columns differed from those discussed by Coulter or Pretorius in that only an anaerobic filter was used, thus excluding the use of a sludge contact chamber.

In the early experiments conducted by McCarty (1966), four particular soluble wastes were used as the substrate for the anaerobic filter; a) methanol, b) methanol plus acetate, c) methanol plus acetate plus propionate, and d) acetate plus propionate. The waste strength was varied between 2140 and 2650 mg/l COD, and hydraulic retention times of 6 and 12 hours were employed. Thus organic loadings varied between 107 and 226 lbs COD/1000 ft³/day (McCarty, 1966). COD removal efficiencies were found to vary between 74 and 88 percent for all conditions. In addition, analysis of effluent solids and solids accumulated within the system showed that the SRT of the unit was greater than 100 days.

In a similar study by Young and McCarty (1969), a protein-carbohydrate synthetic substrate was utilized. Influent COD concentrations of 1500 and 3000 mg/l and HRTs of between 4.5 and 72 hours were employed. At the lowest loadings of 26.5 and 53.0 lbs COD/1000 ft³/day, COD removal efficiencies of greater than 90 percent were observed, whereas at the highest loadings of 212 lbs COD/1000 ft³/day the efficiency dropped to between 36.7 and 63.0 percent. In this study, calculations of SRT showed that it varied from 84 to 665 days (Young and McCarty, 1969).

The most obvious disadvantage of the submerged filter is the probability that as biological sludge accumulates on the static surfaces, clogging may occur. Also the large static packing material limits the quantity of viable organisms to that which could exist in suspended systems.

A summary of data compiled on studies concerning the use of anaerobic filter processes is presented in Table 41.

The success of the anaerobic packed bed has led other investigators to the use of fluidized or expanded bed technology for the removal of contaminants from wastewaters. A schematic of this type of process is shown in Figure 69. Jeris et al., (1974) cited four advantages in using this technology over the packed bed systems: "a) greater surface area available for growth per unit of reactor volume, b) very small headloss, c) no danger of clogging, and d) easier carrier removal procedure." The primary difference between the term "fluidized" and expansion refers to the change in reactor volume when the unit is in operation. Fluidization refers to a large increase in reactor volume (up to 100 percent), whereas the expanded beds only increase in volume 10 to 20 percent over the static non-flowing volume.

In two related studies by Jeris et al., (1974), fluidized bed technology was used for denitrifying secondary sewage effluent. The unit employed was columnar in shape and both activated carbon and sand (Jeris et al., 1974; Jeris and Owens, 1975) were used independently as support media for the microorganisms. The hydraulic retention times within the reactor varied from 3 to 10 minutes. Investigations on a pilot scale unit where a HRT of 6.5

TABLE 41. REVIEW OF OPERATIONAL PARAMETERS AND PERFORMANCE OF STUDIES
CONCERNING ANAEROBIC PACKED COLUMN SYSTEMS.

Influent Feed Constituent	Loading Rate lbs/ft ³ /day	HRT hrs	Influent Waste Concentration mg/l	Removal Efficiency %	Reference
Sewage	0.0084 (BOD ₅)	35	200 (BOD ₅)	81.0 (BOD ₅)	Coulter <u>et al.</u> , 1956
Sewage	0.0167 (COD)	45	500 (COD)	78.0 (COD)	Pretorius, 1971
Sewage	0.0226 (COD)	33	500 (COD)	89.0 (COD)	Pretorius, 1971
Sewage	0.0312 (COD)	24	500 (COD)	90.0 (COD)	Pretorius, 1971
Methanol	0.107 (COD)	12	2140 (COD)	84.0 (COD)	McCarty, 1966
Methanol + Acetate	0.123 (COD)	12	2360 (COD)	74.0 (COD)	McCarty, 1966
Methanol + Acetate	0.123 (COD)	12	2360 (COD)	78.0 (COD)	McCarty, 1966
Methanol + Acetate	0.123 (COD)	12	2360 (COD)	82.0 (COD)	McCarty, 1966
Methanol + Acetate + Propionate	0.131 (COD)	12	2650 (COD)	86.0 (COD)	McCarty, 1966
Methanol + Acetate + Propionate	0.266 (COD)	6	2650 (COD)	80.0 (COD)	McCarty, 1966
Acetate + Propionate	---	6	---	---	McCarty, 1966
Acetate + Propionate	0.224 (COD)	6	2240 (COD)	82.0 (COD)	McCarty, 1966
Protein-Carbohydrate	0.0265 (COD)	36	1500 (COD)	92.1 (COD)	Young and McCarty, 1969
Protein-Carbohydrate	0.053 (COD)	18	1500 (COD)	91.5 (COD)	Young and McCarty, 1969
Protein-Carbohydrate	0.106 (COD)	9	1500 (COD)	79.3 (COD)	Young and McCarty, 1969
Protein-Carbohydrate	0.212 (COD)	4.5	1500 (COD)	36.7 (COD)	Young and McCarty, 1969
Protein-Carbohydrate	0.0265 (COD)	72	3000 (COD)	93.4 (COD)	Young and McCarty, 1969
Protein-Carbohydrate	0.053 (COD)	36	3000 (COD)	88.4 (COD)	Young and McCarty, 1969
Protein-Carbohydrate	0.212 (COD)	9	3000 (COD)	63.0 (COD)	Young and McCarty, 1969

Figure 69. Schematic of the expanded bed reactor.

minutes was employed resulted in 99 percent of the nitrite and nitrate nitrogen being removed (Jeris and Owens, 1975). However, this removal efficiency was highly dependent upon the addition of methanol, in stoichiometric amounts, as a carbon source. When the addition of methanol was not enough, efficiency decreased proportionally. Tolerance to short HRTs in the fluidized bed denitrification process were most likely due to the large mass of microorganisms which could be concentrated within the fluidized bed, 20,000 to 40,000 mg/l (Jeris and Owens, 1975), and to the use of methanol, which served as a readily available carbon source.

In studies by Jewell (1974) an aerobic attached film expanded bed (AFEB) unit was used in the treatment of domestic sewage strength wastes. The media used as a support surface for microbial attachment was spent ion exchange resin and PVC commercial filter media. All particles had a size of approximately 1 mm in diameter. Aeration of the system was accomplished by utilizing an external aerator which kept the dissolved oxygen of the liquid within the AFEB high enough to support an aerobic microbial population. Loadings as high as 700 lbs COD/1000 ft³/day were employed at retention times of 8 hours. Removal efficiencies of 80 to 90 percent were reported (Jewell, 1974).

Leuschner (1976) used a similar AFEB unit for the anaerobic treatment of domestic sewage strength synthetic substrate as well as primary effluent. The media that comprised the expanded bed was the same as that used by Jewell (1974). Synthetic substrate loadings varying from 0.052 to 0.552 lbs COD/ft³/day were employed at HRTs ranging from 3 to 12 hours. COD removal efficiencies varied from a low of approximately 50 percent for the highest loadings to a high of 90 percent for the lower loadings. Primary settled sewage was also used as a substrate and COD removal efficiencies of 75 percent were observed at a loading rate of 0.02 lbs COD/ft³/day and a HRT of 12 hours.

In addition to waste treatment studies, the fluidized bed has also been under investigation for its potential applications in the fermentation industry. Atkinson and Knight (1975), for instance, developed a "completely mixed microbial film fermentor" (CMMFF) for the continuous fermentation of brewers yeast. Scott and Hancher (1976) have investigated "tapered fluidized bed" technology

in the immobilized enzyme conversion of lactose to glucose and galactose and in enzyme-catalyzed hydrogen production, as well as biological denitrification.

In summary, the literature clearly shows that anaerobic contact processes such as the packed column and the AFEB have been used in the waste treatment field for the anaerobic fermentation of low strength soluble organic materials and that it is a highly promising technique. Utilization of these types of systems on high strength particulate wastes has not yet been investigated. Only conventional completely mixed anaerobic fermentation technology has been applied to the concentrated solids residues.

The attached film expanded bed process has been used in this study for the anaerobic fermentation of dairy cow manure diluted to 1 to 2 percent total solids. The high concentrations of biological solids which have been shown to persist in the anaerobic AFEB process under stresses of high organic and hydraulic loading rates suggested a need to examine the potential of this system for the high rate fermentation of dairy cow manure.

MATERIALS AND METHODS

Description of System

A schematic of the anaerobic AFEB reactor and associated appurtanences is shown in Figure 70. The reactor itself was constructed from a plexiglass cylinder, 8.89 cm in diameter and 132 cm in height. The empty volume was 8.2 liters. A mixture of expended ion exchange resin and PVC particles about 0.5 to 1.5 mm in diameter occupied about 3 liters of the 8.2 liters. The surface area for microbial attachment contributed by the media is about 5.6 m²/l of media (Leuschner, 1976). The entire anaerobic AFEB system was situated in an environmental chamber where the temperature was held constant at 35°C ($\pm 1^\circ\text{C}$). The influent feed container was continuously mixed while being pumped into the reactor to insure that a homogeneous feed resulted. The manure was fed to this unit on a semi-continuous basis. The number of times the unit was fed during the day was determined by the feed rate. When the unit was operated at approximately a 1-day HRT, the unit was fed once daily; likewise, when the

Figure 70. Schematic of the anaerobic attached film expanded bed reactor and associated experimental equipment.

unit was operated at approximately a 3-hour HRT, the unit was fed eight times daily at 3-hour intervals. The feed entered the bottom of the reactor and passed through the bed in an upwards direction.

The inert particle bed was continuously expanded by recycling liquid from the top of the unit back through the bottom. The rate at which the liquid was recycled was 650 ml per minute, thus approximately 115 system volumes were pumped through the reactor per day. This was not a critical variable, but high recycle rates were required to obtain uniform flow distribution in this small reactor and to fluidize the resin particles.

Liquid effluent and the gas produced by the system both flow out of the reactor at the top of the unit and into an enclosed effluent container. The effluent container is drained from the liquid effluent after each hydraulic retention period. The gas produced by the reactor passes through the enclosed effluent container and into a wet test gas meter.

Within the anaerobic AFEB reactor, two distinct zones are considered to exist, these being the expanded bed zone and the clarification zone. The expanded bed zone constituted the bottom 3.8 liters of the column. This volume represented the degree of expansion the media experienced. It is considered that within this volume of the reactor the majority of biological conversion of organic material to carbon dioxide and methane gas occurred.

The second zone within the anaerobic AFEB system is the clarification zone. This comprised approximately 3.2 liters in volume. It is assumed that relatively little biological activity occurred in this portion of the system. Its main function is to act as a settling chamber for any media particles escaping from the bed.

The manure used as the feed for this system was initially diluted to 6 percent total solids. This manure slurry was then passed through a coarse screen to remove large particles such as long fibers, undigested corn kernels, pebbles, etc. The purpose of removing this material was to insure that large fibrous particles would not clog feed lines. The strained manure was

subsequently stored in a refrigerator at 4°C to minimize any degradation of the waste. Upon feeding this manure to the reactor it was again diluted to the desired percent total solids and placed in the influent feed container.

A photograph of the entire anaerobic AFEB system is shown in Figure 71. Figure 72 shows the inert particles coated with a microbial film.

A control reactor was operated with the same feed, but under HRT values which would insure stable operation of a CSTR reactor. The control reactor was a 3.5 liter unit which was fed on a semi-continuous fill and draw basis daily. This reactor was mixed twice daily, prior to wasting and after feeding. Gas produced by this reactor was measured by a liquid displacement system. This reactor received the same strained manure which served as the feedstock for the anaerobic AFEB unit. The control reactor was kept in the same environmental chamber as the anaerobic AFEB unit which was maintained at 35°C $\pm$ 1°C.

Laboratory Analyses and Testing Schedule

Analyses on influent and effluent samples from the anaerobic AFEB reactor were performed for a variety of parameters at various time intervals. The influent was analyzed every time a new batch of feedstock was prepared while effluent samples were analyzed in accordance with the analysis schedule shown in Table 42.

All wet analyses applied in this study were performed according to Standard Methods (1971) with the exception of the ammonia and TKN nitrogen determinations which were performed according to the method described by MacKenzie and Wallace (1954).

All analyses performed on the anaerobic AFEB unit were similarly performed on the control reactor, thus Table 42 applies to both the completely mixed control reactor as well as the anaerobic AFEB unit.

Figure 71. Photograph of the anaerobic attached film expanded bed reactor.

Figure 72. Close-up photograph of fluidized inert media particles coated with a microbial film in the reactor zone of the anaerobic attached film expanded bed reactor.

TABLE 42. ANALYSIS SCHEDULE FOR THE EXPANDED BED REACTOR.

Parameter	Influent Analysis of Each Batch	Daily Eff. Analysis	Bi-Weekly Eff. Analysis	Steady State Analysis	Daily Gas Analysis
Total Solids	X		X	X	
Total Volatile Solids	X		X	X	
Total COD	X			X	
Soluble COD	X			X	
Volatile Acids			X	X	
NH ₄ -N	X		X	X	
TKN	X			X	
pH	X	X	X	X	
Alkalinity	X		X	X	
Gas Measurement					X
Temperature		X	X	X	

In addition to the above analyses, serum bottle tests were performed to determine relative microbial activities within different sections of the anaerobic AFEB reactor. Three types of samples were withdrawn for this type of analyses, liquid from the clarification zone, liquid and media from the expanded bed zone, and media from the expanded bed zone. Clarified liquid and expanded bed media samples were directly pipetted from the reactor and put into serum bottles. The serum bottles were free of oxygen since they were gassed with nitrogen for a few minutes prior to injection of the liquid sample. Each sample was 20 ml in volume. The media sample was removed from the reactor with a pipet. A 20 ml volume sample of media and liquid was removed and placed in a beaker. The sample was then washed with distilled water until only media remained. Distilled water was then added to give this sample a volume of ml and then transferred to the serum bottle. Each sample was inoculated with a known amount of substrate. Both the effluent and expanded bed media samples were inoculated with a 5 ml solution containing 1000 mg/l of acetic acid and 1500 mg/l of sodium bicarbonate. The media sample was inoculated with a 5 ml solution of 1500 mg/l of acetic acid, 1500 mg/l of sodium bicarbonate, and supplemental nutrients of nitrogen (NH_4Cl), phosphorus, and potassium (K_2HPO_4 and KH_2PO_4).

These samples were then incubated in a water bath for 24 hours at 35°C. A syringe was inserted in the top of the bottle and used to collect gas. Gas volumes were recorded for these samples and gas composition was then analyzed.

RESULTS AND DISCUSSION

The anaerobic AFEB reactor was operated for a total of 190 days in this study. During this period, five different conditions were examined, each until steady state was achieved. A summary of the operating conditions under which this unit was analyzed is presented in Table 43.

TABLE 43. DESCRIPTION OF FIVE CONDITIONS AT WHICH THE ANAEROBIC AFEB WAS OPERATED.

Condition No.	Description	HRT* days	Solids Loading*
1	1% T.S. Feed, Fed 3.5 l/day, fed once every 24 hours	1.2	8.34** (0.52)†
2	2% T.S. Feed, Fed 3.5 l/day, fed once every 24 hours	1.2	16.7 (1.04)
3	2% T.S. Feed, Fed 7.0 l/day, fed once every 12 hours	0.6	33.4 (2.08)
4	2% T.S. Feed, Fed 14.0 l/day, fed once every 6 hours	0.3	66.7 (4.16)
5	2% T.S. Feed, Fed 28.0 l/day, fed once every 3 hours	0.15	133.0 (8.32)

* HRT and solids loading calculations were based on the liquid volume of the reactor, 3.8 l.

** Kg T.S./m³/day

† (lbs/ft³/day)

†† As a comparison, conventional high rate municipal digesters are usually designed for 10 to 20 days HRT and loading rates of 0.10 to 0.50 lb total solids/ft³/day.

TABLE 44. RESULTS OF SERUM BOTTLE ANALYSIS.

Parameter	Media & Liquid Sample	Washed Media Sample	Liquid Sample
Gas Production (l/l sample-day)	0.57	0.28	0.25
Gas Composition			
CH ₄	66.5	77.5	54.7
CO ₂	33.5	22.5	45.3
CH ₄ Production (l/l sample-day)	0.38	0.22	0.14

Serum Bottle Studies

Serum bottle analysis described in the previous section was performed while the unit was operating at a 2 percent total solids feed and at a flow of 3.5 liters per day (condition 2). The purpose of this analysis was to determine qualitatively which portion or portions of the reactor was responsible for the destruction of organics and production of methane gas. This would assist in establishing what the reactor volume was within the system. From a knowledge of the reactor volume, hydraulic retention times and organic loading rates were established for all conditions at which this unit was operated. In addition, the data on gas composition was used as an indication of the quality of the gas produced by this reactor. Serum bottle test results on clarified liquid, media and liquid, and washed media samples (run in duplicate) are presented in Table 44.

Lower gas production rates per volume of reaction vessel were observed in the serum bottles because the samples were taken from the parent reactor just prior to feeding and because organic feed concentrations introduced into the serum bottle during testing were a lower than the substrate concentrations maintained in the anaerobic AFEB reactor. Another possible explanation of the lower serum bottle gas production rates in the serum bottles is that upon introduction into the quiescent test vessels, the media quickly settled to the bottom, allowing only the top layer of the media to come into contact with the substrate in the bulk liquid.

Nevertheless, the serum bottle test results have their real value in the comparative analysis of gas production rates among various sections of the anaerobic AFEB reactor sampled, providing some basis for identifying the major location of active biomass in the fermentor. The data in Table 44 indicate that on a per reactive volume basis, gas production rates from the expanded media portion of the anaerobic AFEB reactor are far greater than the gas production rates realized from the clarified zone of this fermentor. In light of these data, it is apparent that the reactive section of the AFEB unit should be considered to be the expanded bed. It is not unreasonable to expect that, had all of the media surfaces been allowed to come into intimate

contact with the substrate in the serum bottle, the media samples tested would have performed at higher rates. Exposure to oxygen could also have limited the rates.

The serum bottle data also shows that although most of the methane production comes from the expanded bed zone of the reactor, the clarification zone has a small contribution to the gas production within the anaerobic AFEB reactor. There could be two possible reasons for methane gas being produced in the clarification zone, the first being that methanogenic bacteria are sloughed off from the media surface and suspended, and continue their activity in the liquid above the expanded bed. Another plausible explanation is that these organisms are introduced into the reactor with the waste. Since ruminant animals have mesophilic methanogens in their stomachs and intestines, these bacteria do appear in their wastes. Thus when feeding the waste to the reactor, a regular infusion of methanogens was being supplied.

It should be noted here that the serum bottle data in Table 44 seem to suggest certain mechanisms of biomass entrainment in the expanded bed. Serving as evidence of microbial film formation on the media particles was the fact that, under serum bottle test conditions, the washed media sample was able to generate methane gas. If, however, a rigid film attachment to media particles was the sole means of microbial support in the expanded bed, and if the effects of washing itself were minimal on microbial activity, the washed media sample would have produced almost as much gas in the serum bottle test as the media and liquid sample. Instead, upon washing, the media sample lost merely half of its gas production activity, possibly indicating that the biomass may be held in the expanded bed by loose floc attachment to the media particles as well as by fixed film formation.

Reactor Performance

The anaerobic AFEB system was operated under five conditions described in Table 43. The hydraulic retention time is the total volume of liquid retained in the reactor divided by the liquid flow through the reactor, the HRTs for the reactor were calculated on the basis of the total liquid

volume (3.8 ℓ). Solids loading was also calculated on the basis of liquid volume. The void volume of the AFEB systems, occupied by the media, was calculated at 3.2 ℓ . This approach tends to cause the data to be conservative since the clarifier zone would occupy a small fraction of the total volume in a full scale system.

A summary of data gathered at steady state for all five conditions is presented in Table 45. The data from which this table was derived appear in Tables 1 through 5, in Appendix E, for Conditions 1 through 5, respectively. In addition, Figure 73 presents a summary of daily gas production rates. In general, the rate of total volatile solids (TVS) destruction, the difference between influent and effluent g TVS/day, increases with increased TVS loading to the reactor. Total volatile solids removal rates ranged from a low of 5.2 g/day for the lowest loading of 0.52 lbs/ft³/day (Condition 1) to a peak utilization rate of 33.0 g/day for the highest loading of 8.32 lbs/ft³/day (Condition 5). For comparison purposes, it must be noted here that typical loading rates for conventional high-rate fermentation range between 0.10 and 0.50 lbs/ft³/day. It is remarkable that under TVS input rates amounting to 20 times the highest rate recommended for conventional digestion, the anaerobic AFEB fermentor did not appear to reach a limit in its rate of volatile solids utilization with increased TVS loading, nor did total volatile acid concentrations increase significantly. It should be emphasized that the small scale and short duration of these tests would also indicate that the full potential of this concept has not been measured. The duration of testing is illustrated in Figure 74.

The percent decrease in volatile solids for dairy cow manure material passed through the reactor ranged from a high of 39.6 percent for Condition 2 (2 percent TS feed, 1.2-day HRT) to a low of 7.2 percent for Condition 5 (2 percent TS feed, 0.15-day HRT) which received the highest loading. A trend of increasing total volatile solids destruction efficiency with increasing hydraulic retention time (decreasing loading) is shown in Figure 75. The better solids destruction efficiencies for the anaerobic AFEB reactor were generally observed at the lower loading rates. An exception to the trend was Condition 1, operated at 1 percent TS feed, 1.2-day HRT, and 0.52

TABLE 45. SUMMARY OF DATA FROM STEADY STATE CONDITION FOR THE ANAEROBIC AFEB REACTOR.

Parameter	Units	Condi- tion 1	Condi- tion 2	Condi- tion 3	Condi- tion 4	Condi- tion 5
T.S. Loading lbs T.S./ft ³ -day	gm/day	0.52	1.04	2.08	4.16	8.32
Influent T.S.	gm/day	35.8	70.4	141.6	281.8	560.0
Effluent T.S.	gm/day	30.1	46.1	122.5	249.4	520.8
T.S. Reduction	%	15.9	34.5	13.5	11.5	7.0
Influent T.V.S.	gm/day	30.2	58.4	118.9	234.5	460.6
Effluent T.V.S.	gm/day	25.0	35.3	101.1	207.8	427.6
Rate of T.V.S. Destruction	gm/day	5.2	23.1	17.8	26.7	33.0
T.V.S. Reduction	%	17.2	39.6	15.0	11.4	7.2
Influent Total COD	gm/day	41.2	79.4	125.4	251.3	602.7
Effluent Total COD	gm/day	43.5	51.7	145.2	303.3	504.1
Total COD Reduction	%	--	34.9	--	--	16.4
Gas Production*	ℓ/day	2.7	10.86	13.68	19.07	24.59
Gas** CH ₄	%	66.5	66.5	66.5	66.5	66.5
Composition CO ₂	%	33.5	33.5	33.5	33.5	33.5
Gas Production* Predicted From T.V.S. Destruction	ℓ/day	3.91	17.37	13.38	20.08	24.81
Gas Production* Predicted From Total COD Reduction	ℓ/day	--	14.58	--	--	51.89
Effluent T.V.A.+	mg/ℓ	532.0	506.0	719.0	402.0	282.0

* Gas production corrected for 0°C and 1 ATM

** Gas composition computed from serum bottle analysis

+ T.V.A. mg/ℓ as acetic acid

Figure 73. Rates of solids destruction and gas production resulting from varying organic loading rates for the anaerobic attached film fermentor operated at 35°C with 1 and 2 percent TS at HRT's varying from 0.15 to 1.2 days.

Figure 74. Chronological development of gas production in the laboratory scale anaerobic attached film expanded bed reactor operated at 35°C.

Figure 75. Effect of the hydraulic retention time on the total volatile solids destruction efficiencies achieved with a bench scale anaerobic attached film expanded bed fermentor at 2.0 percent TS manure feed.

lbs/ft³/day TVS loading, which resulted in half the solids destruction efficiency produced by Condition 2 operated at twice the loading, 2 percent TS feed, 1.2-day HRT, 1.04 lbs/ft³/day. The sharp drop in activity observed for Condition 1, however, is best illustrated in comparing rates of TVS destruction and gas production for the entire range of organic loadings applied as shown in Figure 75.

The gas production predicted from total volatile solids destruction was generally in good agreement with actual gas production observed. The gas composition was determined from the serum bottle analysis. It was not possible to obtain a gas sample from the anaerobic AFEB reactor which was not contaminated with air. Thus the serum bottle analysis was used.

The gas production predicted from total COD reduction was in poor agreement with gas production observed in the serum bottle. In three conditions the effluent total COD concentration was actually found to be greater than the influent total COD. These data are felt to be unrepresentative of what was actually occurring. Total COD analysis is highly subjective to obtaining a representative sample because of the high dilution required in running the analysis on dairy cow manure. Since manure is highly concentrated in particulate solids, a representative sample is difficult to obtain. Therefore, the COD data in this study are felt to be unreliable.

The effect of solids loading on gas production is presented in Figure 73 for all five conditions. These data indicate that the maximum gas production rate was not observed in this process, and that it could be significantly greater than 30 liters per day for the 3.8 liter reactor.

Control Reactor Performance

The control completely mixed semi-continuous reactor was operated throughout the study in parallel with the anaerobic AFEB unit. The control unit was operated at two different conditions, namely 1.0 percent total solids feed at a 10-day HRT, and 2.0 percent total solids feed at a 10-day HRT. The data collected at the first conditions was used to compare with the data collected from the anaerobic AFEB reactor during Condition 1, and likewise

the second condition was used as a basis for comparison with the data collected for Conditions 2 through 5, at which the anaerobic AFEB was operated.

Figure 76 presents the daily record of gas production and effluent pH for both conditions at which the control reactor was operated. As can be seen, after initial start-up the gas production for both conditions remained relatively constant (± 0.18 liters for Condition 1 and ± 0.29 liters for Condition 2). This would indicate that the quality of the manure remained relatively constant and any increase in gas production in the anaerobic AFEB reactor was not a result of changes in the manure characteristics.

A compilation of all data collected at steady state for both conditions at which the control reactor was operated is presented in Tables 6 and 7 of Appendix E. These data were summarized and appear in Table 46. The total volatile solids destruction for both conditions was approximately 32 to 35 percent, which is somewhat higher than what was found in most conditions at which the anaerobic AFEB was operated. The gas production predicted from total volatile solids destruction is equivalent to gas production which was observed for Condition 2 (2 percent TS feed, 1.2-day HRT; however, in Condition 1 (1 percent TS feed, 1.2-day HRT) it is double what was actually observed. The solids destruction figures for this condition are probably in error. If the observed gas production is used to calculate total volatile solids destruction, a value of 17.8 percent reduction would be obtained. This then corresponds closely to the 17.2 percent total volatile solids destruction achieved by the anaerobic AFEB reactor with the same substrate. The calculated total volatile solids destruction is probably more accurate than what was actually measured. This is further substantiated by the fact that in Condition 2 for both systems the total volatile solids destructions were very close (32.0 percent and 39.6 percent). The only difference between Conditions 1 and 2 was the percent total solids fed the units. As further evidence of the measured total volatile solids reduction being in error for Condition 1 at which the control unit was operated, gas production predicted from total COD reduction for both conditions are close to observed gas production values.

Figure 76. Daily gas production and effluent pH from the operation of a completely mixed semi-continuous fermentor at 1.0 and 2.0 percent T.S. manure feed and 10 day HRT.

TABLE 46. SUMMARY OF DATA FROM STEADY STATE CONDITIONS FOR SEMI-CONTINUOUS COMPLETELY MIXED CONTROL REACTOR.

Parameter	Units	Condition 1	Condition 2
Loading lbs T.S./ft ³ -day	gm/day	0.062	0.124
Influent T.S.	gm/day	3.50	7.00
Effluent T.S.	gm/day	2.49	5.18
T.S. Reduction	%	28.9	26.0
Influent T.V.S.	gm/day	2.96	5.81
Effluent T.V.S.	gm/day	1.90	3.95
T.V.S. Reduction	%	35.8	32.0
Influent Total COD	gm/day	4.21	7.90
Effluent Total COD	gm/day	3.57	5.04
Total COD Reduction	%	15.2	36.2
Gas Production*	ℓ/day	0.44	1.45
Gas Composition** CH ₄	%	60	60.0
CO ₂	%	40.0	40.0
Gas Production Predicted From T.V.S. Destruction*	ℓ/day	0.88	1.55
Gas Production Predicted From Total COD Destruction*	ℓ/day	0.37	1.67
Effluent T.V.A.+	mg/ℓ	426.0	375.0

* Corrected to 0°C and 1 ATM

** Gas composition assumed

+ mg/ℓ as acetic acid

Total volatile acid concentrations in the control reactor averaged about 400 mg/l as acetic acid. This value is within the range which was observed for the anaerobic AFEB reactor.

SUMMARY AND CONCLUSIONS

The anaerobic AFEB reactor was successfully operated at five different conditions with HRT as low as 0.15 days. Loading rates, greater by an order of magnitude over the loading rates used by conventional anaerobic digesters, were employed without apparent problems of volatile acid or ammonia inhibitions. The conclusions of this study are as follows:

1. the anaerobic AFEB reactor was able to ferment a high strength particulate organic residue, namely cow manure, at rates higher than have been achieved previously;
2. total volatile solids destruction efficiencies of 39.5 percent were achieved at an HRT of 1.2 days, using a 2 percent TS feedstock, whereas equal efficiencies occurred at an HRT of 10 days with the same substrate in a conventional anaerobic fermentor produces;
3. this process was found to be stable and capable of producing gas at HRTs as low as 0.15 days, although total volatile solids destruction efficiencies decreased to 7.2 percent;
4. at the highest loading rates of 8.2 lbs TS/ft³/day (HRT of 0.15 days) a maximum gas production of 6.60 volumes of gas per volume of reactor per day was observed;
5. serum bottle analysis showed that fermentation of this residue was occurring in the expanded bed zone of the reactor and that viable microbes were attached to the media;
6. serum bottle analysis could not quantitatively determine the relative amounts of activity between the media zone of the reactor and the clarification zone, but it did show qualitatively that both portions of the reactor were responsible for methane production.

Principal Authors

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CHAPTER VI

KINETIC ANALYSIS

Anaerobic fermentation technologies have proven useful in the reduction of municipal sewage sludges and some industrial wastes. As this study has shown, agricultural residues can also be effectively treated by a variety of innovative anaerobic fermentation processes to yield a more stable residue and a useable energy product, methane. In fact fermentation of agricultural residues results in a more stable and less sensitive system than has been reported with sewage sludge. In the past, a better understanding of biological systems has been made possible through kinetic analysis of the process. A knowledge of process kinetics can be used as a basis for the successful design and operation of a biological system and for predicting the overall efficiency and stability of the process.

THEORETICAL CONSIDERATIONS

There are two general approaches used to mathematically model biological processes - a rational fundamental approach based on the understanding of microbial growth and an empirical approach which is dependent upon laboratory or full scale analyses of the substrate. The rational model is most often used when it is desirable to obtain a clearer understanding of microbial processes, but because of its simplicity the empirical model is often used in actual design situations. Many attempts have been made to develop alternative analyses, but they usually are much more complex and because of this they either fail to actually symbolize the actual microbial growth processes or they are difficult to apply. No attempt will be made to present a review of these approaches here. The rational approach is based on pure bacterial culture enzyme kinetics and assumes that the microbial growth rate (and therefore the rate of substrate removal) is a function of substrate concentration in contact with the microorganisms, or the mixed

liquor substrate concentration in completely mixed systems, and the concentration of microorganisms in the reactor. This rationale is embodied in the Monod model (Monod, 1949) which is represented by:

$$\mu = \frac{\bar{\mu} S}{K_s + S} \quad \text{Eq. 30}$$

where μ is the specific growth rate of the microorganisms (time^{-1}), $\bar{\mu}$ is the maximum specific growth rate, S is the substrate concentration (mass/volume) and K_s is the half velocity coefficient (mass/volume). For a continuous, single pass, completed mixed reactor the specific growth rate is conventionally considered to be equal to the inverse of the hydraulic retention time ($\mu = \text{HRT}^{-1}$). From a mass balance standpoint this is true if there are no significant inputs of biomass to the system other than microorganism growth in the reactor. In light of the fact that dairy cow manure is a residual product of the anaerobic fermentation of materials within the rumen, this model is difficult to apply.

The other approach considers the reactor biomass as a catalyst and assumes an overall first order mass transfer kinetic model to describe a "catalyzed" reaction. This empirical model assumes the rate of substrate removal to be proportional only to the substrate concentration (S) surrounding the catalyst surface (Prakasam, *et al.*, 1974; Goodman and Englande, 1974), the microorganisms. The functional form of this model, presented in Chapter II, will be restated here:

$$\frac{dS}{dt} = -kS \quad \text{Eq. 31}$$

where, t , is time and, k , is the substrate removal rate coefficient. Despite its simplicity, Morris (1976), in a comparison of the two methods, found the empirical model to be quite applicable in the prediction of reactor performance with respect to the anaerobic fermentation of dairy cow manure at mesophilic temperatures.

This general rate expression may be used to determine the performance of a variety of theoretical fermentors. A batch reactor is the simplest of reactor concepts. The batch reactor is a stirred or unstirred reactor which has been initially filled with feed material, seeded with an ample culture of microorganisms and allowed to operate without feeding for a long period of time. If the well-mixed reactor is fed periodically but allowed to ferment as a batch reactor between feedings, it is considered to operate in a semi-continuous or semi-batch mode. A continuous stirred tank reactor (SRT) is a continuously-fed well-mixed reactor in which the influent stream, upon entering the fermentor, instantaneously takes on the physical and concentration characteristics of the reactor contents and in which the effluent stream leaving the system is the same in composition as the reactor contents.

The plug flow reactor definition has already been stated in depth in Chapter 3, Simplified Fermentor. This reactor type is usually considered to operate in the continuous mode. An ideal plug flow condition is one in which the particles leave the reactor in the same sequence they entered. In this reactor, each infinitesimal volumetric element has the same residence time in the reactor. Because each of these fluid elements move through the reactor independently, they are considered as mobile batch reactors, explaining why ideal plug flow and batch reactor kinetics are exactly the same. In reality some mixing through diffusion and dispersion always occurs in the plug flow design, so that it departs from the batch kinetic definition.

The first order kinetic model development for the continuous stirred tank reactors (CSTRs) arranged in series has also been discussed in Chapter 3. It has been mathematically shown that the ideal plug flow reactor is kinetically equivalent to an infinite series of CSTRs. This enables plug flow reactors to be evaluated on a small scale with a series of connected completely mixed reactors.

The expressions describing the substrate removal performance for various theoretical reactor types are presented in Table 47.

TABLE 47 . REACTOR PERFORMANCE BASED ON THE EMPIRICAL MODEL .

Fermentor Type	Effluent Biodegradable Volatile Solids Concentration
Batch	$S_{bl} = S_{bo} \exp(-k \text{ HRT})$
CSTR	$S_{bl} = \frac{S_{bo}}{1 + k(\text{HRT})}$
Series of CSTRs	$S_{bl} = \frac{S_{bo}}{(1 + \frac{k \text{ HRT}}{n})^n}$
Semi-continuous	$S_{bl} = \frac{S_{bo} (t) \exp(-kt)}{\text{HRT}(1 - (1 - \frac{t}{\text{HRT}}) \exp(-kt))}$
Plug Flow	$S_{bl} = S_{bo} \exp(-k \text{ HRT})$

S_{bl} = effluent biodegradable volatile solids concentration, g/l

S_{bo} = influent biodegradable volatile solids concentration, g/l

HRT = hydraulic retention time

k = first order reaction rate coefficient

n = number of CSTRs in series

t = incremental time period in semi-continuous mode

The first order model and the equations in Table 47 will serve as a basis for kinetic evaluation of the reactors operated in this study. This model was chosen because it has been demonstrated to be applicable to the anaerobic fermentation of dairy cow manure (Morris, 1976) and because it does not incorporate terms such as viable cell mass, which are difficult to verify with direct measurement.

The mesophilic and thermophilic control reactors were usually operated on a semi-continuous basis (fed daily) and mixed by manual agitation twice a day. However, it can be seen from the semi-continuous reactor performance equation that at the longer hydraulic retention times, $t \ll \text{HRT}$, the limit of that expression yields a new expression:

$$S_{bl} \text{ (Semi-continuous)} = \frac{S_{bo}(t) \exp(-kt)}{\text{HRT}(1 - (1 - t/\text{HRT}) \exp(-kt))} \quad \text{Eq. 32}$$

and

$$S_{bl} \text{ (CSTR)} = \lim_{t \rightarrow 0} \left[\frac{S_{bo} t \exp(-kt)}{\text{HRT}(1 - (1 - t/\text{HRT}) \exp(-kt))} \right] = \frac{S_{bo}}{k \text{ HRT} + 1} \quad \text{Eq. 33}$$

which is the performance equation for a CSTR. This illustrates that a semi-continuous control fermentor could be considered equivalent to a CSTR when operated at relatively long HRTs.

The liquid-solids separation reactor was operated in a semi-continuous mode at SRTs ranging from 4.62 to 22.5 days with and without recycle. If it is assumed that most of the biological activity takes place in the liquid portion of the reactor, the system can then be described with the semi-continuous kinetic equation while consider S_{bo} and S_{bl} in terms of the total amount of biodegradable solids entering and leaving this unit each day. Because this unit was fed on a daily basis, the time interval, t , of semi-continuous operation should be considered to be 1.0 day.

The anaerobic attached film expanded bed (AFEB) reactor was also operated in the semi-continuous mode. On the basis of experimental data

presented in Chapter 5, the reactive volume of this unit was considered to be the volume of the expanded media bed itself, 3.8 liters. The time intervals of semi-continuous operation of this unit ranged from 0.15 to 1.0 days.

The pilot unit plug flow reactor is difficult to characterize in terms of a predictive performance model. Designed to simulate the plug flow condition, the reactor exhibited the flow patterns of a dispersed plug flow reactor, as interpreted from solids data and tracer study information presented in Chapter 3. The performance for any non-ideal plug flow reactor would fall somewhere between the solids destruction efficiencies of a CSTR and an ideal plug flow reactor (Grieves, et al., 1964).

KINETIC COMPARISONS OF REACTOR PERFORMANCE

The first order reaction rate coefficient, k , for any particular fermentor is, by definition, an indication of the rate of substrate utilization per unit of substrate present. For each of the theoretical fermentor types included in Table 48, the reaction rate coefficient can be directly calculated from a simple rearrangement of the corresponding kinetic expression using steady state data on influent and effluent BVS concentrations. The resulting equations allowing the computation of k for each kind of fermentor are tabulated in Table 48.

The first order reaction rate coefficient can also be calculated graphically if enough conditions are conducted with a particular type of reactor. This is the conventional method used to calculate k for continuous or semi-continuous fermentors. In Chapter 2, for instance, the reaction rate coefficient for the thermophilic CSTRs was calculated by measuring the slope of a plot of $(S_{bo} - S_{bl})/HRT$ versus S_{bl} , as shown in Figure 5.

The first order reaction rate coefficients for the various fermentor types investigated are summarized in Table 49. Sufficient data was not developed to use the above graphical technique to solve for k in all types of reactors. All values in Table 49 were obtained using one representative steady state data point and the appropriate reactor equation given in

TABLE 48. DIRECT COMPUTATION OF K USING
EMPIRICAL MODEL KINETICS.

Fermentor Type	Reaction Rate Coefficient
Batch	$K = \text{HRT}^{-1} \ln \frac{S_{bo}}{S_{bl}}$
Ideal Plug Flow	
Single Pass	$K = \text{HRT}^{-1} \ln \frac{S_{bo}}{S_{bl}}$
Recycle	$K = \text{HRT}^{-1} \ln \left[\frac{S_{bo} (1 - \alpha) + \alpha S_{bl}}{S_{bl}} \right]$
CSTR	
Single Pass	$K = \frac{S_{bo} - S_{bl}}{S_{bl} \text{HRT}}$
Recycle	$K = \frac{S_{bo} (1 - \alpha) + S_{bl} (\alpha - 1)}{S_{bl} \text{HRT}}$
Series of CSTRs	
Single Pass	$K = \frac{n}{\text{HRT}} \left[\left(\frac{S_{bo}}{S_{bl}} \right)^{1/n} - 1 \right]$
Recycle	$K = \frac{n}{\text{HRT}} \left[\left[\frac{S_{bo} (1 - \alpha) + S_{bl} \alpha}{S_{bl}} \right]^{1/n} - 1 \right]$
Semicontinuous	
Single Pass	$K = 1/t \ln \left[\frac{(HRT-t) S_{bl} + S_{bo} t}{S_{bl} \text{HRT}} \right]$
Recycle	$K = 1/t \ln \left[\frac{S_{bl} (HRT-t + \alpha t) + S_{bo} t (1 - \alpha)}{S_{bl} \text{HRT}} \right]$

Where α = fraction of effluent flow recycled.

TABLE 49 REACTION RATE COEFFICIENTS FOR THE REMOVAL OF THE BIODEGRADABLE ORGANICS IN DAIRY MANURE IN VARIOUS FERMENTOR TYPES INVESTIGATED.

Fermentor Type and Operating Mode	Method of Computation	Biodegradable Volatile Solids, First Order Reaction, Rate Coefficient, K, Day ⁻¹
Mesophilic Control, 3.0ℓ, 22°C		
12-day HRT	Equation 37	0.079
30-day HRT	Equation 37	0.049
Mesophilic Control, 30ℓ, 35°C		
10-day HRT	Equation 37	0.16
30-day HRT	Equation 37	0.086
Mesophilic Control, 3.5ℓ, 35°C		
Strained Manure Feed		
1% T.S. Feed, 10-day HRT	Equation 37	0.054
2% T.S. Feed, 10-day HRT	Equation 37	0.046
Thermophilic Control, 15ℓ, 60°C		
5-day HRT	Graphical	0.157
10, 20, 30-day HRT	Graphical	0.081
Thermophilic Liquid-Solid Separation Reactor, 180ℓ, 55°C		
5-day HRT		
1-day Float Residence Time (FRT)	Equation 37	0.23
5-day FRT, Float not remixed	Equation 37	0.16
5-day FRT, Float remixed	Equation 37	0.22
10-day FRT, Float not remixed	Equation 37	0.22
5-day FRT, Recycle ($\alpha = .84$)	Equation 38	0.074
Anaerobic AFEB, 3.8ℓ, 3.5°C, Strained Manure Feed		
1.2-day HRT, 1% T.S. Feed,		
t = 1.0 day	Equation 37	0.79
1.2-day HRT, 2% T.S. Feed,		
t = 1.0 day	Equation 37	1.0
.6-day HRT, 2% T.S. Feed,		
t = 0.5 days	Equation 37	1.1
.3-day HRT, 2% T.S. Feed,		
t = 0.25 days	Equation 37	1.7

TABLE 49. (CONTINUED).

Fermentor Type and Operating Mode	Method of Computation	First Order Reaction Rate Coefficient, k , Day ⁻¹
Anaerobic AFEB, 3.8ℓ, 3.5°C Strained Manure Feed (cont.)		
.15-day HRT, 2% T.S. Feed, $t = 0.125$ days	Equation 37	1.5
Pilot Scale Unmixed Horizontal Displacement Reactor, 5600ℓ, 35°C		
30-day HRT, 10-12% I.S. Feed		
Assuming Ideal Plug Flow Conditions*	Equation 31	0.049
Assuming a High Degree of Dispersion*	Equation 33	0.073
Four CSTRs in Series, 12ℓ, 22°C		
12-day HRT, 8% T.S. Feed	Equation 35	0.013
30-day HRT	Equation 35	0.012
Four CSTRs in Series, 12ℓ, 35°C		
10-day HRT	Equation 35	0.034
10-day HRT, 40% Recycle	Equation 36	0.023
12-day HRT	Equation 35	0.030

*In accordance with tracer study results, an effective HRT of 17 days was assumed in the calculation of K for the pilot-scale UHD reactor.

Table 48 except where noted differently. In examining reaction rate coefficient data it should be remembered that the kinetic constant, k , as presented in Table 49 is specific for a reactor system, population of bacteria, substrate type and concentration and a set of environmental conditions. Theoretically, the reaction rate coefficient k will be affected by anything which can interfere or enhance a biologically-catalyzed reaction.

The highest reaction rate as shown in Table 49 was attained by the anaerobic attached film expanded bed (AFEB) reactor (35°C). This fermentor had a reaction rate coefficient averaging around 1.50 days^{-1} when operated on 2% total solids concentration manure feed compared to 0.046 days^{-1} obtained from a 10-day HRT control operating on the same feedstock. In addition, the anaerobic AFEB k value was roughly an order of magnitude (10X) higher than any of the largest k values encountered with the mesophilic or thermophilic control fermentors operated at the shorter hydraulic retention times.

There are, no doubt, a number of properties unique to the anaerobic AFEB system which may explain its vast kinetic superiority. Several such characteristics are hypothesized:

1. The cell mass is attached or entrapped by a stationary phase in the reactor; this makes it possible to operate at high flow rates through the reactor without danger of washout of the viable microorganisms. It has been hypothesized that a maximum of 25 to 30 gm/l of active microbes can be held in these types of microbial attached film systems (Jewell and MacKenzie, 1972).
2. An intimate contact of substrate with the microorganisms. A constant recycling of the feed material over the biological films attached to the expanded media maintains a higher concentration gradient around the microbial cell.
3. A natural selection for microorganisms capable of attachment or entrainment by the expanded media may result in a population with greater solids destruction efficiency.

Upon closer examination of the data it can be seen that for some of the control fermentors operated over a wide range of HRT's, two reactions rate coefficients are reported. Occurring at both mesophilic and thermophilic temperatures, the dairy cow manure substrate seems to be comprised of at least two classes of components which are degraded at different rates. The more easily degraded substances in the manure, such as starch, urea, simple sugars, small chain proteins, etc., are removed quickly and impart a high value for k as measured at the shorter retention time. At the longer retention times, the readily degraded substrate fraction becomes depleted and the more slowly available materials, such as cellulose and hemicellulose, are utilized at a slower rate, thus decreasing the value for the reaction rate coefficient. This phenomenon was evident from the first order plot $(S_{bo} - S_{bl})/HRT$ versus S_{bl} , of the thermophilic control fermentor data shown in Figure 5 . In this graph the data points seemed to be grouped about two separate linear plots, the slopes of which defined two reaction rate coefficients of significantly different magnitudes. This also seems to be supported by data in Figure 2 of the thermophilic chapter which indicates essentially a constant solids removal efficiency as the organic loading rate was increased from 0.3 to 2.22 lbs VS/ft³/day.

The reaction rate coefficient can also be affected by the temperature of fermentation. This effect can be seen in comparing the kinetic constants of the control fermentors and CSTRs in series operated at 35°C and 22°C. The data indicate a doubling of the reaction rate coefficient for these systems when the temperature is increased from 22°C to 35°C, as would be expected. A further increase in the rate constant, however, was not realized when the temperature was increased from 35°C to 60°C. The reaction rate coefficients for the thermophilic control fermentors are about equal to the k values of the mesophilic control systems. It was shown in Chapter II that although the reaction rate coefficients for the mesophilic and thermophilic fermentation of dairy cow manure are comparable, the thermophilic systems showed a substantial performance advantage in being able to withstand higher hydraulic and solids loadings without distress or washout.

The kinetic information from the thermophilic control fermentors correlated well with the reaction coefficient values obtained from the various operating modes of the thermophilic liquid-solid separation reactor. The first order reaction rate constant calculated for the removal of the highly biodegradable fraction of the manure substrate at a 5-day HRT was 0.16 days^{-1} for the thermophilic control fermentors (15 ℓ) and averaged approximately 0.21 days^{-1} for the thermophilic liquid-solid separation unit (180 ℓ). This correlation between the k values of these two semi-continuous reactors having a greater than ten-fold difference in fermentor volume, is an indication of the applicability of the bench-scale kinetic information to the design and scale-up of thermophilic systems.

IMPLICATIONS FOR PRACTICAL DESIGN

The kinetic development presented here points out the difficulty of deriving a unified empirical design expression for the fermentation of a multi-component substrate, holding valid over a wide range of operational conditions. The development of such an all-inclusive model would require more fundamental research into the nature of the degradable fractions comprising the dairy cow manure substrate, and their respective patterns of removal by anaerobic fermentation. As is true with all complex substrates composed of liquids and solids, the inability to measure the active biomass in the reactor influent and effluent samples make it difficult to compare results. In part, inclusion of the biodegradable mass in the first order decay kinetic analyses is one simplified method of minimizing this problem.

On the basis of the kinetic information presented here, a careful and simplistic approach must be taken in using the empirical kinetic model to predict the solids destruction efficiency and, hence, the gas production capability of any particular reactor design. This would involve following certain hydraulic retention time constraints in applying the correct reaction rate coefficient values to the pertinent predictive equations. The performance equations for each of the major types of fermentors studied, listed in Table 50, are only valid within certain specified ranges of HRT values. It should also be mentioned that since the substrate terms in these

TABLE 50 . PERFORMANCE PREDICTION EQUATIONS USED WITH
THE EMPIRICAL MODEL .

Reactor Type	HRT Range Days	Applicable Empirical Performance Equation
Conventional Digester at 22°C	12-30	$S_{bl} = \frac{S_{bo}}{1 + .064 \text{ HRT}}$
Conventional Digester at 35°C	< 10	$S_{bl} = \frac{S_{bo}}{1 + .16 \text{ HRT}}$
	20-30	$S_{bl} = \frac{S_{bo}}{1 + .086 \text{ HRT}}$
Conventional Digester at 60°C	≤ 5	$S_{bl} = \frac{S_{bo}}{1 + .16 \text{ HRT}}$
	10-30	$S_{bl} = \frac{S_{bo}}{1 + .081 \text{ HRT}}$
Thermophilic Liquid-Solid Separation Reactor, 55°C, fed once daily (t = 1.0)	4.5-7.5	$S_{bl} = \frac{S_{bo}(t) \exp(-.21t)}{\text{HRT} (1 - (1-t/\text{HRT}) \exp(-1.21 t))}$
Anaerobic AFEB	.15-1.2	$S_{bl} = \frac{S_{bo}(t) \exp(-1.50t)}{\text{HRT} (1 - (1-t/\text{HRT}) \exp(-.073t))}$
Pilot Scale UHD Reactor, 35°C, fed once dairy	> 17 (Effective HRT)	$S_{b \text{ CSTR}} \geq S_{bl} \geq S_{b \text{ PF}}$ Assuming Dispersion $S_{b \text{ CSTR}} = \frac{S_{bo}(t) \exp(.073 t)}{\text{HRT}(1 - (1-t/\text{HRT}) \exp(-.073t))}$ or $S_{b \text{ CSTR}} = \frac{S_{bo}}{.076 \text{ HRT} + 1}$ Assuming Plug Flow $S_{b \text{ PF}} = S_{bo} \exp(-.073 \text{ HRT})$

equations are expressed as biodegradable volatile solids, the refractory fraction of the dairy cow manure influent must be known.

CONCLUSIONS

The kinetic analysis of the data generated from the operation of various types of conventional and innovative reactors has allowed a clearer understanding of the potential of the anaerobic fermentation process. In light of the data presented in this study, the following conclusions seem appropriate:

1. Dairy cow manure, being composed of substances of varied biodegradabilities, may result in rapid removal of specific components followed by a slower rate of removal of materials more resistant to bacterial attack.
2. The most rapid utilization of substrate during this investigation was observed with the anaerobic attached film expanded bed (AFEB) reactor which was able to effect a reaction rate coefficient of 1.50 days^{-1} , 10 to 30 times greater than the k-values obtained from the control fermentors.
3. Increased temperature brought about an increase in the reaction rate coefficient in the range of 22°C to 35°C . Fermentation at thermophilic temperatures ($55\text{--}60^{\circ}\text{C}$), however, resulted in reaction rate coefficients no higher than those obtained at 35°C .
4. Controlled solids flotation within a fermentor does not seem to adversely affect the rate of substrate utilization at thermophilic temperatures. The reaction rate coefficients obtained from the liquid-solids separation reactor were in fact slightly higher than those obtained from the completely mixed control fermentors. The reaction rate coefficient also seems to be unaffected by the various float management techniques described and employed with the liquid-solids separation unit.

5. In the course of using any kinetic model as a basis for design, care must be taken to understand the conditions under which that model may be safely applied. One of the most important constraints to remember in using the empirical model is that a certain reaction rate coefficient value may be valid only within a specific range of hydraulic retention times.

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CHAPTER VII

MICROBIOLOGY OF ANAEROBIC FERMENTATION OF DAIRY COW MANURE

INTRODUCTION

Significant recent advances have been made in the understanding of the bacteriology and biochemistry of anaerobic fermentation. However, the understanding of the complex interactions of the various species and the identification of the practical or actual limiting factors for conversion of organics is still unclear. The basic premise upon which this portion of the study was based was to develop fundamental information regarding the major microbial populations in well managed complex fermentors in order to clarify the major factors which control fermentation. Although this effort will be continued at a low level, the early termination of this study (see Acknowledgment Section) resulted in incomplete data for this portion of the study.

The goal of the microbiological portion of this study was to examine the relationship between the activities and concentrations of specific microbial populations to the overall performance of reactors producing methane from animal residues. A clearer understanding of the controlling biological parameters will assist in developing fermentation strategies that optimize the conversion of organic residues to methane. There are essentially no studies that attempt to relate microbial activities and concentrations to reactor performance. In fact, there is an appalling lack of information on the nature of the important microorganisms involved in the conversion of organic residues to methane, no less their activities and concentrations.

There are several approaches that can be taken to study the microbial population of anaerobic fermentors. Some were beyond the scope of this study. An example is the isolation, identification of species and enumer-

ation of the major polymer-fermenting species in the reactors. This study was restricted to what seemed to be practical approaches. Approaches were emphasized that would provide meaningful feedback on engineering studies rather than emphasizing studies of fundamental microbiological characteristics of reactors. Even with these practical constraints on experimental design, it was impossible to carry out complete studies because of the short time frame of the project and the shift of project orientation at the end of the first year. This portion of the study had been on-going for six months when the project orientation was shifted.

The following objectives were the main focus of this short study:

1. Correlation of short term laboratory measurements of the rate of methanogenesis with actual rates of methane production in reactors.
2. Enumeration of the total viable anaerobic bacterial count of reactor contents.
3. Enumeration of the methane-producing bacteria.

BACKGROUND

Rates of Methanogenesis

A method for measuring the instantaneous rate of methanogenesis in a reactor using a small sample and short incubation times would be a valuable tool. Instantaneous rates could be useful for on-line monitoring of the operation of experimental units, studies of diurnal variations, and studies of activities of portions of non-homogeneous systems such as expanded bed or plug flow units. In addition, by supplementing contents of the system with the major substrates for methanogenesis, i.e., acetate, formate and H_2 and CO_2 , it might be possible to obtain the actual potential for methane production by the methanogenic population independent of the overall rate of the system.

The system can be thought of as a sequence of reactions:

- a) Polymers $\longrightarrow$ Acetate + Formate + H_2 + CO_2 (Fermentation by non-methanogenic bacteria)
- b) Acetate + Formate + H_2 + CO_2 $\longrightarrow$ CH_4 (Activities of methanogenic bacteria)

Supplementation of reactor contents with the specific substrates for (b) could provide an estimate of the potential for methanogenesis. In an unsupplemented system, rates are dependent not only on the rate of (b), but also on the rate of the reaction in (a) and the concentration of the polymer (feed substrate).

Total Viable Count

These studies were designed to obtain a measure of the concentration of organisms involved in the fermentation of feed substrate to the precursors of methane.

Enumeration of Methanogenic Bacteria

These studies were designed to obtain a measure of the concentration of organisms that convert the precursors of methane to methane.

EXPERIMENTAL PROCEDURES

Steady state samples from all reactors were to be shipped from the Cornell laboratories to the anaerobic bacteriology laboratory at the New York State Department of Health Laboratories at the time that steady state operating conditions were achieved for all reactors. Although at times more than 20 reactors were being operated, most samples that were obtained were from the thermophilic units reported in Chapter II. Thus, a large amount of data on high temperature reactors were obtained even though the main focus had been originally planned on mesophilic communities.

Approximately 300 grams of reactor contents were shipped 200 miles by surface transportation in containers that kept the samples at about 4°C. Shipping time from sampling to receipt of the material was about 15-20 hours. Two gram samples were weighed or pipetted for rate studies and 1 gram was used for preparing dilutions for total counts and counts of methanogens within about 24 hours after the sample was taken. Anaerobic procedures were used throughout the laboratory study. All rate studies were conducted as soon as samples had been accessioned and counting studies were also immediately initiated. Counts were determined after 2 weeks of incubation.

Total viable counts were in an agar medium containing glucose, cellobiose, starch, trypticase, yeast extract supplemented with rumen fluid. Colony counts were recorded. Methanogen counts were in a liquid medium, either unsupplemented, supplemented with acetate, or supplemented with formate and H_2 - CO_2 . Triplicates of each dilution were examined for methane at the end of 2 weeks and concentrations were computed by the most probable number method based on the number of tubes in which methane was detected. Additional information on the procedures are given in the Appendix.

RESULTS AND DISCUSSION

Laboratory Rate Determinations

It was feasible to obtain rates of methane production with 2 gram samples of reactor contents after 3 hours of incubation in the laboratory. Rates were obtained by incubating samples in an anaerobic environment with buffer, reducing agent, and without and with separate supplements of acetate and formate. The temperature was the same as that of the reactor. Gas production was measured by gas chromatography.

Supplementation with acetate or formate always stimulated rates of production of methane because the methanogenic species did not have to depend on the conversion of complex substrates to precursors of methane, such as acetate or formate, to produce methane. H_2 was produced by all samples that

produced methane. H_2 was probably the actual substrate used for reducing carbon dioxide to methane. Rates were reproducible for duplicate samples from a single reactor (this was checked with 4 separate samples from thermophilic reactors). Three hour readings represent only a small proportion of the total capacity for methane formation with supplemented samples, and therefore, are valid rate and not endpoint measurements.

Experiments with thermophilic reactors (5 and 10-days HRT) showed that short term laboratory gas production rates were proportional to loading rates which in turn were proportional to rates obtained with the pilot reactors. A typical example is shown in Table 51. Similar trends were obtained with 5-day HRT thermophilic reactors but two unexplained differences appeared. More scatter in pilot reactor rates than laboratory rates was noted for the 10-day HRT samples and the reverse was true for the 5-day HRT samples. Even more puzzling was the fact that the mean of the ratio of laboratory/reactor rate for the 10-day HRT and 5-day HRT samples were 2.04 and 0.43, respectively. Formate-supplemented laboratory rates were used to calculate the preceding ratios but the trends were the same for acetate supplemented and unsupplemented rates. There is no obvious explanation of the discrepancies between the 5 and 10-day HRT studies.

TABLE 51 RATES OF CH_4 PRODUCTION IN 3 HOUR LABORATORY EXPERIMENTS AND IN REACTORS (THERMOPHILIC - 10 DAY HRT).

% Solids in feed	$\ell CH_4 / \ell$ contents/day ¹		
	Laboratory(A) ² (after shipping & treatment)	Cornell Reactor(B) (before shipment)	A/B
4	0.86	0.44	1.95
6	1.24	0.69	1.85
8	1.43	0.73	1.95
10	2.46	1.02	2.41

¹Values are means of 3 separate samples in steady state.

²Formate-supplemented.

An interesting trend was noted in the ratio of rates obtained with acetate- and formate-supplemented samples from the 5-day HRT thermophilic reactors. This is illustrated in Table 52 for the reactor fed 10% solids. The acetate-supplemented rate increased more than the formate-supplemented with length of run. This was also noticed in the reactors fed 4, 5 and 8% solids. This suggests that when the acetate/formate ratio is high, gas and methane production is high in the reactors. The reactor fed 8% solids gave an acetate/formate rate of 0.56 for the 1/18/77 samples (3.11 l/l/day), 2.89 for the 2/14/77 sample (3.12 l/l/day). Although daily production of gas appears to be the same, other data show a decrease and a subsequent increase in the 8% solids 5-day HRT reactor. When 3/30/77 and 3/31/77 samples of this reactor were examined after it had been shifted to a 2.5-day HRT and was producing about 4 l/l/day, the acetate/formate ratios were 1.75 (3/30/77) and 1.74 (3/31/77). Additional information on the relationships of these two reactors could lead to a more sound definition of anaerobic fermentor steady state conditions.

There is a suggestion in these data that there are variations in the capacity of the population to convert acetate to methane and carbon dioxide relative to the capacity to produce methane from H_2 plus carbon dioxide or from formate. There is also the suggestion that this increase in relative capacity for acetate utilization is related to increases in total gas production. Unfortunately hard conclusions cannot be made at the present time because of the inability to correlate short term rates with reactor rates in the thermophilic 10 day versus 5-day HRT reactors. There is, however, sufficient reason to believe that a careful study of sampling methods, reaction mixture ingredients and concentrations and a larger number of sampling points may lead to a procedure for estimating methanogenic capacity and detecting gross changes in capacity for producing methane from acetate versus H_2 plus carbon dioxide or formate. These data in turn may provide insight into some of the factors that influence the yields of gas from the reactors.

TABLE 52 CHANGE IN RATIO OF LABORATORY RATES WITH ACETATE AND FORMATE
WITH SAMPLES FROM 10% SOLIDS FED, 5 DAY HRT, THERMOPHILIC REACTOR.

Sample Date	l gas/l/day (reactor)	$\frac{\text{Acetate Rate}}{\text{Formate Rate}}$ (Lab)
1/18/77	3.93	0.50
1/20/77	3.68	0.49
2/3/77	3.30	0.57
2/8/77	not known	0.92
2/14/77	4.03	1.46
2/22/77	5.98	not done

Counts of Organisms

The most interesting bacterial population data were the total viable counts. These correlated reasonably well with reactor performance for the thermophilic 5 and 10-day HRT thermophilic reactors. As shown in Table 53 as methane production per day increased with increasing load, total count increased. There is a disproportionate increase in count with increasing rates of production of methane. Again there is a discrepancy between 10-day HRT and 5-day HRT samples. Only 1 set of 10-day HRT samples was examined, however, whereas the data in Table 53 for 5-day HRT samples represent means of 3 sets of samples.

No meaningful correlations between viable counts of methanogenic bacteria by the MPN method and individual reactor activity or short term rate studies were observed. Part of the lack of correlation stems from the fact that the MPN method's precision is only good enough to detect differences in population of 1 order of magnitude. This type of variation may be meaningless if there is a linear relationship between concentration of viable methanogens and individual rates. When the averages of 3 sets of 5-day HRT samples were compared to the average rates of methanogenesis, however, an inverse correlation between counts obtained on the $\text{H}_2\text{-CO}_2$ plus

TABLE 53. RELATIONSHIP BETWEEN TOTAL VIABLE COUNT, LOADING AND METHANE PRODUCTION IN THERMOPHILIC REACTORS.

Reactor	Reactor 1 CH ₄ /l/day(A)	10 ⁸ Bacteria/g(B)	B/A
<u>10 day HRT¹</u>			
4% solids	0.40	0.2	0.5
6% "	0.50	0.4	0.8
8% "	0.60	0.6	1.0
10% "	0.85	1.2	1.4
<u>5 day HRT²</u>			
4% solids	0.63	1.7	2.7
6% "	1.10	3.7	3.4
8% "	1.51	6.4	4.2
10% "	1.89	8.7	4.6

¹Data from 1 sample set.

²Values are means of data from 3 sample sets.

formate medium and methanogenesis was obtained (Figure 77). The figure also shows that total viable counts are proportional to rates of methane production. It is not yet possible to explain the negative relationship between viable methanogens and methanogenesis. This contrasts with the positive relationship between short term rate studies and reactor performance. One plausible explanation is based on the observation that short term rate studies showed a gradual increase in the utilization of acetate with time of operation of these reactors. It is conceivable that the methanogenic organisms that developed during the shift were not detected by the MPN method used. In order to explain the data, one would have to assume that this new population represented increasing proportions of the total methanogenic population as loading rate was increased. It should also be emphasized that the analyses were arbitrarily done after 2 weeks which would skew the results against more slower growing bacteria.

Figure 77. Microbial analyses of thermophilic (60°C - 5 day HRT) cow manure fermentors in relation to the gas production rate. Methanogens determined by H_2/CO_2 /formate media (see Appendix).

Feedstock - Dairy Cow Manure

A sample of feedstock obtained on 3/14/77 gave low rates of methane production when tested in the 3 hour rate study at 37°C. The methane production rate (ml CH₄/gm/day) were 0.03, 0.02 and 0.03 for unsupplemented acetate- and formate-supplemented reaction mixtures, respectively. Methanogen MPN's at 37°C showed 9.3×10^4 methanogen/g in the H₂/CO₂/formate medium. No rates and less than 30 methanogens were detected at thermophilic temperatures. Feedstock obtained on 3/31/77 was tested at 37°C and showed rates of 0.03, 0.16 and 0.34 ml CH₄/g/day for unsupplemented, acetate- and formate-supplemented reaction mixtures, respectively. Total viable count was 8.0×10^7 bacteria per gram and methanogen MPN was 1.1×10^7 /g in H₂-CO₂/formate medium.

DISCUSSION

The problem of relating the microbiological characteristics to the performance of reactors was only briefly examined. Short term rate studies and total viable counts may provide valuable indices of reactor performance, but more detailed studies are necessary to clarify relationships. Concentration of study on a single reactor with more sampling, tests of adequacy of sampling and reproducibility of results and diurnal variations would be more valuable at this stage than examination of a variety of reactors with many treatment variables. More refined microbiological analyses should be carried out to define the species and to follow population changes at the species level.

Most of the report deals with thermophilic reactors because insufficient data were available to draw conclusions at this point about other systems. Data for all samples thus far analyzed, however, are included in the Appendix.

A picture does emerge from the limited data available that total viable counts are in the range of 10^8 to 10^9 bacteria per gram and methanogens are about 10^7 per gram for high rates of methanogenesis. These

numbers are empirical in the sense that they relate to the specific methods used for enumeration. There are variations in these numbers that have already been discussed. Dairy cow manure contains mesophilic methanogens in reasonably high numbers. It is likely that thermophilic methanogens are present below the level of detection by our procedures. Both the mesophilic and particularly thermophilic methanogens increase considerably in the reactors.

RECOMMENDATIONS FOR FUTURE WORK

1. Short term rate studies of methanogenesis should be continued. The potential exists for increasing the sensitivity of the method by use of flame ionization detection for methane instead of thermal conductivity detection. Supplementation should include feed substrate and H_2 - CO_2 . Possible explanations of the discrepancies between short term rates and reactor rates seen in the present study could not be tested experimentally because of time limitations, but it should be possible to discover the reasons with additional studies. Examination of short term rates can be a very useful tool for examining the capacity of the methanogenic population of the reactor to produce methane.
2. The total viable bacterial count correlated reasonably well with the performance of the thermophilic fermentors. Additional studies should be carried out to extend this correlation, especially to mesophilic units. Because sophisticated microbiology and lengthy incubation times are required, these counts are not particularly useful for process control. They should, however, be continued and extended to identify the most important organisms involved in the degradation of natural organic polymers. Until this is done we will not have a fundamental understanding of the fermentation process.
3. Counts of methanogens are also not very useful for process control. The thrust of research should be in the direction of isolation and characterization of the major methanogens involved

in fermentation of agricultural residues. This, together with the characterization of organisms producing methane precursors from feed substrates (item 2) may lead to improved strategies for fermentation based on optimization in terms of the important microbial inhabitants of the fermentation system.

4. Possibly more important than specific recommendations is the recommendation for continued collaboration between engineers and microbiologists. The microbiologist's goals should be to provide as much help as possible in the application of state-of-the-art techniques by the engineer to the fermentation problem and to provide a description of the microbial population involved in the fermentation. The latter goal envisions long-range but significant impact on technology and has its analogies in the role the microbiologist has had in the development of important fermentation industries (beverages, cheese, yeast, antibiotics, amino acids, etc.) These industries would not have significant economic or public health impact if not for the fundamental understanding of the specific organisms involved in the various processes.

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CHAPTER VIII

IMPLEMENTATION OF ANAEROBIC FERMENTATION INTO AGRICULTURAL OPERATIONS

General acceptance and rapid implementation of energy producing fermentation technology will depend largely upon the economic feasibility of integrating these systems into existing agricultural residue management schemes. Although anaerobic fermentation potentially carries with it a number of attractive advantages, it must render its benefits at a reasonable cost so as not to impart a substantial increase in the total cost of residue management. Certainly, a large factor in an individual farmer's decision to adopt a methane-generating fermentor will be the capital and operating costs of integrating anaerobic fermentation into the farm residue-handling scheme. It is possible that the popularity of the idea of "energy self-sufficiency on the farm" may serve as the main mechanism of promotion among the agricultural community. However, if the inclusion of the anaerobic fermentation process into agricultural residue management systems is to be a lasting trend, the net yearly cost of such an integration step will have to be minimal.

The purpose of this discussion is to briefly survey the economics of various waste-handling systems and calculate the costs of integrating the anaerobic fermentor into some representative management schemes. This is a somewhat formidable task because of the large number of waste handling systems in use today. However, by taking advantage of certain systems analysis techniques, a cost evaluation of the inclusion of anaerobic fermentation into a wide variety of residue-handling systems can be greatly simplified. This chapter describes the utilization of a system's costing model for this purpose.

BACKGROUND

Various systems analysis techniques have been applied to animal waste handling. Coote, et al. (1975), described a mathematical model of a dairy

farm and used linear programming to evaluate some manure management alternatives. Nordstedt, et al. (1971) developed another systems analysis technique, dynamic programming, to study storage and disposal of liquid manure. A network analysis approach was taken by Ogilvie, et al. (1975), to determine the least cost dairy manure handling system. Safley and Price (1976) have also used this method in analyzing dairy manure systems. The latter study provided the computer model which was modified for use in this study.

SYSTEMS ANALYSIS MODEL

One method of evaluating the potential impact of anaerobic fermentation on the total cost of dairy farm residue management is to simply estimate the yearly cost of various residue handling systems with and without the incorporation of anaerobic fermentation systems. A computer program with the capability of performing a rigorous costing of a number of agricultural residue-handling systems simultaneously was developed in a previous study at Cornell (Safley, et al., 1976). This model was modified to incorporate the anaerobic fermentation process into the program's waste management network. Combinations of residue-handling systems considered in the revised systems selection model are summarized in the "original" and "add-on" network flow schemes outlined in Figures 78 and 79, respectively.

The residue management alternatives used in the computer model, Figures 78 and 79, was divided into five phases defined as follows: 1) initial collection; 2) treatment or further collection; 3) transport to storage; 4) removal from storage; and 5) transport to land. All arcs in the network diagrams were numbered for identification purposes. Thus, each system (network branch) consisted of five arcs or phases of operation which had to be separately evaluated and summed up to obtain a total system cost.

The systems in the network were evaluated according to the following net annual cost model:

$$NASC_x = FC_x + CL*L_x + CE*E_x + CD*D_x - CN*N_x - CP*P_x - CK*K_x - CM*M_x \quad \text{Eq.34}$$

Figure 78. Decision network for conventional dairy manure handling systems.

Figure 79. Additional decision network for evaluation of the anaerobic fermentation system applied to dairies.

where:

- $NASC_x$ = net annual system cost of system x
- FC_x = fixed annual cost of system x
- CL = labor cost per unit
- L_x = units of annual labor for system x
- CE = cost of a unit of electricity
- E_x = annual units of electricity for system x
- CD = cost per unit of diesel fuel
- D_x = units of diesel fuel for system x per year
- CN = unit value of nitrogen
- N_x = annual production of nitrogen for system x
- CP = value per unit of phosphorus
- P_x = phosphorus produced by system x per year
- CK = potassium unit value
- K_x = number of units of potassium for system x per year
- CM = cost per unit of methane
- M_x = annual production of methane for system x

The calculation of costs in the network analysis involved approximately 300 variables which were defined in the computer program. These variables included costs of machinery and structures, duration of activities, construction data, daily manure production rates, and other information relative to manure handling systems. Values for these variables were obtained from the literature, surveys of manufacturers, data collection, and construction cost data. Such variables comprised equations representing each arc in the network. Five kinds of arc equations were used in the model describing the following: 1) initial construction cost; 2) fixed annual cost; 3) labor requirements; 4) electricity usage; and 5) diesel fuel consumption.

Credits taken on the fertilizer value of manure material ultimately returned to the land were calculated on an overall system basis for potassium, phosphorus, and nitrogen. The quantity of phosphorus and potassium produced per animal per day was taken to be 0.131 kg (0.29 lb) and 0.261 kg (0.57 lb),

respectively. However, on the basis of literature values, it was assumed that only 60% of the phosphorus and 90% of the potassium effectively reached the field. Calculations of the nitrogen fertilizer value of manure delivered to the land, however, had to be calculated for each individual system because of the high variability of nitrogen losses observed with different manure-handling schemes. A production value for nitrogen was assumed to be a constant 0.314 kg (0.69 lb) per cow per day. Nitrogen losses for systems in the network were estimated on the basis of actual analyses (Safley, 1976) or extrapolations from the literature. These nitrogen loss estimates are summarized in Table 54.

COMPUTER PROGRAM

The computer program used in the calculation of network systems cost was written in PL/C language, tailored for use on a conversational monitor system (CMS) in conjunction with an IBM 370-168 computer. Once logged into Cornell CMS, "METHANE" could then be called. The computer would respond with documentation concerning METHANE and ask a series of questions defining the particular farm operation. These questions, when properly answered, would supply the program with the following information:

1. the number of animals for the system
2. the amount of storage desired
3. costs and design data for particular types of manure storages
4. costs and capacities of solid and liquid manure spreaders
5. appropriate size tractors for the manure spreaders, tractor scraper, and front-end loader
6. sizes and costs of electric motors for manure collecting devices and pumps
7. amounts and cost of bedding per barn type--free stall or tie stall
8. average conditions for the roads and field used for transporting and spreading manure
9. costs of labor, manurial nutrients, and energy.

TABLE 54. ESTIMATES OF NITROGEN DELIVERED TO THE FIELD.

System	% Nitrogen to Field	% of Field Nitrogen in Ammonia Form	% Effective Nitrogen to Field	System	% Nitrogen to Field	% of Field Nitrogen in Ammonia Form	% Effective Nitrogen to Field
1*	80	25	60	24	45	10	45
2	63	35	41	25	45	10	40
3	64	15	54	26*	45	10	40
4*	70	25	53	27	60	42	35
5	66	53	31	28	82	58	34
6	66	53	31	29	82	58	34
7	66	53	66	30	82	58	82
8	84	40	50	31*	40	50	20
9	84	40	50	32*	40	50	20
10	84	40	84	33*	40	50	40
11	80	18	80	34*			34
12	80	18	66	35*			74
13*	82	58	82	36*			34
14*	82	58	34	37*			34
15*	82	58	34	38*			74
16	78	36	50	39*			34
17	54	28	39	40*			34
18	58	25	43	41*			74
19*	80	18	80	42*			34
20*	80	18	66	43*			34
21*	82	58	34	44*			74
22*	82	58	82	45*			34
23*	82	58	34				

* Estimates were made from similar systems.

One important feature of the program, METHANE, was that most of the information significantly affecting the relative economics of the network systems could be inputted by the user himself. The high accessibility of crucial input variables make the program highly amenable to critical re-examination and sensitivity analysis.

Once the user-supplied information has been entered, the computer can perform the analysis and output the results in tabular form for all of the network systems. A list of the tables yielded from the entire computer output is as follows:

1. Initial Construction Cost--This is the complete cost of the system, including necessary machinery and structures.
2. Fixed Annual Cost--This is a percentage of the initial cost of a system based on interest, repairs, insurance, taxes, and depreciation. Also, the cost of bedding material is included.
3. Labor--The per animal time requirements for various activities are used to determine the system's annual labor requirement.
4. Electricity--The user-supplied input included the voltages and amperages of electric motors for the machinery. The annual electricity consumption can then be found by knowing the amount of time a motor is to be run.
5. Diesel--All tractors are assumed to run on diesel fuel. Based on the weight of the tractor-spreader rig, its operating speed, and the power take-off horsepower requirements, the total horsepower demand can be determined. Then, using work done by Hunt (1973), the fuel consumption can be determined.
6. Nitrogen--By chemically analyzing manure samples for nitrogen, a percentage of the as produced manurial nitrogen was determined to effectively reach the field. It was assumed that the $\text{NH}_3\text{-N}$ fraction of the nitrogen that reached the ground was lost unless immediately incorporated into the soil.

7. Total Operating Cost--This is a summation of the annual fixed cost, labor cost, electricity cost, and diesel fuel cost.
8. Net System Cost--This represents the difference between the total operating cost and the fertilizer value of the manure.

TYPICAL RESULTS

The computer program has been used to examine a variety of input conditions to evaluate the costs of integrating anaerobic fermentation technology into dairy farm residue management. As previously mentioned, the economics of using anaerobic fermentation processes may be described by a comparison of the costs of entire residue handling systems with and without the incorporation of methane generation technology. The computer program facilitates such an analysis by providing a rigorous cost breakdown of all systems shown in Figures 78 and 79. A typical set of data for input variables having the greatest potential effect on cost comparisons among waste management systems are tabulated in Table 55.

The results of computer runs of program METHANE using the input data of Table 55 are summarized in Tables 56 through 58 which present representative costing for waste-handling systems with and without the inclusion of anaerobic digestion. Basically three kinds of cost information are given in these tables, annual operating costs, annual credits taken on benefits obtained from the system, fertilizer value of the manure and net methane yields, and the net annual system cost or a summation of the costs and credits. Annual costs for some simple and commonly used residue-handling systems on a 100-cow dairy farm are shown in Table 56. Net annual system costs for these systems varied from around \$1600 for a simple tractor scraper-solid manure spreader system to nearly \$5700 for a gutter cleaner-solid manure spreader scheme. In most cases, credits taken on the fertilizer value of manure delivered annually to the field remained between \$5000 and \$7000. When a conventional manure stabilization process is added to a waste handling system, the net annual cost can be more than doubled as reflected in the costs given in Table 57. For example, a system in Table 57 consisting of an alley scraper, piston pump, earthen storage, and injection land application has a net annual

TABLE 55. TYPICAL INPUT DATA FOR THE PROGRAM "METHANE".

Key Variables		Value
Size of dairy farm		100 cow
Manure storage capacity		120 cows
Length of irrigation pipe		1200 M
Time to drive to the field for spreading		3 min
Interest		5%
Labor Costs		\$3.00/hr
Electricity Costs		\$0.35/KWH
Diesel Fuel		\$.51/gal
Fertilizer Prices		
Nitrogen		\$.44/kg (\$.20/lb)
Potassium		\$.22/kg (\$.10/lb)
Phosphorus		\$.40/kg (\$.18/lb)
Methane Generation Costs		UHD Reactor Conventional Reactor
Capital cost		\$22,600 \$25,700
Net methane production		768 885
l/cow/da (STP)		
Value of Methane gas		2.00-4.00 2.00-4.00
\$/MM BTU		
Labor to run digester, hr/da		.5 1.0
Electricity consumption,		150 2000
KWH/yr		

TABLE 56. NET ANNUAL COST OF WASTE-HANDLING SCHEMES
NOT EMPLOYING ANAEROBIC FERMENTATION FOR A 100-
COW DAIRY FARM.

System Components	System No.	Operating Cost, \$	Fertilizer Value, \$	Net System Cost, \$
Gutter Cleaner-Solid Spreader	2	10,798	5,101	5,697
Alley Scraper-Piston Pump- Earthen Storage-Injection (Land	11	11,007	7,071	3,936
Alley Scraper-Piston Pump- Earthen Storage-Liquid Spreader	12	10,723	6,364	4,359
Alley Scraper-Solid Spreader	16	8,278	5,555	2,723
Tractor Scraper-Stack- Solid Spreader	17	7,295	5,000	2,925
Tractor Scraper-Solid Spreader	18	6,771	5,202	1,569
Tractor Scraper-Piston Pump-Earthen Storage- Injection	19	9,985	7,071	2,914
Tractor Scraper-Piston Pump-Earthen Storage- Liquid Spreader	20	9,701	6,364	3,337

TABLE 57. NET ANNUAL COST OF WASTE-HANDLING SCHEMES EMPLOYING TREATMENT SYSTEMS (EXCLUDING ANAEROBIC FERMENTATION) FOR A 100-COW DAIRY FARM.

System Components	System No.	Operating Cost, \$	Fertilizer Value, \$	Net System Cost, \$
Alley Scraper-Licom Storage-Injection	10	18,999	7,274	11,725
Alley Scraper-Piston Pump- Anaerobic Lagoon- Injection	13	15,652	7,173	8,479
Slotted Floor-Oxidation Ditch-Storage-Liquid Spreader	26	15,124	5,050	10,074
Slotted Floor-Dry Ditch-Solid Spreader	27	14,338	4,798	8,540
Alley Flush-Collection Sump-Anaerobic Lagoon- Aerobic Lagoon- Irrigation	32	15,146	4,040	11,106

TABLE 58. NET ANNUAL COST OF WASTE-HANDLING SYSTEMS EMPLOYING ANAEROBIC FERMENTATION
FOR A 100-COW DAIRY FARM TAKING \$5.00/MMBtu CREDIT ON BIOGAS GENERATED.

System Components	Reactor Type	System No.	Tot. Oper. Cost, \$	Fertilizer Value, \$	Value of Net Methane Production	Net System Cost, \$
Gutter Cleaner-Fermentor	Plug Flow	36	12,077	4,747	4,952	2,378
Earthen Storage-Irrigation	Conventional		13,301	4,747	5,708	2,846
Alley Scraper-Fermentor	Plug Flow	37	13,357	4,747	4,952	3,658
Earthen Storage-Liquid Spreader	Conventional		14,581	4,747	5,708	4,126
Alley Scraper-Fermentor	Plug Flow	38	13,640	7,173	4,952	1,515
Earthen Storage-Injection	Conventional		14,864	7,173	5,708	1,983
Tractor Scraper-Fermentor	Plug Flow	40	12,306	4,747	4,952	2,607
Earthen Storage-Liquid Spreader	Conventional		13,530	4,747	5,708	3,075
Tractor Scraper-Fermentor	Plug Flow	41	12,589	7,173	4,952	464
Earthen Storage-Injection	Conventional		13,813	7,173	5,708	932
Flush-Fermentor-Earthen Storage-Injection	Plug Flow	44	12,076	7,173	4,952	-49
	Conventional		13,330	7,173	5,708	419
Flush-Fermentor-Earthen Storage-Irrigation	Plug Flow	45	10,449	4,747	4,952	750
	Conventional		11,673	4,747	5,708	1,218

system cost of \$3936 while the same system incorporating an anaerobic lagoon (replacing earthen storage) increases the net annual cost to \$8479, as shown in Table 57.

Unlike most of the waste treatment processes mentioned in Table 57, the integration of anaerobic fermentation into a dairy farm residue-handling system seems to have only a modest effect on the net annual system cost if a credit is taken on the net production of biogas. The level of effect can be seen in comparing the costs for the simple waste-handling schemes in Table 56 with costs of similar systems incorporating anaerobic fermentation as tabulated in Table 58. As an example, a system consisting of an alley scraper, plug flow anaerobic fermentor, earthen storage, and injection land application has an annual system cost of \$1515 (see Table 58) compared to \$3936 for the previously discussed example (see Table 56) without anaerobic digestion. Net biogas production credits taken in Table 58 were based on \$5.00 per million Btu. Furthermore, energy production credits were taken as though all of the net biogas produced could be used on site or sold to a pipeline. If a significant portion of the net biogas production cannot be used, this benefit must be reduced.

On the other hand, if all of the gas can be used or sold and if the market price for natural gas rises substantially in the coming years, the economics for integration of anaerobic fermentation may rapidly appear more favorable.

CONCLUSIONS

The cost information from running the program, METHANE, indicates that the real cost of integration of anaerobic fermentation processes into dairy farm waste-handling systems may already be modest, even for the 100-cow dairies. This assumes that construction costs for the simplified plug flow reactors are equal to those given earlier by Jewell et al., (1976). Present activities include the large scale construction and operation of this design at a 65 dairy cow capacity in order to verify this earlier estimate. At natural gas prices not much higher than most unregulated intrastate natural

gas rates, the incorporation of anaerobic fermentation into waste-handling systems may serve, in some cases, to completely defray existing residue management costs by virtue of the increased credit taken on the biogas. Thus the energy value and the potential income from this technology appears to hold promise for even small-scale operations.

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CHAPTER IX

SOCIAL CONSIDERATIONS OF METHANE PRODUCTION ON DAIRY FARMS

INTRODUCTION

The practical feasibility of using anaerobic fermentation on dairies depends on the perception of the potential users. Important aspects of farmer attitudes include such issues as how the new management and labor requirements fit into the existing farm operations, the scientific and mechanical skills required, new government regulations that apply, and a host of socio-technological ramifications.

In addition there are a large number of practical issues such as the farmer's need for an alternate source of energy, for better residue handling systems, and for improved fertilizer options. Anaerobic fermentation systems confront all of these needs and therefore have potential economic benefits beyond the production of methane gas.

Less obvious questions about the feasibility of methane production from dairy and feedlot manure are those associated with the complex web of financial, commercial, and governmental systems. Because these systems may produce more methane than can be used on their own farms, small energy producers (such as dairy and feedlot operators) might face such challenges as collection and distribution problems, constraints of small-scale operations, quality and quantity, reliability, training, and technical support. In other words, in addition to being practical as an on-the-farm energy source, methane production from animal and crop residues must also be practical when viewed as a new and revolutionary commercial industry competing within the mounting pressures and dilemmas of the overall energy industry.

When viewed in this light, critical questions become: What will be the demand for farm-produced methane? How will it be collected and distributed?

How will quality control be maintained? What government regulations apply (the existing natural gas industry is heavily regulated)? What competitive forces will impact on the farmer/producer? How will the independent-minded farmer/natural gas production worker/energy crisis era entrepreneur react to all of these new problems and opportunities? As a farm operator considers accepting or rejecting an innovation, he at the very least implicitly weighs technical, economic, and practical conditions on his farm and within his industry.

The challenge of adapting new technologies to existing social and economic systems is a complex process. Even in the most seemingly straight forward cases, adjustments cannot be made quickly, and the role of any one set of technical experts is rather limited. Two prominent international examples of the technology transfer process will be outlined in this report.

This report will attempt to highlight the implications gathered through preliminary study of the issues outlined above, but because of time and budget constraints will not generate definitive answers. During the months of February through May 1977, 20 dairymen from New York and Vermont were interviewed. In addition, dairy and animal scientists, social scientists, businessmen, financial and insurance officers, and government officials at the local level as well as in Washington, D.C., were also interviewed. The combination of these contacts offers a strong beginning at identifying the issues surrounding the practical feasibility of incorporating methane production from dairy and beef feedlot residues.

An attempt will be made to look at all the systems and all the important elements therein as a first attempt at understanding how the technology will affect the innovating farmer. In so doing the most prominent measures of effectiveness (benefits) and alternatives will be identified. Costs are outlined, including capital inputs but also opportunity costs and labor/management requirements. Risks and reliability are also investigated, including a look at those dairy and beef feedlot managers confronted within their industries as they now exist as well as the uncertainties of the new technology and its impact on their future operations.

The goal of this first look at an ongoing multi-year project is to begin to bring the more critical areas of uncertainty into focus for the complete methane system. This should, in turn, help decision-makers anticipate problems and allocate resources with increased knowledge of systems outside the technical and economic aspects of the problems.

PROJECT DESCRIPTION

It was necessary to provide a relative perspective of the methane technology so that individuals could visually relate the size and possible problems to their own situations. To facilitate this, a scale model kit was constructed containing a large number of farm structures--such as barns, machinery sheds, houses, silos, and farm ponds. All of these could be arrayed on a display board to represent the farmer's actual operations. (See Figures 80 and 81).

The farmer being interviewed would arrange a model approximating the layout of his farm on the board and while doing so would often talk about specific challenges he faced, including waste handling concerns, problems of terrain, the need for extensive repairs on a barn, etc. Although an interview guide was available to insure that all important topics were covered in most cases almost all of these items of interest were covered in the free-flowing discussion generated by the use of the farm-scale model kit.

Description of the Pilot Study

Several issues were raised by almost every dairy farmer interviewed. Other issues were closely related to the size of the farm or age of the farmers involved. Importantly, some issues that would seem to be central are of very little concern to the farmers interviewed. A summary of the findings is presented below. Also included is a brief comparison of a similar study conducted in Worcester County, Massachusetts. Taken together they provide an early indication of dairy farmer attitudes toward adoption of the soon-to-be available methane technology.

The goal of the dairy visits was two-fold. Of primary importance was the need to understand the attitudes of potential adopters toward the emerging

Figure 80. Dr. Harold R. Capener adjusts the models to reflect different arrangements of a dairy operation. The range of sizes and styles of the models enables the farmer to visualize the anaerobic methane generator in relation to his particular farm. The kit's flexibility allows future plans to be taken into account as well.

Figure 81. The Dairy Farm Site Kit was constructed to aid in explaining the anaerobic methane generation system. The kit includes scale models of ponds and lagoons as well as barns, machinery sheds, houses, and other farm structures.

technology. In conjunction with this a good deal of information was needed about the farmer's own operations, particularly regarding waste handling procedures and pollution control issues. A second function was to expose the farmer to the potential of generating methane gas from his dairy wastes.

Because the alternative production systems have not been completely developed by Cornell engineers, detailed financial information was not fully explored of the farmers. However, the "Dairy Farm Information Guide" and the farm-scale model kit did help obtain information about the farmers' attitudes toward methane generation from their farm wastes, their herd size, and other dairy facts. Information was also gathered about the residue handling systems.

The general tone and spirit of most interview sessions was very friendly. Quite often the farmer's hopes and fears for the future were very much a part of the conversation. Most sessions were remarkably free from the rather sterile atmosphere of many questionnaire-based interviews indicating that the approach with the scale models provides a unique basis of assaying public attitudes.

All of the interviews reported herein can be considered a pilot study. Both the farm-scale model kit and the interview guide represent new efforts and little value was seen in prematurely formalizing them. Furthermore, there was difficulty in asking specific questions about alternative forms of methane technology that have not yet taken shape.

The sample was not random, rather individual farms were picked for their general size and availability. Several of the interviews were conducted while on a residue management systems tour in the states of New York and Vermont. This tour itself preselected larger and probably more articulate and innovative farmers. In fact, most of these farmers were openly looking for innovations to adopt. That was the major reason they were participating in the tour.

The combination of new interview tools and techniques, the tentative technology, and the sample of dairy farmers interviewed made for an interesting pilot study of farmer attitudes toward adoption and adaptation of methane generation from animal residues.

The interviews ranged from about 45 minutes to almost 3 hours. All farmers were open, cooperative, and interested, and on occasion other members of the family became involved.

Although the farm scale model was a very useful device over all, in some cases it had limited value. Some farmers became overly concerned when they could not find an exact representation of their barn or other features of their operation. In those cases the model seemed to inhibit an articulate discussion rather than foster a free-flowing conversation.

From the interviewer's viewpoint, however, the model often seemed to open up a good mechanism for outlining the anaerobic fermentation concept and the various alternatives which might in time become a recommended system. In almost all cases it was interesting to see oneself as well as the person being interviewed discussing the subject at hand with arms outstretched and fingers pointing, moving models around and actively participating in an interview session.

In the most encouraging cases, the farm scale model may have actually triggered some original planning on the part of the farmer. In any event, it is an interesting methodological device and deserves further attention and development.

Analysis of the Findings

Perhaps the most important implication for methane technology transfer to come out of this investigation is the fact that there is a residue handling revolution now in progress in the dairy industry. All farmers are generally aware of the various alternative systems available or in testing stages. Most seem to have active plans for constructing more advance handling systems and a number had new systems already in place.

Not all of the farmers residue handling problems have been solved by these systems and the farmers interviewed are generally well informed as to the capabilities and limitations of each alternative. Many say they are waiting

a short while longer because they are hoping for a more definitive answer to their residue handling/pollution control problems.

Virtually all of the farmers interviewed had heard of several solutions to dairy residue handling problems. Many expressed concern that there were no full-scale demonstrations of methane production systems being studied at Cornell. The farmers were very quick to spot unsolved problems in methane production. The general feeling was that methane generation is still undeveloped and experimental, it is still too far away to be given serious or detailed consideration at this time. What the respondents wanted was more information but most importantly a demonstration model.

On the other hand, the farmers were willing to talk in considerable detail about the major farm management issues and currently available solutions. They appeared to be very knowledgeable about various residue handling methodologies, pollution standards and enforcement of regulations. Most farmers were found to be both conscientious and conservative in their use of energy supplies. They clearly placed a premium on independence and self-reliance. When methane generation can be positively demonstrated they will probably quickly become a very interested audience.

Most of the farmers interviewed use little hired help, although generally the larger the farm, the greater the requirement for outside labor. Smaller dairy farm operators know that their numbers are shrinking and they are very concerned about the economics of their businesses. Since many are nearing retirement they are understandably conservative regarding additional capital expenditures in the coming years. In contrast, the larger dairy owners seldom mentioned capital constraints. When they learned of the probable cost of the reactors they did not seem surprised.

It should be emphasized again that most of the farmers interviewed saw the value of the anaerobic fermentors as an alternative answer to their residue handling problems in that it assists in rendering the waste odorless and allowed long-term storage. This was especially true for the 100 to 200

head-size operations, where the owners were quite aggressive in their search for a solution to these problems.

Agricultural engineers and dairy scientists who are not working on the methane project directly also seem to view methane systems as yet another residue-handling alternative, with the gas produced being an added plus. Until arrangements can be explored for marketing the excess methane gas produced, the real benefits may accrue to residue handling.

An equally important impression to come out of this process is how site-specific the adoption process will have to be. Each operation has significantly different opportunities and limitations. Not only do such characteristics as soil and drainage conditions differ among adjoining farms, but so do the perceived energy requirements or the adaptability of the operations to a liquid wastes handling system that must accompany a methane fermentor.

For example, on several of the larger dairies the ownership was shared between a father and one or two sons. Since there was usually only one family residence near enough to the probable reactor site, only one family could take direct advantage of the methane produced. The others often lived miles away.

Farmers do not seem to be anxious to commit more of their management time. Importantly for its design implications for the Cornell project, they also do not want to accept more sophisticated labor requirements. They tended to ask specific questions about the reliability of the reactors. The implication was that they have to depend on reliable power and cannot afford to have their family or farm subjected to a system that breaks down or is too technical for them to manage well.

Briefly summarizing, many larger dairy farmers are planning or have already installed some type of automatic residue handling system. Even though they are knowingly taking some risks, since no system is fully tested or without some limitations, they perceive the problems of an untried system as less bothersome than daily spreading of manure on land and the threat of pollution control regulations. These larger farm operators appear to be willing to take

risks and spend capital if the capabilities and practicality of the system can be demonstrated. They do not anticipate a trouble-free solution and are quite accustomed to modifying almost any system they purchase to their particular farming site and operation.

The Massachusetts Study

In 1975 a study was carried out with 40 dairy farmers in Worcester County, Massachusetts (Cole, 1976). Two college students conducted the interviews. Several of the findings compare very closely with those obtained from dairy farmers in New York and Vermont with several significant exceptions.

The purpose of the Massachusetts dairy farm study was to "determine the circumstances relevant to the proposal of methane generation on farms." The study concluded that if methane generation is found to be economical, its adoption would be seriously considered by many farmers. The researchers found that a methane generation system could fit rather easily into a dairy farm's ongoing operations and that there should be a ready on-farm demand for the gas.

Forty farms were sampled, and the distribution was, by and large, representative of the county in terms of herd size as well as geographical coverage.

The questionnaire was not specific with respect to the type of methane technology that would be adopted. The farmers responding were asked about their attitudes toward investments and payback schedules for \$5,000, \$10,000 and \$20,000 for 5, 10, and 20 years each. In essence, the farmers were asked about their attitudes toward a hypothetical methane production system. They were asked much more specific questions about herd size, equipment investment, fuel usage, and farming operations generally.

The Worcester County study concluded that the type of farmer who might be most interested in future methane production on his farm is: under the median age (47), with higher than average education, and already knows a good deal about the methane potential on his farm. He will place a high value on self-sufficiency and consider himself to be innovative. Lastly, he will have at least 50 cows and a gross annual income of over \$40,000.

The Cornell study has attempted to be more specific about the technological features of the methane gas generation, its estimated price range, and other features of size, location, performance, and potential benefits. While the Massachusetts study pointed toward methane generation as the key variable, the New York and Vermont farmers focused their attention on the residue handling issues where the methane generator would render the animal waste materials odorless, thus making it possible to consider prolonged periods of storage while promoting a more efficient reclamation and utilization of the nutrients.

The Massachusetts study concluded, among other things, that although the dairy industry was facing serious difficulties and future growth would require large labor and capital commitments, even so, it was found that farmers are accustomed to sizeable investments and the price of methane reactors would not surprise them. However, compared with other demands for capital a methane production system might be viewed as a low-priority item. Economics and practical management issues played a major part in the Worcester dairy farmer's decision making. There was a perception expressed that methane generation from animal wastes was still experimental and too far in the future to merit serious consideration. They too want to see the working models.

A few important differences are noted between the findings of the Cornell pilot study and the Massachusetts project. First, the Worcester study found little interest in new manure management systems. Quite the opposite was true in New York and Vermont. Second the Worcester study dairy farmers were not especially concerned with any present or future waste-handling/pollution regulations, and only planned to alter their operations if forced to do so by regulating officials. By 1977 the New York and Vermont farmers were highly sensitized to the environmental pollution regulations, especially as they impacted on waste-handling operations.

From both of the studies there is favorable evidence that a profit-making anaerobic reactor would be viewed with interest by many dairy farmers, especially as other incentives seem likely to grow in importance. Such factors as increasing energy costs and scarcity of natural gas and other fuels will

contribute. Expanding environmental regulations will probably also make the methane alternative more attractive.

Innovation, the System, and the Farmer

The implications of adopting the methane technology have a wider consequence than the impact of the innovative farmer. Several other connecting systems are involved. This study, in turn, has implications for wider technology transfer issues such as the creation of a new industry, energy and agricultural crisis control, government policy-making, and education and development. Perhaps the best way to highlight these additional facets is to draw from two well-known parallel examples.

New agricultural development projects have a relatively long history and tradition, domestically and internationally. This tradition serves to illustrate several important lessons regarding the development of technological innovations and their adoption by farmers. A brief review of these key insights can provide valuable guides as the next phases of the methane project are planned.

First, it has been found that technology transfer is complex, takes time and impacts very evenly on the related components of the larger system of which the technology is a part. By way of illustration, the experience of the "Green Revolution" can be cited.

In the press for increased food production certain technologies saw the development and transfer of new seed strains as the basic component of what came to be called the "Green Revolution." Others foresaw the process as not only related to an arduous genetic breeding program, the products of which be transferred to farmers, but also as one element in a network of related support systems of agrarian change.

The differences between these two perspectives help explain why the "Green Revolution" has not met with the expected overall success. On one hand, although higher yielding genetic varieties have been developed, they have

not, general in, resulted in greater agricultural production per capita, nor reduced malnutrition. In fact, according to many authorities, the total direct quantitative effects have been meager. On the other hand, the negative indirect effects have been noticeable and measurable. With the growth of a market-oriented crop system, a growing agricultural labor force, increasing power of the landowners and the declining share of wages, the "Green Revolution" has been significantly associated with economic inequality.

Basically the "Green Revolution" and the technology that formed its core were beyond the reach of the small farmers who comprise the bulk of the agricultural systems which needed the positive change. Several reasons can be cited as to why average farmers had difficulty benefiting from the improved technologies:

1. Risk of failure. Although small, these risks were often too great for the marginal operator.
2. Loss of access to scarce resources as owners of larger farms used up the technologies (seed, fertilizer, water, credit, training).
3. Low levels of political power limited their access to a hearing, a review of the distribution of the scarce resources or a part in the planning and policy process.

Stated in other terms, the flow of scientific information in the form of improved seeds and other technologies was far from perfect. Even worse was the lack of any reverse flow of information from the farmers back through the channels to the original technicians about the constraints they were facing in gaining access to the technology.

From the "Green Revolution" and most other development programs, two general conclusions can be drawn. First, development (which almost always includes adoption of a new technology) is complex. Historical economic processes do not often lend themselves to simplistic, quick adjustments. Second, the potential role of any specialized group of technicians, or developers, is rather limited when the system wide context of developmental change is adequately considered.

While the specialized technical groups are obviously indispensable in the original development of the technological advance (seeds or methane digesters) the fact remains, however, that technical experts (or any other single set of individuals) are but one link in a rather long development chain.

Placed in the larger perspective, then, it should be emphasized that methane production is also an industrial and governmental challenge as well as an agricultural or technology transfer exercise. Extensive use of the potential for producing methane from the biomass (or from animal manures) will quickly bring regulatory transportation, and distribution challenges to the surface.

For example, consider the possibilities of excess quantities of methane gas being produced on dairy farms or feedlots. It may then become practical to have a fleet of bulk trucks collecting methane from such units and delivering it to central depots in much the same manner as bulk milk trucks now operate. This then would require new and probably highly technical innovations on the part of the truck lines and would also require a significant number of important behavioral and managerial changes within all affected systems including the natural gas industry.

Perhaps the scope of the undertaking is not unlike the second example, the attempt of the Chinese to develop a backyard steel industry. By way of a brief review of this event--in the summer of 1958 China launched a radical new program for speeding up industrial development. The leadership under Mao Tse-tung abandoned the Soviet model of industrial development and initiated their own "Great Leap Forward" which was a cooperative movement of grand dimensions. The new communes averaged about 5000 households per unit and were multifunctional.

In addition to agriculture, the communes and the production brigades within them were expected to undertake industrial production. The small-scale steel plants established at this time have become the primary symbol of the economic disaster that the great leap forward brought. Not only were the production brigades poorly qualified as steel producers but none of the supporting

infrastructure was ready to provide the required support. Within a few months the movement was restrained. Massive economic dislocations were evident as a direct result of uneconomical production methods being attempted in the steel plants run by the communes.

The dairy methane producer, much like the backyard steel productions in China, would be largely independent and relatively isolated. It is recognized that the problems of management, scale, quality control, and distribution are much different between the existing natural gas industry and a biomass based methane gas system just as the existing steel industry is vastly different from the one attempted in Communist China. While the complexity of steel manufacture far exceeds that of anaerobic methane fermentation on a small scale, it may be that some of the lessons apply.

Farm-produced methane will surely take years to implement on any large scale, and will involve a great number of farmers, policy-makers, businessmen, and support personnel. It will also involve cooperation, training, leadership, organizational and individual behavioral change.

Cornell's present project provides a valuable opportunity to create the technological design of a methane reactor adapted to small-to-medium-scale U.S. farms up to larger scale beef feedlots.

The challenge that remains is to generate appropriate research funding to the technology transfer portion of the equation. Here the tasks will be to study the range of conditions conducive for farm operators to seriously consider incorporating methane fermenters into their farm operational systems. Obviously a practical, economical, and technologically reliable anaerobic methane production system will impact on current waste handling systems, national energy systems, pollution control and other government policies, credit and other financial systems, farm labor patterns, family behavior, and operator/manager skills. Additionally, it will have important effects on national and other levels of policymaking and will change and be changed by the educational and extension systems.

SUMMARY AND CONCLUSIONS

Several implications seem to stand out from this pilot study of the socio-economic concerns of the feasibility of on-the-farm anaerobic methane production. At least in the pioneering stages, the advantages to individual farmers of methane production from dairy cattle manure will depend greatly on specific characteristics of the farm sites. This is one of the clearest lessons that has come out of the discussions with farmers in New York and Vermont (and an impression which has been consistently reinforced by the secondary investigations of the beef feedlot industry). A large number of site-specific elements comprise the critical factors in any feasibility analysis or, more importantly, the decision-making processes that the farm operators will go through.

For example, factors such as age of the farm operator, size of the dairy enterprise, condition and value of existing buildings, nature of the manure-handling systems, topography of the farmstead, presence of other capital equipment, location of the farm house (is it close enough to use methane for energy?), long-range plans of the farmer (are there sons who are interested in continuing the operation?). All of these variables tend to be farm-site specific in terms of adoption decisions.

A brief review of regional differences in terms of design considerations may be in order. Water is the major constraint for agricultural operations in many parts of the United States. In some sections irrigation is the highest farm energy cost. In arid regions there simply may not be enough water remaining to conduct all the traditional agricultural operations, let alone using more water to liquify manure for methane generation. This will mean that technology to accept high solids concentrations such as discussed in Chapter 4 would have to be available. On the other hand, methane can supply the power for irrigation.

Perhaps the most obvious regional difference is climate. In the hot dry Southwest evaporation plays a major role in waste-handling systems, particularly with beef feedlots. The moisture (and nitrogen) is lost very rapidly and in some feedlot cases manure dries out very rapidly. These conditions are vastly

different from cold, humid regions where it seems very difficult to manage a dry-manure handling system.

In several sections of the country there may be alternative energy-producing technologies competing for the attention of the farm operator. For example, in some high concentrated beef feedlots areas the conditions may be attractive for wind-powered energy alternatives. In the coming years, a farmer may have to choose whether he should innovate with methane and produce gas or innovate with wind power and produce electricity. It is unlikely that many farmers will adopt two radical new energy technologies at the same time.

Labor costs are another regional variable that will bear investigation in greater depth. Since labor is such a high cost item, any regional variations will have considerable influence on the cost effectiveness of the developing technology.

Feed is significantly different in various sections of the country. What effects do various feeds have on manure composition and, finally, the anaerobic process in a fermenter? Will a farmer be required to check with a methane technical expert before he changes his rations or includes any chemical additives? As the research progresses these and other regional issues need to be confronted. From the preliminary study, it is clear that regional and farm-site differences are very important and assumptions valid for one region may not hold across other regions.

In cases where small dairies have no residue disposal problem, where odor control is not a concern, and where daily land disposal of the waste remains an acceptable practice, utilization of the anaerobic fermentation process would have to be primarily adopted for energy generation alone and all associated costs would have to be justified by the value of the energy. Under these conditions there would be little incentive for producing methane in excess of what could be used on the farm or incorporated into some type of larger distribution system.

Other potential benefits may be realized but are difficult to quantify. For instance, what is the value of improved environmental quality through more

effective runoff or odor control? When environmental pollution reaches extreme stages, these may be some of the major contributing factors in terminating a farm operation. It is not uncommon to find migrating suburbanites settling next to farming operations who subsequently object to the odors associated with animal production and feel these should not be part of their environment. In such cases, odor control technology may have a high value, since this may be the critical feature determining whether the farmer can continue his operation. To date, little attempt has been made to place dollar values on such technology. Rather, the approaches used have been to briefly consider benefits, assign a yearly value to them and calculate the net energy produced in each operation per million BTU's generated, and then express these benefits in dollars. During the coming years these qualitative issues will undoubtedly demand further investigation.

In summary, the major benefits available on dairies and beef feedlots would be the possible labor reduction, timeliness of waste handling, odor control, more efficient nutrient recovery, and energy production. In some feedlots the value of residues may be large enough to provide the basis for recovery of the residue for refeeding purposes. The combined value of these benefits would vary from a high of $\$80/10^6$ k cal (\$20 per million BTU) on small dairies to about $\$39/10^6$ k cal (\$10 per million BTU) on a larger beef feedlot. These benefits seem to equal or exceed the cost of generating the energy in all farming systems.

The evidence is strong that decisions made in agriculture are not solely based on the economic feasibility of adopting a specific practice, but often include such intangibles as competitive price, desire to increase productivity with a fixed labor base, and motivation to produce a better product. The crucial feature of adoption of anaerobic fermentation on farms will be the operators' acceptance of it. Not only must the proposal be technologically feasible and economically beneficial, but such a system must fit into the flow of ongoing operations. In this regard, an important barrier to small dairy and feedlot operation methane generation systems may be the lack of ready markets for exporting the excess gas from the farm, credit limitations,

lack of management skills, and similar issues, rather than technological constraints. An efficiently designed fermentor system will have positive effects upon:

1. Labor timeliness and seasonal efficiency in residue management operations.
2. Pollution control. The contamination of water through runoff, odors, and the number of flies and other pests will be better controlled.
3. Conservation. In addition to methane, nitrogen, phosphorus, other valuable fertilizer compounds, and solid materials can be returned to the land or otherwise be conserved and utilized.
4. Energy. At a minimum, farm-produced methane can be used efficiently on the farm, thereby releasing existing energy supplies to be utilized by other consumers. Ideally, the excess methane can be distributed through other commercial natural gas suppliers.

The fraction of the total investment on dairies and beef feedlots that is associated with residue management is less than 10 percent of the total investment. If a new investment in a methane reactor decreases a farmer's environmental and waste management problems and provides enough heat for his home and buildings, this may be enough to convince him of the value of adopting this practice even though he may have to expend between \$15,000 and \$30,000 for a unit.

At present, however, with no standard for regulating methane production, with no technical backup service and with no market for excess methane production there is very little in the way of incentives for the dairy farmer, other than improvement of his waste-handling mechanism. Essentially, the anaerobic system is seen as a new and complex "hassle" by many farmers.

Considering the larger system-wide approach that needs to be thought of much of the selling job must be directed at policy-makers, energy experts, and natural gas consumers. A market and marketing system will need to be established. When the economic impact is great enough, the remaining farmers will probably respond to the new technology.

It will be relatively more feasible to adapt anaerobic fermentation to an existing liquid manure-handling system than to a dry manure-handling system. At present the farmers seem to know that there are no completely tested alternative solutions to their waste-handling problems. Most in the New York-Vermont region seem to be leaning toward a liquid system. They seem to want the manure out of the barn and into storage as automatically as possible. Labor demand rather than cost seems to be the primary constraint on most medium to large dairies.

If the focus of those who are responsible for the technology of methane production remains too tightly aimed at gas production alone, without facilitating the companion features of the larger systemic elements, a disjointed developmental process can surely be predicted.

The implications of the larger system elements deserving attention are as follows:

1. The farmer adopting a methane generating system will almost certainly be impacted by several government agencies having special interests in financing, regulation, control, marketing, distribution, etc. These could include several units within the U.S.D.A., such as the Farmers' Home Administration, the Cooperative Extension Service, and the Economic Research Service. Other agencies would include the Environmental Protection Agency, the Department of Commerce, the U.S. Public Health Service, and the Public Utilities Commission.
2. Conceivably unless some of the system-level components are coordinated in relation to this approach to energy alleviation, the farmer could find himself in a position of financial, regulatory, marketing, or some other form of chaos.
3. More systematic study is needed of the implications of technological development of methane production such as the necessary infrastructure to handle realistic transfer and application of the technology.
4. More systematic study is needed of the farmer's site-specific problems and his needs, capabilities, and resources. This kind of intelligence

information then could be fed back and front loaded in the technology design processes to forestall technical errors, adjustments, and surprises.

5. The frontal attack on the energy questions is obviously complex and therefore demands a multifaceted/multidisciplinary approach.
6. Mobilizing the team approach and facilitating continuing developmental work such as that being proposed by the cadre of physical, biological, and social scientists at Cornell would seem a sound and a wise policy.
7. While the first step in development of the technology is taking place, even its successful completion is inextricably related to the questions of methane for what purpose? For what uses? By whom? Under what farm site conditions?
8. What is needed to optimally pursue the questions raised in this report is a source of funds that will allow the utilization and transfer issues to be simultaneously explored with methane generation technology. Up to this point it has been difficult to find any funding source with timely response. It is hoped this report will serve to highlight the need to pursue a more balanced plan of methane gas generation for dairy and beef feedlot operators.

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CHAPTER X

ESTIMATING THE FEASIBILITY OF USING AGRICULTURAL CROPS FOR ANAEROBIC FERMENTATION - INTRODUCING THE ENERGY-FOOD INDEX FOR EVALUATION OF BIOMASS FUELS

INTRODUCTION

The concept of using fuels from biomass as the major source of energy in the future is receiving increasing attention. Although the near term production of energy from agricultural residues appears limited to a maximum of 10 percent of the total U.S. energy needs (Jewell *et al.*, 1976), some estimate that direct and indirect use of solar energy should become the only source within 100 years, except for a small amount that would be generated from renewable sources such as wave and hydroelectric power (Lovins, 1977).

A simplistic analysis indicates that the total present U.S. energy needs could be met by converting crops to fuels, if about 20 percent of the surface area of the U.S. was used and it produced an average of 5.0 biodegradable organic tons (dry) per year. But the utilization of huge amounts of prime cropland for fuel production instead of food production raises some obvious conflicts. If the proposed large land or ocean-based "energy plantations or farms" utilize undesirable or marginally productive areas, the total area requirement would increase significantly due to the decreased productivity of these areas. After resolution of these and other conflicting issues in the open market place it is doubtful as to whether large energy farms will be feasible if they are developed solely for energy generation. However, most food crops provide only a portion of the total fixed solar energy in the form of animal food with the remainder resulting as agriculture residues. For example, about 600,000 acres of corn are produced for grain in New York State, and more than one million tons (dry) of organic matter is left in the fields to become part of the soil. If the corn stalks were harvested for energy production, the crop would produce both energy and food.

Energy farms that would utilize this presently unused agricultural residue would have the capability of generating about 6×10^{12} Btu per year or about an amount of energy equal to \$12 million per year. However, during the fermentation process new bacterial cells would form during methane production and if these were harvested they would result in the production of 18,000 tons per year of new single cell protein - an amount of protein produced on about 20,000 acres of cropland. On a National scale the production of one quad of energy (10^{15} Btu) would result in the synthesis of a quantity of protein equal to that produced by nearly 3 million acres of corn. Thus the ideal energy farm should incorporate the concept of dual maximization of both fuel and food production. This chapter presents preliminary development of a procedure to evaluate the energy production potential and the food value of various crops. Such an index would be useful in comparing alternate crops and make consideration of the energy farm more realistic in terms of maximizing the value of the outputs. It would also make the energy farm much more acceptable since it would not use cropland for the production of energy products alone.

Energy and food values will be combined to form Index values. A variety of methods for computing values for food and energy and for combining them will be explored here. It is less important to calculate actual values for all crops at this time than it is to demonstrate methods that may be easily used in developing this concept. For the sake of brevity, only eleven crops were used in this study: barley, brown rice, cabbage, corn, field peas, oats potatoes, rye, soybeans, sweet potatoes, and wheat.

Data Requirements and Definitions

The total mass of the food plant may be conceptually divided into two parts, food and unused residues. The food portion of the plant is the total mass, including water, that is removed from the field as a human food product. It is the part of the plant that is used to compute statistical yields of the various food crops. This consists of water, vitamins, carbohydrates, protein, fiber and other dry matter which includes some nutrients.

The residue portion of the plant consists of water, volatile organic solids, and ash. Energy in the form of methane gas is derived through anaerobic fermentation by the destruction of volatile solids, so the relative size of this portion of the plant will indicate the energy value of the crop, assuming that 50 percent of the volatile solids are destroyed in all cases. Of course, this is a simplifying assumption that needs to be refined in future development.

The following nine variables are defined in Figure 82 and are used in the development of the energy/food index.

1. Yield - The wet weight tons of food per acre, under average management and climate conditions.
2. Food Dry Matter - Dry weight of protein, carbohydrates, and other dry matter in the food part of the plant.
3. Residue Dry Matter - Dry weight of volatile solids and ash in the waste part of the plant.
4. Food Dry Matter Content - The wet weight food dry matter fraction of the food part of the plant.
5. Dry Matter as Food - The fraction of the total dry matter in the plant that is in the food portion of the plant, or the food dry matter.
6. Carbohydrate Content - The wet weight carbohydrates fraction of the total yield.
7. Protein Content - The wet weight protein fraction of the yield, independent of amino acid composition.
8. Waste Dry matter Content - The wet weight waste dry matter fraction of the waste.
9. Ash Content - The wet weight ash fraction of the waste.

The assumed values of each of the seven relevant variables are given for the eleven crops in Table 59.

Figure 82. Categories of plant food matter and plant residues..

TABLE 59. YIELD (IN TONS PER ACRE) OF FOOD FOR HUMAN CONSUMPTION ONLY, AND PLANT MATTER CONSTITUENTS (ALL WET WEIGHT FRACTIONS).

Crop	Yield	F O O D				R E S I D U E	
		Food Dry Matter Content	Dry Matter As Food	Carbohydrate Content	Protein Content	Residue Dry Matter Content	Ash Content
Cabbage	10.17 ^{12/}	.076 ^{11/}	.90 ^{5/}	.053 ^{11/}	.022 ^{2/}	.158 ^{2/}	.030 ^{2/}
Corn	2.58 ^{12/}	.251 ^{11/}	.45 ^{8/}	.205 ^{11/}	.037 ^{11/}	.927 ^{3/}	.092 ^{3/}
Oats	0.71 ^{12/}	.923 ^{4/}	.34 ^{7/}	.607 ^{4/}	.125 ^{4/}	.902 ^{2/}	.065 ^{4/}
Field Peas	1.28 ^{12/}	.161 ^{11/}	.65 ^{9/}	.113 ^{11/}	.035 ^{2/}	.901 ^{13/}	.061 ^{13/}
Potatoes	10.61 ^{12/}	.222 ^{11/}	.82 ^{5/}	.191 ^{11/}	.022 ^{2/}	.212 ^{2/}	.013 ^{2/}
Brown Rice	2.08 ^{12/}	.880 ^{11/}	.34 ^{7/}	.777 ^{11/}	.091 ^{2/}	.935 ^{4/}	.219 ^{4/}
Rye	0.62 ^{12/}	.900 ^{11/}	.34 ^{7/}	.752 ^{11/}	.112 ^{11/}	.894 ^{3/}	.037 ^{3/}
Soybeans	0.76 ^{12/}	.330 ^{11/}	.28 ^{6/}	.060 ^{11/}	.380 ^{10/}	.640 ^{11/}	.008 ^{11/}
Sweet Potatoes	4.17 ^{12/}	.315 ^{11/}	.64 ^{1/}	.279 ^{11/}	.016 ^{2/}	.892 ^{2/}	.024 ^{13/}
Wheat	0.83 ^{12/}	.894 ^{4/}	.34 ^{7/}	.716 ^{4/}	.135 ^{2/}	.926 ^{3/}	.167 ^{3/}
Barley	0.93 ^{12/}	.898 ^{11/}	.34 ^{7/}	.728 ^{11/}	.121 ^{2/}	.869 ^{3/}	.060 ^{3/}

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ENERGY/FOOD INDEX DEVELOPMENT

The next step in developing the index is to compute the values of the crops in terms of both energy and food per unit of yield. It will also be useful to calculate the tons of protein and carbohydrate, as well as the cubic feet of gas, that may be expected per acre under average yield conditions.

The food value will be reflected by the tons of protein and carbohydrates per ton of yield. The simplest means of combining protein and carbohydrates is to value them equally, by just adding the carbohydrate content and protein content columns in Table 59. For example, barley would have a "food value" of 0.85 tons of carbohydrates and protein per ton of yield. A more elaborate method for computing a food value could reflect the extent to which the food contributes to the minimum human dietary requirements of protein and carbohydrates. According to one source, a 70 kg man would require 60 grams of protein per day, and 2800 kcal of energy (NAS 1968). If all of this energy is obtained from carbohydrates, 700 grams of carbohydrates are required daily at 4 kcal per gram. This is a ratio of 6 to 70 for protein to carbohydrates. The food value could therefore be computed from weighted values of protein and carbohydrates:

$$\text{Food Value} = (\text{Protein Content}/6) + (\text{Carbohydrate}/70)$$

For example, the food value for barley computed in this way would be $0.121/6 + 0.728/70 = 0.031$. Note that the food value computed in this way is more like an index in that it has meaning only when it is used as a comparison to other values.

Other methods of computing the food value are also possible. The percentage of digestible nutrients can be substituted for the sum of the protein content and carbohydrate content; or a totally different approach may be used in which the prices of protein and carbohydrates, as well as for energy, are the respective weights for the components of the index. This would not be a static index, as are the other indices discussed here. Rather, it would

require constant revision and updating as prices change over time. Such an index would include factors such as scarcity when they are included in the price.

It is possible to enhance the animal food value of a crop through the use of anaerobic fermentation of the residue material. The microorganisms that result from the process appear in the effluent, and may form a substantial source of protein, although the availability of this protein may be as low as 50 percent of the total. The amount of protein produced can be approximated by:

$$P = VS \cdot PD \cdot BY \cdot PB \quad \text{Eq. 35}$$

where: P = Weight of Protein Synthesized.
          VS = Weight of volatile solids available for conversion to methane.
          PD = The fraction of volatile solids actually degraded. Its average is taken as 0.5, but may vary by as much as ± 40 percent for plant wastes, as it does for animal wastes.
          BY = Weight of bacterial yield per weight of volatile solids destroyed, assumed to be 0.2. This is also highly variable depending on the fermentation process design.
          PB = Weight of protein per unit weight of bacteria, assumed to be 0.4.

Therefore, a pound of volatile solids converted to methane should produce an average of about 0.04 pounds of protein (Table 60). However, the fraction of volatile solids destroyed may be as low as 0.3, and the quality of the protein may be only half that of the food part of the plant. So a pound of volatile solids destroyed may produce only 0.012 pounds of protein; but if the protein is of equal quality, and the fraction of volatile solids degraded is 0.7, a pound of volatile solids could produce as much as 0.056 pounds of protein. These are the most pessimistic and the most optimistic possibilities, respectively.

TABLE 60. PROTEIN ADDED TO THE FERMENTED RESIDUE BY MICROBIAL YIELD.

Crop Residue Fermented	Protein ² Added Minimum (%)	Protein ² Added Average (%)	Protein ² Added Maximum (%)	Rank	Synthesized ¹ Protein	
					Pess.	Opt.
Cabbage	0.38	0.95	1.78	11	0	0
Corn	9.31	22.27	43.44	7	.003	.016
Oats	15.96	39.91	74.50	4	.020	.093
Field Peas	2.78	6.94	12.78	8	.001	.005
Potatoes	2.51	6.27	11.70	10	.001	.003
Brown Rice	17.25	43.12	80.49	2	.016	.073
Rye	17.95	44.87	83.75	1	.020	.094
Soybeans	2.65	6.62	12.35	9	.010	.047
Wheat	12.64	31.60	58.99	6	.017	.080
Barley	16.10	40.24	75.11	3	.019	.091
Sweet Potatoes	12.90	32.32	60.20	5	.002	.008

¹ tons per ton of yield

² weight of protein synthesized per weight of yield/weight of protein in food per weight of yield

The additional protein provided through synthesis may be crucial in determining the viability of a crop as both an energy and food source. A food value that reflects dietary requirements may be considered less accurate because diets vary widely.

The energy value of a crop is the amount of gas produced through the destruction of volatile solids in the residue portion of the plant per unit of yield. The residue of the plant includes roots, stems, or other non-food parts, not typically removed from the field for human consumption, but that which is easily removed. Since the efficiency of the destruction of volatile solids is not known for each of the eleven crops, it is assumed that 50 percent

is converted to gas in all cases. Therefore, the energy value will be proportional to the amount of volatile solids available in the residue, per unit of yield. The weight of volatile solids that may be expected per unit weight of yield of food can be derived from the input variables:

$$\frac{\text{Weight VS}}{\text{Weight Yield}} = \frac{(\text{Food Dry Matter Content}) * (\text{Dry Matter as Residue}) * (\text{Volatile Solid Content})}{(\text{Dry Matter as Food}) * (\text{Residue Dry Matter Content})} \quad \text{Eq. 36}$$

where: Dry Matter as Residue = 1 - Dry Matter as Food
               Volatile Solid Content = Residue Dry Matter Content - Ash Content

A summary of the first part of the index identified with the food value is shown in Table 61. This illustrates three approaches to evaluating the food index of a crop.

The quantity of gas that can be expected per acre for average yield conditions is as follows:

$$\frac{\text{Gas}}{\text{Acre}} = 6000 * \text{Yield} * \frac{\text{Weight VS}}{\text{Weight Yield}} \quad \text{Eq. 37}$$

The above expression assures that all crop residues are 50 percent biodegradable and that 12 ft³ of gas is produced per pound of volatile solids degraded.

The gas production potential and energy values are given in Table 62 for the eleven crops.

The last step in developing an index involves combining the food and energy values in such a way that the comparison of their values lead to differentiation of the combined energy and food values. Simply adding the energy and food values will give index values that are difficult to interpret. The influence of the energy value would be greater because it has a higher range (0.007 - 1.675) than the protein + carbohydrate food value (0.075 - .868). This would also imply that a ton of volatile solids has the same value as a

TABLE 61. CALCULATED FOOD VALUES USING THREE DIFFERENT APPROACHES.

	Protein* + Carbohydrate* 1	Protein* + Carbohydrate* + Synthesized Protein* Pess. Opt. 2 3		Protein/6 + Carbohydrate/70 4
Cabbage	.075	.075	.075	.004
Corn	.242	.245	.258	.009
Oats	.732	.752	.825	.030
Field Peas	.148	.149	.153	.007
Potatoes	.213	.214	.216	.006
Brown Rice	.868	.884	.941	.026
Rye	.864	.884	.958	.029
Soybeans	.440	.450	.487	.064
Sweet Potatoes	.295	.297	.305	.007
Wheat	.851	.868	.931	.033
Barley	.849	.868	.940	.031

* Tons/Ton Yield

TABLE 62. ENERGY AND FOOD PRODUCTION PER UNIT AREA FOR FOOD CROPS.

Crop	Cubic Feet of Gas Per Acre	Energy Value (Tons of VS Per Ton of Yield)	Tons of Protein and Carbohydrates Per Acre
Cabbage	427	.007	.763
Corn	4,443	.287	.624
Oats	7,084	1.663	.520
Field Peas	622	.081	.189
Potatoes	2,928	.046	2.260
Brown Rice	16,324	1.308	1.805
Rye	6,231	1.675	.536
Soybeans	3,821	.838	.334
Sweet Potatoes	4,303	.172	1.230
Wheat	7,082	1.422	.706
Barley	9,056	1.623	.790

ton of protein and carbohydrates. A means of combining the two values must be established that minimizes assumptions regarding the relative values of food and energy.

One approach to standardize the index would be to reduce or expand the range of a variable in such a way as to render comparisons among variables meaningful. One alternative is to divide all values by the maximum of the variable. For example, the standardized food value for brown rice would be 1.0 (the maximum) and 0.086 (the minimum) for cabbage (see column 1, Table 61). This implies that the maximum of either food or energy is equally valued in the computation of the index. The highest score for this index is rye, at 1.995 out of a possible 2.000.

Another form of standardization is to divide all the values of a variable by the variable's standard deviation. This has somewhat more statistical meaning, since 95 percent of all values of a normal distribution will fall within a distance of two standard deviations from the mean. For example, the total food value (given as carbohydrates plus protein) has a mean of 0.507, a range of 0.793, and a standard deviation of 0.362. The full range, that is 100 percent of the values, of the total food value falls within a distance of $0.793/0.362$, or 2.433 standard deviations of the mean. By dividing by the standard deviation instead of the maximum value, the highest for the food value is 2.663 for brown rice, and a minimum of 0.230 for cabbage. Standardizing in this manner indicates the extent of dispersion of the values rather than their fraction of the maximum value. The food and energy values shown in Tables 63 to 66 compare methods of standardization.

Standardizing by the highest value implies that the maximum of both energy and food values are equally valuable. If another crop is added to the list that has a higher food or energy value than was previously used, all the indices would require recalculation. However, in Tables 65 and 66 there are fewer changes in rank order in indices that are standardized by the highest value than by the standard deviation. This indicates a lower sensitivity to value computation technique for those indices standardized by the highest value.

TABLE 63. FOOD/ENERGY VALUES STANDARDIZED BY STANDARD DEVIATION.

	Energy Value 1	Protein + Carbohydrates 2	P & C + Pessimistic Protein Synthesis 3	P & C + Optimistic Protein Synthesis 4	Percent Dietary Requirements Provided 5
Cabbage	.009	.230	.224	.224	.246
Corn	.399	.742	.733	.768	.505
Oats	2.306	2.245	2.245	2.456	1.639
Field Peas	.112	.454	.445	.454	.414
Potatoes	.063	.653	.637	.642	.355
Brown Rice	1.814	2.663	2.638	2.801	1.459
Rye	2.323	2.650	2.639	2.851	1.634
Soybeans	1.162	1.350	1.343	1.449	3.566
Sweet Potatoes	.239	.905	.887	.907	.370
Wheat	1.973	2.610	2.591	2.770	1.818
Barley	2.251	2.604	2.592	2.797	1.698

TABLE 64. FOOD/ENERGY VALUES STANDARDIZED BY MAXIMA

	Energy Value 1	Protein + Carbohydrates 2	P & C + Pessimistic Protein Synthesis 3	P & C + Optimistic Protein Synthesis 4	Percent Dietary Requirements Provided 5
Cabbage	.004	.086	.085	.079	.069
Corn	.172	.279	.278	.269	.142
Oats	.993	.843	.851	.861	.461
Field Peas	.048	.171	.169	.159	.116
Potatoes	.027	.245	.242	.225	.100
Brown Rice	.781	1.000	1.000	.983	.410
Rye	1.000	.995	1.000	1.000	.460
Soybeans	.500	.507	.509	.508	1.000
Sweet Potatoes	.103	.340	.336	.318	.104
Wheat	.849	.980	.982	.971	.511
Barley	.969	.978	.982	.981	.478

TABLE 65. FOOD/ENERGY INDICES, WITH VALUES STANDARDIZED BY STANDARD DEVIATION.

Crop	Protein + Carbohydrates + Energy 1	P & C + Pessimistic Protein Synthesis + Energy 2	P & C + Optimistic Protein Synthesis + Energy 3	Percent Dietary Requirements Provided Energy 4
Cabbage	.240	.234	.234	.255
Corn	1.141	1.131	1.167	.904
Oats	4.551	4.551	4.762	3.945
Field Peas	.566	.557	.566	.526
Potatoes	.717	.701	.705	.419
Brown Rice	4.477	4.452	4.616	3.274
Rye	4.973	4.962	5.173	3.952
Soybeans	2.512	2.506	2.611	4.728
Sweet Potatoes	1.144	1.126	1.146	.609
Wheat	4.583	4.564	4.743	3.791
Barley	4.855	4.843	5.048	3.949

TABLE 66. FOOD/ENERGY INDICES, WITH VALUES STANDARDIZED BY HIGHEST VALUE.

Crop	Protein + Carbohydrates + Energy 1	P & C + Pessimistic Protein Synthesis + Energy 2	P & C + Optimistic Protein Synthesis + Energy 3	Percent Dietary Requirements Provided Energy 4
Cabbage	.090	.089	.083	.073
Corn	.450	.449	.441	.314
Oats	1.836	1.843	1.854	1.454
Field Peas	.219	.217	.207	.165
Potatoes	.273	.269	.252	.127
Brown Rice	1.781	1.781	1.763	1.191
Rye	1.995	2.000	2.000	1.459
Soybeans	1.007	1.009	1.009	1.503
Sweet Potatoes	.443	.439	.421	.207
Wheat	1.830	1.831	1.821	1.361
Barley	1.947	1.951	1.950	1.446

Figure 83. Energy - Food index (EFI), energy values and food values.

Although sensitivity may be one criterion for choosing a standardization technique, one that provides a somewhat more statistically meaningful value may be desired. Dividing by the standard deviation will indicate the extent of the spread of the value. Assuming that there is a normal distribution of values, one would expect 84 percent of the values to fall five standard deviations of the lowest value, 98 percent within six, and virtually 100 percent within seven. The largest value standardized in this way is 3.566, indicating that none of the values have very large spreads. Here, the maximum values have meaning, whereas the highest value obtained when standardizing by the highest value is by definition equal to 1.0. However, dividing by the standard deviation gives an index that is more sensitive to the method used to compute the index in the first place (see Table 65). Also, the introduction of a new crop to the index list will require the recalculation of all indices, regardless of the size of the energy and food values. This is due to the probable change in the standard deviation.

SUMMARY

A general procedure was developed to emphasize the concept that value of fixed solar energy in crops needs to be thought of in terms of both energy and human food. Such an index would assist in the development of a rational plan to utilize land and crops to optimize the value of the products.

Five ways of computing a food value for a crop are discussed by:

1. Adding the protein and carbohydrate contents;
2. including the amount of protein synthesized during anaerobic fermentation;
3. reflecting the extent to which a crop fulfilled the minimum dietary requirements for protein and carbohydrates;
4. reflecting the current prices of protein and carbohydrates; and
5. the percent of digestible nutrients in the food part of the plant.

As a first attempt, only the first three methods were examined. The energy value of a crop was computed on the basis of the quantity of volatile

solids available for anaerobic fermentation. Other indices may be computed using other energy recovery systems such as incineration or pyrolysis, or may reflect the current price of energy.

Regardless of the method used to combine the two values, the assumptions made will affect the rank order of index values. The methods of computation and standardization must be chosen subjectively in anticipation of future needs.

There are other factors that might influence a crop's value, in both energy and food such as:

1. The existence of a market for the food products and the adaptability of the farm or community to the use of methane gas;
2. The availability of capital, machinery, and labor needed for a change-over to a new crop, including storage and processing;
3. The sensitivity of a new crop to changes in the environment, such as hours of sunlight (or the number of growing degree days), temperature, humidity of the air, and the moisture, nutrient content, texture, and alkalinity of the soil;
4. Requirements for erosion control due to the removal of crop remains.

Although these factors are not included here, they may prove significant in future research in this area.

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CHAPTER XI

NUTRITIVE ANALYSIS OF ANIMAL RESIDUES BEFORE AND AFTER ANAEROBIC FERMENTATION

INTRODUCTION

The organic matter to be used in anaerobic fermentation must compete with such alternative uses as refeeding of the raw waste. Also, the undigested residue in the effluent from the fermentors may have value as a feed material. Relatively little information is available in these two areas even though the potential value from recovery of the residue both before and after fermentation may be significant (see Figure 2 in the summary). This portion of the study was initiated to provide information on the basic nutritive value of the dairy animal residues before and after fermentation. Although it was recognized that the nitrogen and protein content of the residues is of major concern, the premature termination of this study prevented development of this information. Only digestible organic matter analyses are presented here.

Fiber digestion is predominantly a phenomenon of ruminants and consequently, the nutritive analyses were based on the methods used to evaluate forages (Goering and Van Soest, 1970). Samples of the influents and selected samples of the effluents from the various fermentors were analyzed for total ash, neutral-detergent fiber (plant cell wall), lignin (KMnO_4), cellulose, hemicelluloses, true digestibility of the dry matter and digestibility of the cell wall.

Due to the early termination of this portion of the study, complete input-output balances could not be completed. Only selected representatives of daily effluent samples were analyzed. No samples were run on the refractory organics at the end of the batch tests. However, since these initial trials were conducted to test equipment and identify optimum conditions for gas

production, it is the analysis of the influents that was of prime importance in order to be able to evaluate various substrates as potential sources of energy.

PRELIMINARY RESULTS AND DISCUSSION

Thermophilic Reactors

Samples from 3 trials conducted on thermophilic reactors were analyzed. The reactors were run at 60°C and fed influents at total solids concentrations of 4, 6, 8 and 10 percent as discussed in Chapter 2. The daily balances are given in Table 67. Of the components of the cell wall-lignin, cellulose and hemicelluloses, there was a greater disappearance of the hemicelluloses than the cellulose fraction as occurs in ruminants. There was incomplete recovery of the lignin and ash indicating some sedimentation occurring in the reactors, or a possible sampling error.

In the 3 influents tested, their potential for gas production could be related to the fiber content and its composition. The highest rate of gas production was achieved in the influent containing the lowest level of cell wall with the lowest lignin to acid-detergent fiber ratio (L/ADR). This sample also had the highest content of cell solubles (microbial matter plus any starch by-pass). Unfortunately, the composition of the cattle feed which resulted in the influents is unknown. The influent which had the highest gas production and lowest cell wall content and L/ADR ratio also had the lowest disappearance of cell wall both in the reactors and in vitro. This would indicate that it came from dairy cattle fed a high concentrate ration since there was quite a high concentration of starch in the feces. Concentrate fiber is not as digestible as that in roughages. Thus of the 3 in vitro measurements the organic matter digestibility estimate is the only one related to the gas production rate despite the good relationship between cell wall disappearance in the reactors and cell wall digestibility in vitro.

TABLE 67. COMPARISON OF NUTRIENT CHARACTERISTICS OF DAIRY MANURE BEFORE AND AFTER THERMOPHILIC ANAEROBIC FERMENTATION (60°C), (ALL VALUES EXPRESSED AS PERCENT OF TOTAL DRY WEIGHT EXCEPT AS NOTED).

48 hr IVR										
	Total Ash	Cell Wall	Lignin (KMnO ₄)	Cellulose	Hemi-celluloses	Cell Solubles	L _k /ADR	% TDDM	% TDOM	% DCW
Influent 3/77	11.24	50.97	9.18	23.77	18.02	37.79	27.86	44.32	55.78	15.01
% Disappearance of:										
	Dry Matter	Organic Matter	Cell Wall	Lignin	Cellulose	Hemi-celluloses	Total Ash	Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I
Reactor 1 4%TS	9.25	8.68	2.45	-12.69	3.94	8.19	13.78	17.08	600	.86
2 6%TS	11.67	11.15	10.49	- 6.62	7.50	23.14	15.76	12.04	900	1.57
3 8%TS	19.25	20.61	20.93	.25	23.50	28.08	8.47	20.19	1200	4.09
4 10%TS	21.20	22.06	19.25	1.80	19.67	27.58	14.40	25.85	1500	5.37
Mean	16.89		15.45	- 2.52	15.91	24.00	12.91	20.02		3.55
x̄ R1 4%TS	18.08	19.06	19.51	- .70	19.85	30.40	10.09	17.45		2.00
R2 6%TS	17.22	18.47	20.39	- .60	19.29	33.89	8.11	14.83		2.34
R3 8%TS	19.92	21.91	23.18	.52	24.73	34.56	4.83	19.65		3.14
R4 10%TS	23.23	24.62	24.28	4.87	25.96	33.84	12.69	24.72		3.60
Mean	20.26		22.45	1.66	23.31	33.56	8.80	20.13		2.97

TABLE 67. (CONTINUED).

48 hr IVR											
								%	%	%	
								TDDM	TDOM	DCW	
								L _k /ADR			
								Total Ash	Cell Wall	Lignin (KMnO ₄)	
								Cellulose	Hemi-celluloses	Cell Solubles	
Influent 11/76								39.20	42.06	51.61	21.72
% Disappearance of:											
								Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I	
								Total Ash	Hemi-celluloses	Cellulose	Lignin
								Dry Matter	Organic Matter	Cell Wall	
Reactor 1	4%TS	27.50	28.19	30.18	14.60	29.88	42.14	23.44	24.29	600	2.03
2	6%TS	21.00	23.00	26.92	7.64	26.61	41.65	9.17	15.35	900	2.02
3	8%TS	18.38	20.40	22.00	1.85	22.70	39.01	6.44	17.27	1200	1.81
4	10%TS	21.50	22.86	24.43	2.77	28.56	35.87	13.48	19.79	1500	1.90
Mean		21.36		25.09	4.18	26.66	38.90	11.10	18.76		1.92

48 hr IVR											
								%	%	%	
								TDDM	TDOM	DCW	
								L _k /ADR			
								Total Ash	Cell Wall	Lignin (KMnO ₄)	
								Cellulose	Hemi-celluloses	Cell Solubles	
Influent 1/77								30.15	45.99	55.23	23.04
% Disappearance:											
								Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I	
								Total Ash	Hemi-celluloses	Cellulose	Lignin
								Dry Matter	Organic Matter	Cell Wall	
Reactor 1	4%TS	17.50	20.30	25.89	4.01	25.73	40.86	6.95	11.09	600	3.10
2	6%TS	19.00	21.25	23.76	2.81	23.77	36.87	.60	17.11	900	3.43
3	8%TS	22.13	24.71	26.60	3.16	28.00	36.58	.43	21.60	1200	3.53
4	10%TS	27.00	28.93	29.17	10.05	29.65	38.06	10.18	28.53	1500	3.54
Mean		22.54		26.81	3.32	27.36	37.78	2.39	21.61		3.45

Mesophilic Reactors

The results of the balances with samples from the mesophilic reactors are shown in Table 68. As with the thermophilic units, there was an extremely variable lignin recovery, ranging from 64 to 110 percent, which makes interpretation of the results difficult. The influent sample on 1/77 was fed to both the thermophilic reactors and mesophilic reactors. The mesophilic unit showed about 44 percent greater digestion of the cell wall and a lower disappearance of microbial matter indicating a greater bacterial growth than in the thermophilic units. It also exhibited a higher rate of gas production.

The two temperatures, 22°C and 35°C, were compared using three influents, A, B and C. Gas production was substantially higher at 35°C than at 22°C. When input-output data were corrected for complete lignin recovery, cell wall digestion at 35°C was about twice that at 22°C.

A large scale digester experiment (190 l reactor at 60°C) had the highest rate of gas production of all the experiments. When the input-output data were corrected for 100 percent lignin recovery, it showed 36 percent digestion of the cell wall.

Not enough data are available from the mesophilic reactor experiments to be able to evaluate the various influents by their chemical composition as potential energy sources. All did have high true digestibility of organic matter estimates and a high cell solubles content.

Cellulose and Hemicellulose Disappearance in the Reactors

The cell wall components that are degraded in the reactors are the cellulose and hemicelluloses. The contribution of these two components to the fermentation will depend upon their concentration in the cell wall, their association with the lignin and their digestibility. It may also be affected by ambient temperature which could result in the selection of a microbial population favoring one of these components over the other. Table 69 summarizes the input-output data averaged according to reactor temperature.

TABLE 68. COMPARISON OF NUTRIENT CHARACTERISTICS OF DAIRY MANURE BEFORE AND AFTER MESOPHILIC (35°C) ANAEROBIC FERMENTATION (ALL VALUES EXPRESSED AS PERCENT OF TOTAL DRY WEIGHT EXCEPT AS NOTED).

								48 hr IVR		
	Total Ash	Cell Wall	Lignin (KMnO ₄)	Cellulose	Hemi-celluloses	Cell Solubles	L _k /ADR	% TDDM	% TDOM	% DCW
Influent 1/77	10.23	54.29	10.30	23.94	20.05	35.48	30.08	47.74	57.57	24.12
% Disappearance of:										
	Dry Matter	Organic Matter	Cell Wall	Lignin	Cellulose	Hemi-celluloses	Total Ash	Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I
35°C, 12-day HRT	22.13	25.97	38.51	10.92	36.61	55.24	-11.61	6.62	80	5.13
								48 hr IVR		
	Total Ash	Cell Wall	Lignin (KMnO ₄)	Cellulose	Hemi-celluloses	Cell Solubles	L _k /ADR	% TDDM	% TDOM	% DCW
Influent A	13.30	49.81	8.60	24.44	16.77	36.89	26.03	41.74	54.62	12.60
% Disappearance of:										
	Dry Matter	Organic Matter	Cell Wall	Lignin	Cellulose	Hemi-celluloses	Total Ash	Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I
35°C, 10-day HRT	25.38	26.84	37.39	-4.12	42.67	51.06	15.82	12.60	96	16.44

TABLE 63. (CONTINUED).

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48 hr IVR										
	Total Ash	Cell Wall	Lignin (KMnO ₄)	Cellulose	Hemi-celluloses	Cell Solubles	L _k /ADR	% TDDM	% TDOM	% DCW
Influent B	10.35	51.29	10.72	22.51	18.06	38.36	32.26	44.48	54.85	13.80
% Disappearance of:										
	Dry Matter	Organic Matter	Cell Wall	Lignin	Cellulose	Hemi-celluloses	Total Ash	Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I
22°C, 12-day HRT	37.0C	38.00	31.64	35.57	34.52	25.72	28.37	46.5C	96	3.36
Assuming complete lignin recovery			-6.09	0.00	-1.62	-15.28				
48 hr IVR										
	Total Ash	Cell Wall	Lignin (KMnO ₄)	Cellulose	Hemi-celluloses	Cell Solubles	L _k /ADR	% TDDM	% TDOM	% DCW
Influent C	11.43	50.66	8.16	24.78	18.32	37.91	25.33	44.81	56.50	15.94
% Disappearance of:										
	Dry Matter	Organic Matter	Cell Wall	Lignin	Cellulose	Hemi-celluloses	Total Ash	Microbial Matter	DM _I /Day (g)	Gas Production 1/100gDM _I
CMF, 35°C, 12-day HRT	33.25	36.82	40.28	8.59	41.94	52.19	5.68	32.19	20	23.10
CMF, 22°C, 12-day HRT	20.25	22.21	20.50	-10.28	21.61	32.76	5.04	24.48	134	1.47

TABLE 68. (CONTINUED).

								48 hr IVR		
	Total Ash	Cell Wall	Lignin (KMnO ₄)	Cellulose	Hemi-celluloses	Cell Solubles	L _k /ADR	% TDDM	% TDOM	% DCW
Influent D	15.15	50.39	9.89	21.77	18.73	34.46	31.24	42.67	57.86	20.01
% Disappearance of:										
	Dry Matter	Organic Matter	Cell Wall	Lignin	Cellulose	Hemi-celluloses	Total Ash	Microbial Matter	DM _I /Day (kg)	Gas Production 1/100gDM _I
Harford 35°C 30-day HRT	40.00	46.05	54.96	29.33	59.52	63.16	6.09	33.02	15.2	33.32
Assuming complete lignin recovery			36.29	0.00	42.60	47.72				

Cell Solubles = 100 - (% cell wall + % total ash)

L_k/ADR = Ratio of lignin to acid-detergent fiber (lignin + cellulose)

48 hr IVR = 48 hour in vitro rumen fermentation

TDDM = True digestibility of dry matter

TDOM = True digestibility of organic matter

DCW = Digestibility of cell wall

TABLE 69. AVERAGE DISAPPEARANCE OF CELLULOSE AND HEMICELLULOSES DURING ANAEROBIC FERMENTATION
(ALL VALUES AS PERCENT DRY WEIGHT).

	Reactor Temperature											
	22°C				35°C				60°C			
	Dry Matter	Cell Wall	Cellulose	Hemicelluloses	Dry Matter	Cell Wall	Cellulose	Hemicelluloses	Dry Matter	Cell Wall	Cellulose	Hemicelluloses
Average Input	100	50.66	24.18	18.32	100	51.78	23.30	19.03	100	54.47	23.47	19.60
Digestibility		13.98	14.47	19.24		33.89	35.79	48.40		22.21	23.22	33.93
Amount Digested		7.08	3.50	3.58		17.55	8.34	9.21		12.10	5.45	6.65
Of the Cell Wall Digested:												
Cellulose Contributed		49.43%				47.52%				45.04%		
Hemicelluloses Contributed		50.57%				52.48%				54.96%		

In all cases the digestibility of hemicelluloses was greater than that of the cellulose. Although the influents contained more cellulose than hemicelluloses, the hemicelluloses made the greater contribution to the fermentative process. There appeared to be a trend for the contribution of the hemicelluloses to increase as the reactor temperature increased and that from the cellulose to decrease.

Chemical Analysis of Effluents

The results of the chemical analysis of the effluents from the reactors are summarized in Table 70. Even if the engineering problems of removing the water (some 95 percent) were overcome, the remaining dry matter is of low availability to ruminants. The plant cell wall remaining in the effluents is highly lignified and would not be digested to any great degree in the rumen. The organic matter that could be digested would be predominantly bacterial cell walls. The effluent dry matter, apart from its manurial value, might be useful as a diluent in high concentrate ruminant feeds to stimulate some rumination. The low quality of the effluent dry matter is indicative that the fermentors were working near maximum extraction of the available energy in the influents.

CONCLUSIONS

In the thermophilic reactor experiments, the chemical analysis of the influent was a guide to estimating the potential for fermentation. The cell wall content and the lignin-acid detergent fiber ratio seem to be the best parameters for evaluation. There was inadequate data to draw similar conclusions for the mesophilic reactors.

The input-output balances, although having variable lignin recoveries, showed that cell wall digestibility in the reactors and gas production were positively related. The 35°C reactors, a temperature approaching the ambient temperature for methanogenic rumen bacteria, yielded the highest rate of gas production and most complete digestion of the cell wall.

TABLE 70. CHEMICAL ANALYSIS OF THE ANAEROBIC FERMENTOR EFFLUENTS.

Effluent Date Sample	In Dry Matter				Availability*				48 hr IVR			
	DM %	Total Ash	Cell Wall	Lignin (KMnO_4)	Cellu- lose	Hemi- cellu- loses	L_k/ADF	True	Apparent	TDDM %	TDOM %	DCW %
60°C Effluents												
11/76	5.51	16.14	53.84	17.73	20.81	15.24	46.00	42.34	17.41	35.30	44.99	10.30
1/77	5.42	13.01	52.40	13.10	22.62	16.68	36.67	50.13	27.35	40.12	51.19	12.41
3/77	5.82	11.59	52.15	11.32	24.20	16.63	31.87	54.30	32.66	40.02	51.21	9.68
Average	5.53	13.58	52.80	14.05	22.54	16.18	38.40	48.79	25.64	38.48	49.13	10.80
35°C Effluents												
1/77	6.04	14.55	42.80	11.78	19.49	11.53	37.67	54.54	32.97	40.04	52.69	.53
C	5.34	16.18	45.33	11.19	21.03	13.11	34.73	52.23	30.02	36.50	50.38	1.03
A	5.97	15.01	41.79	12.01	18.78	11.00	39.01	54.04	32.33	40.28	54.13	1.34
D	4.80	23.64	37.32	11.64	14.64	11.54	44.29	46.85	23.16	36.08	46.37	2.06
Average	5.54	17.37	41.94	11.66	18.49	11.80	38.67	52.04	29.78	38.23	50.89	1.24
22°C Effluents												
C	6.38	13.61	50.49	11.28	23.76	15.45	32.19	53.36	31.64	35.71	51.68	7.93
B	5.04	11.78	55.65	10.96	23.39	21.30	31.91	51.95	29.67	34.34	45.70	5.92
Average	5.71	12.70	53.07	11.12	23.58	18.38	32.05	52.66	30.57	35.03	48.69	6.93

*From chemical analysis: True is "true digestibility of organic matter" and Apparent is "apparent digestibility."

Hemicelluloses were digested to a greater degree than celluloses although both contributed about equally to the fermentation. There was a trend to decreased digestion of cellulose as temperature increased.

The effluents from the reactors, apart from the bacterial debris, could be considered non-nutritive residues, and had low refeeding values as far as energy is concerned. No work was conducted on the protein content.

RECOMMENDATIONS

Complete input-output balances should be conducted on the fermentors. This should include a nitrogen fractionation of the crude protein.

If possible a wider range of fecal wastes should be tested. These should come from cattle fed known diets. Cattle digestibility studies on these diets would be of value.

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APPENDIX A

Conventional Controlled Fermentation Data

TABLE A1. STEADY STATE DATA FROM THERMOPHILIC
CONTROL DIGESTER OPERATED AS A 3 1/3
DAY HRT, 4% TS FEED

Date	Flow l/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/l	NH ₄ N mg/l	TKN mg/l	T.V.A. mg/l	Temp. °C	CST sec.
Average	4.5	4.0	38.92	7.0	2428	216	1228	-	-	1116.4
EFFLUENT										
2-23-77	4.5	3.20	36.09	6.0	-	350	-	-	61	-
2-25-77	4.5	3.28	36.53	6.3	-	322	-	-	60	-
3-02-77	4.5	3.22	36.64	6.7	-	308	-	-	61	-
3-04-77	4.5	3.25	36.82	6.5	-	273	-	-	60	-
3-08-77	4.5	3.39	37.61	6.2	-	273	-	-	60	-
3-10-77	4.5	3.40	38.30	6.1	-	224	-	-	61	-
3-15-77	4.5	3.53	38.68	6.0	-	238	-	-	61	-
3-22-77	4.5	3.60	38.06	5.9	1800	266	1204	3046	61	-
3-24-77	4.5	3.59	38.30	6.0	1800	308	1204	3206	60	-
3-28-77	4.5	3.70	37.86	6.1	1900	791	1204	1430	60	759.7
Average*	4.5	3.63	38.07	6.2	1833	335	1204	2561	61	759.7

Gas Production

Date	l/day	CH ₄ /CO ₂	Date	l/day	CH ₄ /CO ₂	Date	l/day	CH ₄ /CO ₂
3-17-77	5.27	-	3-22-77	5.32	-	3-27-77	6.20	-
3-18-77	5.53	-	3-23-77	4.99	-	3-28-77	5.31	-
3-19-77	5.09	-	3-24-77	3.96	-	3-29-77	5.49	-
3-20-77	5.33	-	3-25-77	5.11	-	Average	5.14	-
3-21-77	4.98	-	3-26-77	4.68	-			

e of last three data points

TABLE A2. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 5 DAY HRT, 4% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	T.V.A. mg/ℓ	Temp. °C	CST sec.
Average	3.0	4.0	88.71	7.1	2554	205	1184	-	-	1438.1
Effluent										
1-18-77	3.0	3.12	86.72	7.4	4300	350	1204	-	61	737.5
1-20-77	3.0	3.38	85.78	7.0	4000	364	1204	-	61	915.4
2-07-77	3.0	3.33	87.39	6.6	3200	266	880	631	60	1069.3
2-09-77	3.0	3.30	87.27	6.5	3100	280	1414	1000	62	-
2-10-77	3.0	3.35	87.93	6.5	3200	308	1232	1357	62	-
Average	3.0	3.30	86.92	6.8	3560	314	1209	996	61	907.4
Gas Production										
Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂		
1-14-77	18.50	-	1-23-77	20.25	-	2-08-77	17.57	-		
1-15-77	20.39	-	1-31-77	17.43	-	2-09-77	16.93	-		
1-16-77	23.20	-	2-01-77	18.95	-	2-10-77	15.63	50/50		
1-17-77	22.60	-	2-02-77	17.83	-	2-11-77	20.54	-		
1-18-77	22.60	51/49	2-03-77	17.82	51/49	2-12-77	15.64	-		
1-19-77	21.06	-	2-04-77	18.70	-	2-13-77	16.19	-		
1-20-77	19.69	51/49	2-05-77	13.06	-	2-14-77	16.68	51/49		
1-21-77	18.71	-	2-06-77	16.80	-	2-15-77	17.94	-		
1-22-77	19.61	-	2-07-77	18.30	-	Ave.	18.60	51/49		

TABLE A3. STEADY STATE FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 10 DAY HRT, 4% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp. °C	CST sec.
Average	1.5	4.0	88.93	7.7	796	539	1536	-	-	-
Effluent										
11-01-76	1.5	3.40	91.47	7.2	5000	504	1470	--	61	--
11-03-76	1.5	2.80	81.79	7.1	5000	497	--	--	58	--
11-05-76	1.5	3.33	85.59	7.1	5900	511	1442	1283	61	--
11-15-76	1.5	2.04	76.47	7.2	5600	651	1484	1494	60	--
11-18-76	1.5	3.75	86.40	7.2	--	--	--	2315	57	--
11-23-76	1.5	3.70	85.85	7.4	--	--	--	--	58	--
11-29-76	1.5	3.86	86.01	7.4	--	--	--	--	61	--
11-30-76	1.5	3.36	84.19	7.5	--	--	--	--	61	--
12-01-76	1.5	3.59	85.24	7.5	--	--	--	--	60	--
12-02-76	1.5	3.35	84.48	7.4	--	--	--	--	60	--
12-03-76	1.5	3.10	82.90	7.5	--	--	--	--	60	--
12-06-76	1.5	3.59	83.85	7.6	6000	707	1624	--	57	--
12-08-76	1.5	3.11	84.60	7.3	--	--	--	--	60	--
12-13-76	1.5	2.96	83.75	7.3	6300	--	--	--	59	--
Average	1.5	3.28	84.47	7.3	5633	574	1505	1697	60	--

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Gas Production								
Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂
11-01-76	13.97	--	11-07-76	7.42	--	11-13-76	11.28	--
11-02-76	11.35	--	11-08-76	9.09	--	11-14-76	11.75	--
11-03-76	--	--	11-09-76	9.55	--	11-15-76	12.13	--
11-04-76	11.42	--	11-10-76	11.31	--	11-16-76	12.31	--
11-05-76	11.87	--	11-11-76	11.15	--	11-17-76	12.31	61/39
11-06-76	15.39	--	11-12-76	11.53	--	11-18-76	--	--

Gas Production (Cont.)

Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂
11-19-76	13.11	--	12-02-76	12.47	--
11-20-76	9.23	--	12-03-76	11.19	--
11-21-76	14.32	--	12-04-76	11.81	--
11-22-76	10.76	--	12-05-76	12.79	--
11-23-76	10.85	--	12-06-76	15.49	61/39
11-24-76	10.87	--	12-07-76	11.46	--
11-25-76	11.76	--	12-08-76	11.62	--
11-26-76	12.51	--	12-09-76	--	--
11-27-76	13.99	--	12-10-76	11.17	--
11-28-76	14.22	--	12-11-76	15.78	--
11-29-76	15.82	--	12-12-76	12.94	--
11-30-76	14.98	--	12-13-76	10.45	61/39
			Average	12.13	61/39

TABLE A4. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 3 1/3 DAY HRT, 6% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	T.V.A. mg/ℓ	Temp. °C	CST sec
Average	4.5	6.0	88.92	7.0	3642	324	1842	--	--	2661.5
Effluent										
2-23-77	4.5	4.75	86.27	6.7	--	462	--	--	62	--
2-25-77	4.5	4.44	86.62	6.6	--	399	--	--	60	--
3-02-77	4.5	4.70	86.91	6.7	--	434	--	--	61	--
3-04-77	4.5	4.53	85.82	6.7	--	483	--	--	60	--
3-08-77	4.5	4.51	86.24	6.8	--	315	--	--	61	--
3-10-77	4.5	4.91	87.36	6.6	--	343	--	--	61	--
3-15-77	4.5	5.11	87.93	6.4	--	294	--	--	61	--
3-22-77	4.5	4.58	87.18	6.2	3600	315	1652	3397	59	--
3-24-77	4.5	5.45	87.98	6.1	3900	469	1778	3487	60	--
3-28-77	4.5	5.73	88.02	6.1	2900	420	1820	1415	61	1537.2
Average*	4.5	5.30	87.73	6.4	3467	393	1750	2766	61	1537.2

Gas Production

Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂
3-17-77	16.54	--	3-22-77	15.37	38/62	3-27-77	10.59	--
3-18-77	20.83	--	3-23-77	15.68	--	3-28-77	8.42	38/62
3-19-77	17.17	--	3-24-77	12.06	38/62	3-29-77	9.24	--
3-20-77	17.67	--	3-25-77	12.57	--	Average	14.10	38/62
3-21-77	16.39	--	3-26-77	10.81	--			

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TABLE A5. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 5 DAY HRT, 6% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Atk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp °C	CST sec
Average	3.0	6.0	88.71	7.1	3817	307	1775	--	--	2523.6
Effluent										
1-18-77	3.0	5.13	86.74	7.5	6200	420	1666	--	61	2282.9
1-20-77	3.0	4.67	85.51	7.2	6200	490	1568	--	61	1512.9
2-07-77	3.0	4.31	85.71	6.8	4400	357	1246	2443	61	2859.5
2-09-77	3.0	4.77	87.42	6.7	5000	336	1540	2247	62	--
2-10-77	3.0	4.90	87.85	6.7	5100	448	1694	1791	63	--
Average	3.0	4.76	86.66	7.0	5380	410	1543	2160	62	2218.4
Gas Production										
Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ CO ₂		
1-14-77	33.82	--	1-23-77	31.15	--	2-8-77	30.33	--		
1-15-77	30.60	--	1-31-77	27.12	--	2-9-77	28.39	--		
1-16-77	38.03	--	2-1-77	34.26	--	2-10-77	27.29	49/51		
1-17-77	36.04	--	2-2-77	30.07	--	2-11-77	28.53	--		
1-18-77	36.10	51/49	2-3-77	31.44	51/49	2-12-77	24.81	--		
1-19-77	33.75	--	2-4-77	32.68	--	2-13-77	29.31	--		
1-20-77	35.34	51/49	2-5-77	21.34	--	2-14-77	22.35	--		
1-21-77	33.80	--	2-6-77	26.14	--	2-15-77	--	--		
1-22-77	33.35	--	2-7-77	30.82	--	Average	30.88	51/49		

TABLE A6. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 10 DAY HRT, 6% TS FEED

Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Influent				Temp. °C	CST sec.
					Atk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ		
Average	1.5	6.0	88.93	7.7	1195	809	2304	--	--	--
Effluent										
11-1-76	1.5	5.06	85.77	7.4	7700	780	2212	--	59	--
11-3-76	1.5	5.26	84.79	7.3	7400	784	2184	--	59	--
11-5-76	1.5	4.49	84.63	7.3	7900	931	2128	--	58	--
11-18-76	1.5	4.89	85.48	7.5	--	--	--	--	57	--
11-23-76	1.5	4.91	84.32	7.5	--	--	--	--	58	--
11-29-76	1.5	4.92	84.15	7.5	--	--	--	--	59	--
11-30-76	1.5	4.95	83.96	7.6	--	--	--	--	59	--
12-1-76	1.5	5.00	84.42	7.5	--	--	--	--	60	--
12-2-76	1.5	4.88	84.23	7.4	--	--	--	--	60	--
12-3-76	1.5	4.91	83.50	7.5	--	--	--	--	59	--
12-6-76	1.5	4.95	83.93	7.5	8800	1078	2604	--	57	--
12-8-76	1.5	4.55	84.41	7.4	--	--	--	--	60	--
12-15-76	1.5	4.43	83.52	7.3	9800	--	--	2274	59	--
Average	1.5	4.84	84.32	7.4	8200	863	2245	1679	59	--

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Gas Production										
Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂	Date	ℓ/day	CH ₄ /CO ₂		
11-1-76	18.95	--	11-7-76	12.35	--	11-13-76	15.29	--		
11-2-76	20.38	--	11-8-76	14.44	--	11-14-76	16.04	--		
11-3-76	--	--	11-9-76	15.92	--	11-15-76	15.96	--		
11-4-76	18.76	--	11-10-76	16.78	--	11-16-77	19.50	--		
11-5-76	20.65	--	11-11-76	17.12	--	11-17-76	19.64	62.38		

Gas Production (Cont.)

Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂
11-19-76	21.30	--	12-2-76	17.71	--
11-20-76	16.23	--	12-3-76	16.98	--
11-21-76	16.52	--	12-4-76	17.97	--
11-22-76	17.54	--	12-5-76	19.03	--
11-23-76	18.26	--	12-6-76	19.10	62/38
11-24-76	17.39	--	12-7-76	13.12	--
11-25-76	18.18	--	12-8-76	18.80	--
11-26-76	19.16	--	12-9-76	--	--
11-27-76	19.69	--	12-10-76	16.71	--
11-28-76	20.31	--	12-11-76	22.12	--
11-29-76	20.54	--	12-12-76	16.75	--
11-30-76	18.78	--	12-13-76	12.78	62/38
12-1-76	17.39	--	Average	18.12	62/38

TABLE A7. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 20 DAY HRT, 6% TS FEED

Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp. °C	CST sec.
Influent										
Average	.750	6.0	88.3	6.7	2310	230	1600	--	--	1390.3
Effluent										
7-07-77	.750	4.4	84.1	7.0	2800	392	1680	180	59	--
7-11-77	.750	4.5	84.4	7.1	2700	364	1330	110	60	--
7-13-77	.750	4.6	82.6	7.0	2800	336	1806	100	60	465.8
Average	.750	4.5	84.4	7.0	2767	364	1605	130	60	465.8
Gas Production (ℓ/day)										
Date	ℓ/day	CH ₄ CO ₂								
6-25-77	15.29	--	Date	ℓ/day	CH ₄ CO ₂					
6-27-77	15.36	--	7-05-77	20.43	--					
6-29-77	15.45	--	7-07-77	17.41	54/46					
7-01-77	20.37	--	7-09-77	17.11	--					
7-03-77	14.52	--	7-11-77	15.12	54/46					
			7-13-77	21.36	56/44					
			Average	17.24	55/45					

TABLE A8. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 2 1/2 DAY HRT, 8% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp. °C	CST sec
Average	6.0	8.0	88.92	7.0	5360	432	2456	--	--	3758.2
Effluent										
2-23-77	6.0	6.16	93.43	6.9	--	476	--	--	60	--
2-25-77	6.0	5.70	86.21	6.9	--	448	--	--	60	--
3-02-77	6.0	6.22	86.35	7.0	--	560	--	--	60	--
3-04-77	6.0	6.02	85.98	7.0	--	532	--	--	59	--
3-08-77	6.0	6.02	86.64	6.9	--	420	--	--	60	--
3-10-77	6.0	5.94	87.36	6.6	--	364	--	--	61	--
3-15-77	6.0	6.57	88.21	6.6	--	476	--	--	60	--
3-28-77	6.0	6.63	86.04	6.8	4000	546	2520	1581	60	--
3-29-77	6.0	6.54	86.69	6.9	3900	588	2184	1929	61	--
3-30-77	6.0	6.22	85.59	7.0	4000	644	2464	974	60	2490.2
Average*	6.0	6.46	86.11	6.9	3967	505	2389	1495	60	2490.2
Gas Production										
Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂		
2-17-77	40.98	--	3-22-77	38.67	--	3-27-77	60.60	51/49		
3-18-77	42.32	--	3-23-77	42.56	--	3-28-77	64.38	52/48		
3-19-77	39.58	--	3-24-77	47.35	--	3-29-77	66.91	52/48		
3-21-77	37.16	--	3-26-77	58.58	--	Average	49.13	52/48		

*Ave. of last of 3 points.

TABLE A9. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 5 DAY HRT, 8% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp. °C	CST sec.
Average	3.0	8.0	88.71	7.1	5089	410	2367	--	--	5013.5
Effluent										
1-18-77	3.0	6.42	85.98	7.5	8200	756	2464	--	60	3132.9
1-20-77	3.0	6.02	85.45	7.3	8100	672	2436	--	60	5693.8
2-07-77	3.0	6.48	87.81	7.0	6300	476	2100	1654	59	4389.8
2-09-77	3.0	5.19	86.27	6.9	6300	574	2408	1530	62	--
2-10-77	3.0	5.28	87.44	7.0	6300	546	2912	2193	61	--
Average	3.0	5.28	86.59	7.1	7040	605	2464	1792	60	4405.5
Gas Production										
Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂		
1-14-77	39.27	--	1-23-77	42.93	--	2-8-77	39.12	--		
1-15-77	42.08	--	1-31-77	30.97	--	2-9-77	38.87	--		
1-16-77	52.07	--	2-01-77	44.19	--	2-10-77	36.84	50/50		
1-17-77	46.51	--	2-02-77	43.50	--	2-11-77	42.35	--		
1-18-77	49.04	53/47	2-03-77	42.79	53/47	2-12-77	42.66	--		
1-19-77	45.15	--	2-04-77	42.25	--	2-13-77	46.16	--		
1-20-77	43.81	53/47	2-05-77	25.67	--	2-14-77	46.79	53/47		
1-21-77	43.95	--	2-05-77	41.34	--	2-15-77	47.08	--		
1-22-77	37.31	--	2-07-77	41.38	--	Average	42.33	52/48		

TABLE A10. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 10 DAY HRT, 8% TS FEED

Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Influent				Temp. °C	Cst sec.
					Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ		
Average	1.5	8.0	88.93	7.7	1593	1078	3072	--	--	--
Effluent										
11-01-76	1.5	6.73	85.40	7.2	9800	1050	2744	--	59	--
11-03-76	1.5	7.04	85.09	7.1	7300	1162	2968	--	59	--
11-05-76	1.5	7.22	85.73	7.0	10300	1120	2996	2250	60	--
11-15-76	1.5	6.15	84.23	7.5	10400	1162	2660	2239	58	--
11-18-76	1.5	6.19	80.46	7.5	--	--	--	3380	56	--
11-23-76	1.5	6.56	84.60	7.5	--	--	--	--	58	--
11-29-76	1.5	6.58	84.80	7.6	--	--	--	--	59	--
11-30-76	1.5	6.51	83.88	7.7	--	--	--	--	59	--
12-01-76	1.5	6.76	83.58	7.5	--	--	--	--	59	--
12-02-76	1.5	6.46	84.06	7.5	--	--	--	--	59	--
12-03-76	1.5	6.46	84.06	7.5	--	--	--	--	59	--
12-06-76	1.5	6.50	84.19	7.4	11400	1456	3206	--	56	--
12-08-76	1.5	6.27	85.75	7.4	--	--	--	--	58	--
12-13-76	1.5	5.84	84.23	7.4	10350	1190	2915	2623	58	--
Average	1.5	6.52	84.23	7.4	10350	1190	2915	2623	58	--
Gas Production										
Date	ℓ/day	CH ₄ CO ₂		Date	ℓ/day	CH ₄ CO ₂		Date	ℓ/day	CH ₄ CO ₂
11-1-76	14.58	--		11-7-76	12.44	--		11-12-76	18.98	--
11-2-76	15.64	--		11-8-76	16.28	--		11-14-76	24.11	--
11-3-76	--	--		11-9-76	17.81	--		11-15-76	20.47	--

Gas Production (Cont.)

Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂
11-4-76	16.11	--	11-7-76	12.44	--	11-13-76	18.98	--
11-5-76	19.51	--	11-11-76	19.78	--	11-17-76	25.56	55/45
11-6-76	23.42	--	11-12-76	20.86	--	11-18-76	29.02	--
11-19-76	27.40	--	12-2-76	21.67	--			
11-20-76	23.17	--	12-3-76	21.69	--			
11-21-76	23.33	--	12-4-76	22.81	--			
11-22-76	19.99	--	12-5-76	21.31	--			
11-23-76	21.95	--	12-6-76	20.54	55/45			
11-24-76	21.64	--	12-7-76	25.06	--			
11-25-76	26.45	--	12-8-76	25.65	--			
11-26-76	21.13	--	12-9-76	--	--			
11-27-76	24.13	--	12-10-76	20.85	--			
11-28-76	24.75	--	12-11-76	30.72	--			
11-29-76	24.81	--	12-12-76	24.07	--			
11-30-76	22.81	--	12-13-76	17.16	55/45			
			Average	21.77	55/45			

TABLE A11. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 2 1/2 DAY HRT, 10% TS FEED
(7.5 ℓ REACTOR VOLUME)

Influent										
Date	Flow ℓ /day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ ℓ	NH ₄ N mg/ ℓ	TKN mg/ ℓ	TVA mg/ ℓ	Temp. °C	CST sec.
Average	3.0	10.0	88.92	7.0	6070	540	3070	--	--	5044.7
Effluent										
2-23-77	3.0	6.66	85.64	7.0	--	574	--	--	61	--
2-25-77	3.0	6.47	86.18	7.0	--	588	--	--	59	--
3-02-77	3.0	7.42	86.70	7.0	--	644	--	--	60	--
3-04-77	3.0	7.32	86.14	7.0	--	644	--	--	60	--
3-08-77	3.0	7.49	86.62	7.0	--	728	--	--	59	--
3-10-77	3.0	7.89	87.70	6.8	--	434	--	--	60	--
3-15-77	3.0	7.79	88.00	6.9	--	448	--	--	59	--
3-22-77	3.0	7.86	87.53	7.0	4000	476	2492	1916	59	--
3-24-77	3.0	8.27	87.30	7.0	4200	476	2828	2303	58	--
3-28-77	3.0	7.50	86.48	7.6	4300	560	2408	1076	59	3117.4
Average*	3.0	7.88	87.10	7.0	4.67	557	2576	1765	59	3117.4

Gas Production								
Date	ℓ /day	CH ₄ CO ₂	Date	ℓ /day	CH ₄ CO ₂	Date	ℓ /day	CH ₄ CO ₂
3-17-77	36.05	--	3-22-77	40.27	52/48	3-27-77	43.88	--
3-18-77	37.77	--	3-23-77	46.22	--	3-28-77	46.74	52/48
3-19-77	23.96	--	3-24-77	38.12	53/47	3-29-77	46.33	--
3-20-77	41.43	--	3-25-77	44.66	--	Average	40.30	52/48
3-21-77	40.57	--	3-26-77	37.90	--			

*Average of Last 3 Points

TABLE A12 STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 5 DAY HRT, 10% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp. °C	CST sec.
Average	3.0	10.0	88.71	7.7	6361	512	2959	--	--	8078.5
Effluent										
1-18-77	3.0	7.31	86.18	7.6	9800	910	2912	--	59	4806.0
1-20-77	3.0	7.28	85.89	7.5	9000	729	2576	--	60	4563.1
2-07-77	3.0	7.24	87.57	7.0	7500	490	2800	2518	60	4555.2
2-09-77	3.0	7.19	86.72	7.0	7600	560	2688	1305	61	--
2-10-77	3.0	7.44	87.46	7.0	7100	686	2996	2265	61	--
Average	3.0	7.28	86.76	7.2	8200	675	2794	2029	60	4641.4

Gas Production										
Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂		
1-14-77	50.13	--	1-23-77	47.95	--	2-8-77	47.82	--		
1-15-77	51.53	--	1-31-77	50.87	--	2-9-77	45.11	--		
1-16-77	52.39	--	2-1-77	55.18	--	2-10-77	46.29	50/50		
1-17-77	58.64	--	2-2-77	51.85	--	2-11-77	52.08	--		
1-18-77	62.10	56/48	2-3-77	52.10	52/48	2-12-77	56.52	--		
1-19-77	59.84	--	2-4-77	55.04	--	2-13-77	59.99	--		
1-20-77	58.07	52/48	2-5-77	40.72	--	2-14-77	60.50	53/47		
1-21-77	53.23	--	2-6-77	48.84	--	2-15-77	63.44	--		
							Average	53.08	52/48	

TABLE A13. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 10 DAY HRT, 10% TS FEED

Influent										
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	TVA mg/ℓ	Temp. °C	CST sec.
Average	1.5	10.0	88.93	7.7	1991	1348	3840	--	--	--
Effluent										
11-01-76	1.5	9.02	84.70	7.7	10800	1302	3790	--	60	--
11-03-76	1.5	8.00	83.88	7.4	13100	1498	3332	--	59	--
11-05-76	1.5	8.48	85.26	7.3	13300	1358	3668	4941	60	--
11-15-76	1.5	7.79	83.44	7.5	12900	1456	3528	--	58	--
11-18-76	1.5	7.91	84.20	7.5	--	--	--	4232	57	--
11-23-76	1.5	8.31	84.33	7.6	--	--	--	--	58	--
11-29-76	1.5	7.76	84.89	7.6	--	--	--	--	58	--
11-30-76	1.5	7.72	83.56	7.7	--	--	--	--	59	--
12-01-76	1.5	7.99	83.85	7.7	--	--	--	--	60	--
12-02-76	1.5	7.68	87.46	7.6	--	--	--	--	59	--
12-03-76	1.5	7.60	83.68	7.6	--	--	--	--	59	--
12-06-76	1.5	7.78	84.24	7.5	13980	1757	4060	--	56	--
12-08-76	1.5	7.68	85.81	7.4	--	--	--	--	59	--
12-13-76	1.5	7.47	82.73	7.5	17000	--	--	--	58	--
Average	1.5	7.94	84.13	7.5	13500	1474	3674	4586	59	--
Gas Production										
Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂		
11-01-76	31.42	--	11-07-76	17.74	--	11-13-76	26.31	--		
11-02-76	32.39	--	11-08-76	22.74	--	11-14-76	26.85	--		
11-03-76	--	--	11-09-76	24.00	--	11-15-76	27.62	--		
11-04-76	29.63	--	11-10-76	26.45	--	11-16-76	29.40	--		

Gas Production (Cont.)

Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂
11-05-76	29.87	--	11-11-76	26.49	--	11-17-76	31.54	--
11-06-76	34.67	--	11-12-76	28.01	--	11-18-76	30.88	--
11-19-76	31.11	--	12-02-76	28.31	--			
11-20-76	33.42	--	12-03-76	26.93	--			
11-21-76	25.37	--	12-04-76	29.12	--			
11-22-76	28.71	--	12-05-76	27.44	--			
11-23-76	26.62	--	12-06-76	26.73	58/42			
11-24-76	29.12	--	12-07-76	30.96	--			
11-25-76	31.17	--	12-08-76	31.96	--			
11-26-76	30.15	--	12-09-76	--	--			
11-27-76	29.71	--	12-10-76	21.78	--			
11-28-76	30.56	--	12-11-76	32.09	--			
11-29-76	29.77	--	12-12-76	29.56	--			
11-30-76	30.51	--	12-13-76	23.27	58/42			
			Average	28.49	58/42			

TABLE A14. STEADY STATE DATA FOR A THERMOPHILIC CONTROL
DIGESTER OPERATED AT A 30 DAY HRT, 10% TS FEED
(7.5 L REACTOR VOLUME)

Date	Flow L/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/L	NH ₄ N mg/L	TKN mg/L	TVA mg/L	Temp. °C	CST sec.
Influent										
Average	.500	10.0	88.8	6.7	231	23	160	--	--	4341.0
Effluent										
7-07-77	.500	6.05	94.9	7.2	3900	504	2772	160	58	--
7-11-77	.500	6.91	83.4	7.1	3900	616	2968	160	58	--
7-13-77	.500	7.47	84.2	7.2	4000	500	3136	170	59	1959.5
Average	.500	6.30	87.5	7.2	3933	560	2959	163	58	1959.5
Gas Production										
Date	L/day	CH ₄ CO ₂								
6-25-77	19.47	--	Date	L/day	CH ₄ CO ₂					
6-27-77	--	--	7-05-77	18.88	--					
6-29-77	16.84	--	7-07-77	17.60	57/43					
7-01-77	20.59	--	7-09-77	20.20	--					
7-03-77	12.65	--	7-11-77	19.28	46/54					
			7-13-77	20.15	59/41					
			Average	18.40	54/46					

TABLE A15. PARAMETER DATA FROM A THERMOPHILIC DIGESTER
OPERATED AT A 10 DAY HRT, 4% TS FEED DURING
A PERIOD OF SUSPECTED AMMONIA-INDUCED
INHIBITION.

Date	Gas Production (l/day)	pH	Ammonia Conc. mg/l, as N	Free Ammonia Conc. mg/l, as NH ₃
1-20-77	19.69	7.0	364	22.3
1-21	18.71	7.0	-	-
1-22	19.61	7.0	-	-
1-23	20.25	7.1	-	-
1-24	19.37	7.2	-	-
1-25	18.62	7.2	910	86.0
1-26	18.86	7.0	-	-
1-27	24.71	7.1	-	-
1-28	18.23	7.0	574	54.2
1-29	16.19	7.0	-	-
1-30	14.70	6.9	-	-
1-31	17.43	6.9	378	18.6
2/1	18.95	6.8	-	-
2/2	17.83	6.8	-	-
2/3	17.82	6.7	-	-
2/4	18.70	6.7	308	9.7

Feed Data Digester #1				
Batch 23	-	7.3	942.4	287.7
Average	-	7.1	204.8	15.6

TABLE A16. PARAMETER DATA FROM A THERMOPHILIC DIGESTER OPERATED AT A 10 DAY HRT, 6% TS FEED DURING A PERIOD OF SUSPECTED AMMONIA-INDUCED INHIBITION.

Date	Gas Production (l/day)	pH	Ammonia Conc. mg/l, as N	Free Ammonia Conc. mg/l, as NH ₃
1-20-77	35.34	7.2	490	46.3
1-21	33.80	7.1	-	-
1-22	33.35	7.2	-	-
1-23	31.13	7.2	-	-
1-24	23.94	7.2	-	-
1-25	23.55	7.3	1309	152.6
1-26	23.39	7.0	-	-
1-27	18.34	7.1	-	-
1-28	28.33	7.1	742	56.6
1-29	28.51	7.1	-	-
1-30	24.50	7.0	-	-
1-31	27.12	7.1	504	38.4
2-1	34.26	7.0	-	-
2-2	30.07	7.0	-	-
2-3	31.44	6.9	-	-
2-4	32.68	6.9	434	21.4

Feed Data Digester #2				
Batch 23	-	7.8	1413.6	431.6
Average	-	7.1	307.2	23.4

TABLE A17. PARAMETER DATA FROM A THERMOPHILIC DIGESTER
OPERATED AT A 10 DAY HART, 8% TS FEED DURING
A PERIOD OF SUSPECTED AMMONIA-INDUCED
INHIBITION.

Date	Gas Production (l/day)	pH	Ammonia Conc. mg/l, as N	Free Ammonia Conc. mg/l, as NH ₃
1-20-77	43.81	7.3	672	78.4
1-21	43.95	7.2	-	-
1-22	37.31	7.3	-	-
1-23	42.93	7.3	-	-
1-24	31.75	7.4	-	-
1-25	30.38	7.4	1904	272.7
1-26	28.77	7.1	-	-
1-27	28.92	7.2	-	-
1-28	32.97	7.1	1134	86.5
1-29	32.03	7.2	-	-
1-30	25.96	7.0	-	-
1-31	44.97	7.0	700	43.0
2-1	44.19	7.1	-	-
2-2	43.50	7.1	-	-
2-3	42.79	7.1	-	-
2-4	42.25	7.0	602	37.0

Feed Data Digester #1				
Batch 23	-	7.8	1884.8	575.5
Average	-	7.1	409.6	31.2

TABLE A18. PARAMETER DATA FROM A THERMOPHILIC DIGESTER
OPERATED AT A 10 DAY HRT, 10% FEED DURING A
PERIOD OF SUSPECTED AMMONIA-INDUCED INHIBITION.

Date	Gas Production (l/day)	pH	Ammonia Conc. mg/l, as N	Free Ammonia Conc. mg/l, as NH ₃
1-20-77	58.07	7.5	728	127.4
1-21	53.23	7.2	-	-
1-22	50.10	7.3	-	-
1-23	47.95	7.4	-	-
1-24	36.84	7.4	-	-
1-25	35.99	7.4	2128	304.8
1-26	31.32	7.4	-	-
1-27	33.10	7.6	-	-
1-28	39.06	7.2	1372	129.6
1-29	45.22	7.3	-	-
1-30	34.90	7.1	-	-
1-31	50.87	7.2	1022	96.6
2-1	55.18	7.1	-	-
2-2	51.98	7.1	-	-
2-3	52.10	7.1	-	-
2-4	55.04	7.1	742	56.6

Feed Data Digester #4				
Batch 23	-	7.8	2356	719.4
Average	-	7.1	512	39.1

TABLE A19. SUMMARY OF DATA FROM AMMONIA INHIBITION CONDITION

Parameter	Unit	Digester 1	Digester 2	Digester 3	Digester 4
Steady State Gas Production for the 10-day HRT condition	ℓ/day	18.6	30.88	42.33	53.08
Lowest Gas Production Recorded during Transition	ℓ/day	14.7	18.34	24.5	31.32
% Decrease	%	21.0	40.6	42.1	41.0
Steady State Ammonia Conc. for the 10 day HRT condition	mg/ℓ as N	314.0	410.0	605.0	675.0
Highest Ammonia Conc. Recorded during transition	mg/ℓ as N	910.0	1309.0	1904.0	2128.0
% Increase	%	290.0	319.0	315.0	315.0
Steady State Free Ammonia Conc. for the 10-day HRT condition	mg/ℓ as NH ₃	12.4	25.2	46.1	63.8
Highest Free Ammonia Conc. Recorded during transition	mg/ℓ as NH ₃	130.3	229.0	404.3	451.0
% Increase	%	1050.0	909.0	877.0	708.0

APPENDIX B
Plug Flow Simulation Data

TABLE B1

Summary of Start-Up Problems in UHD Reactors.

1. 30 Day HRT reactor--This unit was begun with 20 liters of raw manure diluted to 8% solids; no digested seed was used. After several days, feeding was initiated with raw manure at a rate equivalent to a 120-day HRT. A mild pH drop was observed, from 6.4 to 5.9, and feeding was interrupted for several days since it was felt that volatile acids were accumulating in the system. As gas production increased steadily to 0.3 l gas/l reactor, feeding was again started at a 120 day HRT. Gas production dropped sharply to 0.1 l gas/l reactor, and the feed was increased to a 60 day HRT to prevent substrate limiting conditions; pH had remained steady but low. However, gas production remained low and the pH plummeted to 5.3. A feeding schedule was designed to increase the population of methane-forming organisms in the system. Ten days of feeding a combination of raw manure and effluent from a healthier reactor brought the pH back to 6.0, but gas production was still a low 0.125 l gas/l reactor. At this point, it was decided that a massive infusion of healthy methane formers was necessary. Therefore, at a rate of HRT = 30 days, the unit was fed a complete diet of effluent from the healthy unit for 11 days. This resulted in a stabilized pH of 6.9, and steadily increasing gas production. Following one day of seeding with digested sewage sludge, the unit was hereafter fed only raw manure. The pH remained in the 7.0 - 7.3 range, while gas production increased daily.

2. 20 day HRT reactor--The start-up of this unit was comparable to the 30 day HRT reactor. It was started without a seed, and feeding was started on a schedule of an 80 day HRT unit. This reactor experienced the same drop in pH and similar gas production problems. The same technique was followed (as in the 30 day unit) in feeding effluent from the healthy reactor. In approximately the same number of days, the reactor reached a condition of stable pH and increasing gas production. The main difference between the start-up of this reactor vs. the 30 day reactor was that this unit was not seeded with digested sewage sludge.

3. 10 day HRT reactor--This reactor, also begun without a seed, never experienced the same problems as the others. Once feeding was started, it was never interrupted at any time. Feeding was initiated at a 40 day HRT, and gradually increased to 20 day, 13 day and finally 10 day HRT over a period of two weeks. From that point on, the unit was fed raw manure at a rate of $HRT = 10$ days, with the exception of three days when 25% of the feed consisted of digested sewage sludge. The desired condition of stable pH with increasing gas production was fairly easy to maintain in this reactor. It should be noted that this reactor was the source of the effluent used to stimulate gas production in the 20 day and 30 day reactors.

TABLE B2

Feeding Modes During Start-Up of a 10-Day HRT Bench-Scale

UHD Reactor (20 L Reactor Volume)

Day	Date	Gas (L/day) (STP)	pH	Feed
1			6.7	None
2			6.8	None
3				None
4	11/3	7.8		None
5	11/4	6.5	6.7	None
6	11/5	5.1	6.9	500 ml dairy cow manure
7	11/6	4.1	6.4	
8	11/7	4.6	6.6	
9	11/8	3.2	6.7	
10	11/9	3.9	6.6	1000 ml dairy cow manure
11	11/10	4.9	6.9	
12	11/11	3.7	6.7	
13	11/12	3.8	6.7	
14	11/13	4.1		
15	11/14	4.2	6.9	
16	11/15	3.9	6.7	1200 ml dairy cow manure
17	11/16	3.9	6.5	1500 ml dairy cow manure
18	11/17	4.3	6.6	1500 ml dairy cow manure
19	11/18	4.6	6.9	2000 ml dairy cow manure
20	11/19	3.7	6.9	
21	11/20	4.2	6.7	
22	11/21	4.8	6.6	
23	11/22	5.4	6.7	
24	11/23	5.5	6.7	
25	11/24	6.7	6.8	
26	11/25	6.0	6.6	
27	11/26	6.2	6.6	
28	11/27	5.8	6.7	
29	11/28	7.1	6.8	
30	11/29	7.5	6.7	
31	11/30	5.9	6.7	
32	12/1	7.0	6.7	
33	12/2	7.5	6.8	2000 ml (3/4 dairy cow manure;
34	12/3	7.3	6.7	1/4 effluent from
35	12/4	6.9	6.8	10-day HRT unit
36	12/5	7.8	7.0	
37	12/6	8.6	6.9	2000 ml dairy cow manure
38	12/7	9.1	7.1	2000 ml dairy cow manure

TABLE B3
Feeding Modes During Start-up of a 20-Day HRT
Bench-Scale UHD Reactor (20 ℓ Reactor Volume)

Day	Date	Gas (ℓ/day)	pH	Feed (8% TS)
1			6.3	None
2			6.1	None
3			6.1	None
4	10/27	11.7	6.1	250 ml dairy cow manure
5	10/28	7.5	6.0	240 ml dairy cow manure
6	10/29	7.1	5.8	500 ml dairy cow manure
7	10/30	6.2	6.0	None
8	10/31	5.6	6.0	None
9	11/1	6.2	5.9/6.7 ¹	None
10	11/2	6.7	6.1/7.0	250 ml dairy cow manure
11	11/3	2.9	6.2	250 ml dairy cow manure
12	11/4	6.1	5.9	1000 ml dairy cow manure
13	11/5	2.4	5.9	500 ml dairy cow manure
14	11/6	4.1	5.6	500 ml dairy cow manure
15	11/7	3.5	5.2/6.7	None
16	11/8	4.1	5.3	1000 ml (1/2 dairy cow manure;
17	11/9	4.2	5.4	1/2 effluent from 10-
18	11/10	5.0	5.4	day HRT unit)
19	11/11	4.1	5.5	
20	11/12	4.4	5.5	
21	11/13	4.4	5.8	
22	11/14	4.3	6.1	
23	11/15	3.9	6.2	
24	11/16	4.3	6.1	1000 ml (3/4 dairy cow manure;
25	11/17	4.3	6.1	1/4 effluent from 10-
26	11/18	4.6	6.4	day HRT unit)
27	11/19	4.6	6.4	
28	11/20	3.5	6.2	
29	11/21	3.3	6.1	1000 ml (effluent from 10-day
30	11/22	3.5	6.2	HRT unit)
31	11/23	3.4	5.9	
32	11/24	3.8	6.2	
33	11/25	4.0	6.3	
34	11/26	4.8	6.5	
35	11/27	5.3	6.5	
36	11/28	7.2	6.6	
37	11/29	8.1	6.7	1000 ml dairy cow manure
38	11/30	9.8	6.7	1000 ml dairy cow manure

¹ Dual pH values indicate pH readings at the effluent and influent ends of the reactor, respectively.

TABLE B4

Feeding Modes During Start-Up of a 30-Day HRT

Bench-Scale UHD Reactor (20 L Reactor Volume)

Day	Date	Gas (L/day)	pH	Feed (8% TS)
1			6.3	None
2			6.1	None
3			6.1	None
4	10/27	11.7	6.1	250 ml dairy cow manure
5	10/28	7.5	6.0	250 ml dairy cow manure
6	10/29	7.1	5.8	500 ml dairy cow manure
7	10/30	6.2	6.0	None
8	10/31	5.6	6.0	None
9	11/1	6.2	5.9/6.7 ¹	None
10	11/2	6.7	6.1/7.0	250 ml dairy cow manure
11	11/3	2.9	6.2	250 ml dairy cow manure
12	11/4	6.1	5.9	1000 ml dairy cow manure
13	11/5	2.4	5.9	500 ml dairy cow manure
14	11/6	4.1	5.6	500 ml dairy cow manure
15	11/7	3.5	5.2/6.7	None
16	11/8	4.1	5.3	1000 ml (1/2 dairy cow manure; 1/2 effluent from 10 day HRT unit)
17	11/9	4.2	5.4	
18	11/10	5.0	5.4	
19	11/11	4.1	5.5	
20	11/12	4.4	5.5	
21	11/13	4.4	5.8	
22	11/14	4.3	6.1	
23	11/15	3.9	6.2	
24	11/16	4.3	6.1	1000 ml (3/4 dairy cow manure; 1/4 effluent from 10 day HRT unit)
25	11/17	4.3	6.1	
26	11/18	4.6	6.4	
27	11/19	4.6	6.4	
28	11/20	3.5	6.2	
29	11/21	3.3	6.1	1000 ml (effluent from 10-day HRT unit)
30	11/22	3.5	6.2	
31	11/23	3.4	5.9	
32	11/24	3.8	6.2	
33	11/25	4.0	6.3	
34	11/26	4.8	6.5	
35	11/27	5.3	6.5	
36	11/28	7.2	6.6	
37	11/29	8.1	6.7	1000 ml dairy cow manure
38	11/30	9.8	6.7	100 ml dairy cow manure

¹Dual pH values indicate pH readings at the effluent and influent ends of the reactor, respectively.

TABLE B5

Summary of Steady State From Four CSTRs in Series and
CSTR Control Operated at a 12-day HRT, 8% TS Feed, 22°C

Parameter	Influent	Four CSTRs in Series				CSTR Control
Total Solids						
gm/l	80.00				69.87	64.85
% red.					12.7	18.9
Total Volatile Solids						
gm/l	71.69				61.86	56.12
% red.					13.7	20.7
Alkalinity (gm/l)	4.358				3.150	4.033
Volatile Acids (gm/l)					4.298	2.547
NH ₃ -N (gm/l)	0.340				0.672	0.737
Org. N (gm/l)	3.179				1.447	1.838
pH	6.88				6.65	7.03
CST (sec)	3103				2087	1915
Digester Gas						
		CSTRs in Series				CSTR Control
		No. 1	No. 2	No. 3	No. 4	Total
Gas Production						
l/day	0.41	0.76	0.88	1.16	3.21	1.96
l/l reactor	0.37	0.253	0.293	0.387	0.268	0.653
Gas Composition (% CH ₄ / % CO ₂)						
Methane Production						
l/day						
l/l reactor						
l gas/gm VS _A					0.044	0.109
l gas/gm VS _D					0.327	0.504

TABLE B6

Summary of Steady State Data From Four CSTRs in Series and
CSTR Control Operated at a 30-day HRT, 8% TS Feed, 22°C

Parameter	Influent	Four CSTRs in Series				CSTR Control
Total Solids						
gm/l	80.00				59.09	59.32
% red.					26.1	25.9
Total Volatile Solids						
gm/l	71.18				51.01	52.38
% red.					28.3	26.4
Alkalinity (gm/l)	4.23				3.08	2.97
Volatile Acids (gm/l)					1.608	0.400
NH ₃ -N (gm/l)	0.390				0.728	0.742
Org. N (gm/l)	2.202				1.532	1.518
pH	6.81				6.80	7.0
Digester Gas						
		CSTRs in Series				CSTR Control
		No. 1	No. 2	No. 3	No. 4	Total
Gas Production						
l/day	1.19	1.32	1.00	0.90	4.41	1.33
l/l reactor	0.400	0.440	0.333	0.300	0.368	0.443
Gas Composition (% CH ₄ / % CO ₂)						
Methane Production						
l/day						
l/l reactor						
l gas/gm VS _A					0.155	0.187
l gas/gm VS _D					0.547	0.707
l CH ₄ /gm VS _D						

TABLE B7

Summary of Steady State Data From Four CSTRs in Series and
CSTR Control Operated at a 10-day HRT, 8% TS Feed, 35°C

Parameter	Influent	Four CSTRs in Series				CSTR Control
Total Solids						
gm/l	80.00			59.60		58.87
% red.				25.5		26.4
Total Volatile Solids						
gm/l	70.36			50.73		50.79
% red.				27.9		27.8
Alkalinity (gm/l)	5.71			3.90		2.70
Volatile Acids (gm/l)				3.587		0.724
NH ₃ -N (gm/l)	0.524			1.041		0.546
Org. N (gm/l)	2.048			1.572		2.053
pH	7.14			7.0		6.8
CST (sec)	3860			1385		
Digester Gas						
	No. 1	CSTRs in Series				CSTR Control
		No. 2	No. 3	No. 4	Total	
Gas Production						
l/day	3.13	4.80	4.78	2.87	15.58	4.56
l/l reactor	1.043	1.600	1.593	0.957	1.298	1.520
Gas Composition (% CH ₄ / % CO ₂)	32/68	38/62	46/54	52/48		
Methane Production						
l/day	1.00	1.82	2.20	1.49	6.51	
l/l reactor	0.333	0.607	0.733	0.497	0.543	
l gas/gm VS _A					0.185	0.216
l gas/gm VS _D					0.661	0.777
l CH ₄ /gm VS _D					0.276	

TABLE B8

Summary of Steady State Data From Four CSTRs in Series and a
CSTR Control Operated at a 12-day HRT, 8% TS Feed, 35°C

Parameter	Influent	Four CSTRs in Series				CSTR Control
Total Solids						
gm/l	80.00			59.34		54.21
% red.				25.8		32.2
Total Volatile Solids						
gm/l	71.02			50.53		45.38
% red.				28.9		35.9
Alkalinity (gm/l)	5.633			6.700		3.933
Volatile Acids (gm/l)				2.939		0.430
NH ₃ -N (gm/l)	0.384			0.716		0.690
Org. N (gm/l)	2.446			1.475		1.578
pH	6.86			6.98		6.87
CST (Sec)	5203			2563		2550
Digester Gas						
	CSTRs in Series					CSTR
	No. 1	No. 2	No. 3	No. 4	Total	Control
Gas Production						
l/dry	2.36	4.98	4.64	4.42	16.40	4.62
l/l reactor	0.787	1.660	1.547	1.473	1.367	1.540
Gas Composition		49.2/		57.2/		
(% CH ₄ /% CO ₂)	50/50	50.8	51/49	42.8		56/44
Methane Production						
l/day	1.18	2.45	2.37	2.53	8.53	2.59
l/l reactor	0.393	0.817	0.790	0.843	0.711	0.862
l gas/gm VS _A					0.231	0.260
l gas/gm VS _D					0.800	0.720
l CH ₄ /gm VS _D					0.416	0.404

TABLE B9

Summary of Steady State Data From Four CSTRs in Series
 Operated at a 10-day HRT with 40% Recycle, 8% TS Feed, 35°C

Parameter	Influent	Effluent			
Total Solids					
gm/l	80.00				59.11
% red.					26.1
Total Volatile Solids					
gm/l	71.84				50.92
% red.					29.1
Alkalinity (gm/l)	2.818				2.470
Volatile Acids (gm/l)					
NH ₃ -N (gm/l)	0.209				0.578
Org. N (gm/l)	2.357				1.494
pH	6.48				6.70
Digester Gas					
CSTRs in Series					
	No. 1	No. 2	No. 3	No. 4	Total
Gas Production					
l/day	4.04	4.97	3.78	3.03	15.82
l/l reactor	1.347	1.657	1.260	1.010	1.318
Gas Composition					
(% CH ₄ /% CO ₂)	49.5/ 50.05	52.6/ 47.4	53.3/ 46.7	51/49	
Methane Production					
l/day	2.00	2.61	2.01	1.55	8.17
l/l reactor	0.667	0.870	0.670	0.517	0.681
l gas/gm VS _A					0.184
l gas/gm VS _D					0.631
l CH ₄ /gm VS _D					0.325

TABLE B10
Profile of Solids Concentrations in
Reactor Systems of Four CSTRs in Series

Temperature	HRT Days	Parameter	Reactor Stage			
			1	2	3	4
22°C	12	TS	70.4	71.1	69.4	69.8
		TVS	62.4	63.2	61.7	61.8
	30	TS	67.1	61.0	59.8	59.1
		TVS	59.0	53.5	52.1	51.1
35°C	12	TS	70.3	68.8	65.6	59.3
		TVS	63.1	60.9	57.9	50.5
	10	TS	65.8	60.1	59.3	59.1
		TVS	58.2	52.0	51.3	50.9
	10 (10% Recycle)	TS	79.3	73.6	70.0	64.4
		TVS	68.9	63.2	60.0	55.5

TABLE B11

Summary of Steady State Data From the Pilot-Scale UHD Reactor and
Control Operated at a 30-day HRT, 8 Percent TS Feed, 35°C

Parameter	Influent	Effluent	
		Pilot-Scale UHD Reactor	CSTR Control
Total Solids			
gm/l	86.1	48.7	51.7
% red.		43.4	40.0
Total Volatile Solids			
gm/l	74.6	38.0	41.4
% red.		49.1	44.5
Alkalinity (gm/l)	12.173	14.312	8.666
Volatile Acids (gm/l)		0.470	
NH ₃ -N (gm/l)	1.938	2.264	2.177
Org. N (gm/l)	1.662	1.488	
pH	7.77	7.90	7.30
CST (see)	3968	2565	1816
Gas Production			
l/day		5065	2.51
l/l reactor		0.910	0.837
Gas Composition (% CH ₄ /%CO ₂)		55.7/44.3	54/46
Methane Production			
l/day		2821	1.36
l/l reactor		0.505	0.453
l gas/gm VS _A		0.357	0.336
l gas/gm VS _D		0.728	0.756
l CH ₄ /gm VS _D		0.406	0.410

TABLE B12

Summary of Steady State Data from the Pilot-Scale UHD Reactor and
Control Operated at a 30-day HRT, 11 to 12 Percent TS Feed, 35°C

Parameter	Influent	Effluents	
		Pilot-Scale UHD Reactor	CSTR Control
Total Solids			
gm/l	112.4	88.5	80.3
% red.		21.3	28.6
Total Volatile Solids			
gm/l	97.0	72.0	66.0
% red.		25.7	32.0
Alkalinity (gm/l)	11.73	16.69	11.76
Volatile Acids (gm/l)		0.250	
NH ₃ -N (gm/l)	2.101	2.712	2.202
Free Ammonia (gm/l)	0.010	0.178	0.071
Org. N (gm/l)	2.27	2.73	2.138
pH	6.5	7.7	7.4
Gas Production			
l/day		6.45	2.77
l/l reactor		1.10	0.923
Gas Composition (% CH ₄ / %CO ₂)		61/39	64/36
Methane Production			
l/day		3750	1.77
l/l reactor		0.671	0.591
l gas/gm VS _A		0.333	0.246
l gas/gm VS _D		1.29	0.894
l CH ₄ /gm VS _D		0.789	0.571

Table B13

Data from Pilot-Scale UHD Reactor Feedstocks

Batch No.	Date	pH	NH ₄ -N mg/l	TKN mg/l	Alkalinity mg/l	Feed NH ₄ -N mg/l	Total Solids g.m/l	Volatile Solids g.m/l
8% - Total Solids Feed								
1	1/4	-	-	-	-	-	27.8	77.2
2	1/6	8.2	1721	-	-	346	89.9	78.9
3	1/11	8.0	1649	-	-	247	77.0	66.8
4	1/18	8.7	1936	3839	11,713	843	82.1	71.4
5	1/25	8.8	1451	2958	9,346	728	62.9	53.8
6	1/27	8.3	1516	3593	12,031	335	86.0	74.8
7	2/1	7.2	1601	3678	11,318	34	87.6	76.0
8	2/9	7.8	2019	37.0	11,692	161	87.6	76.0
9	2/15	7.7	1856	37.64	12,422	119	87.6	76.1
10	2/22	8.0	1869	3308	11,207	228	84.6	73.4
11	3/1	7.6	2090	3728	12,790	108	86.1	74.2
12	3/8	7.6	1991	3576	12,331	103	81.3	69.8
13	3/15	7.6	2.67	3502	12,721	112	79.2	66.3
14	3/17	7.8	2006	3540	11,589	160	77.7	66.4
15	3/22	8.0	1753	3569	12,552	214	85.6	74.2
16	3/29	7.5	1732	3728	12,164	72	81.1	68.7
17	4/4	7.3	1747	3319	11,230	46	706	60.4
18	4/14	7.5	1767	3347	12,414	77	82.8	70.6
19	4/14	7.2	1798	3210	9,630	38	72.4	62.7
20	4/19	7.6	2362	4724	14,059	122	90.9	72.1
21	4/28	7.1	2028	4336	12,988	34	81.9	71.5
22	4/28	7.6	2069	4768	13,023	107	83.1	72.6
10 - 12% Total Solids Feed								
23	5/9	6.7	2715	5652	14,682	18	102.9	89.7
24	5/11	7.1	2275	5071	11,609	38	113.4	100.2
25	5/19	7.0	2224	4170	10,922	30	106.6	92.1
26	5/24	6.5	2342	4685	13,595	10	116.8	101.8
27	5/31	6.6	2300	5600	14,048	12	119.2	103.7
28	6/2	7.5	2073	4643	11,846	86	118.6	102.8
29	6/9	6.5	1698	3600	9,705	7	111.5	98.1
30	6/16	6.5	1878	4035	11,429	8	112.8	98.9
31	6/21	6.5	2227	4168	12,234	10	117.0	98.1
32	6/22	6.7	2530	5241	13,770	17	105.0	90.4
33	6/30	6.5	2356	5236	12,468	10	110.8	95.3
34	7/7	6.5	2880	6585	13,854	12	104.9	89.5
35	7/12	6.4	1460	5094	11,980	5	104.6	89.9
36	7/14	6.7	2584	4845	12,458	17	112.1	98.0

Table B14

Volatile Acid Data for Reactor Systems of Four CSTRs in
Series and CSTR Controls

Sample	Date	Reactor Stage	HRT	mg/l as acetic acid				Total as Acetic	
				Acetic	Propionic	Butyric	Valeric		
Four CSTRs in Series 22°C	3-29-77		Days						
		1	12	2200	672	571	46	3489	
		2		3040	992	979	116	5127	
		3		2920	1088	870	93	4971	
	4		2200	1504	598	139	4441		
	6-08-77	1	30	510	760	285	174	1729	
		2		360	1440	88	162	2050	
		3		340	1842	60	-	2224	
		4		200	1408	-	-	1608	

	Control Reactors at 22°C	3-24-77		12	380	304	109	151	944
		4-12-77		12	450	1560	122	116	2248
		4-14-77		12	550	3620	204	174	4448
		6-08-77		30	240	160	-	-	400

	Four CSTRs in Series 35°C	3-29-77	1	12					
2				3080	2016	762	116	5974	
3				960	2336	1080	162	4538	
4				200	2496	163	162	3021	
6-08-77		1	10	1200	1728	238	174	3340	
		2	20%	700	992	150	64	1896	
		3	Recycle	200	1480	82	116	1878	
		4		160	1880	95	-	2135	
6-29-77		1	4	1580	1232	928	290	4030	
		2	20%	1080	1440	843	232	3595	
		3	Recycle	300	1920	1392	452	4064	
		4		200	2208	952	348	3708	

Control Reactors at 35°C		4-12-77		12	180	200	-	-	380
		4-14-77		12	590	160	-	-	750
		6-08-77		10	580	144	-	-	724

Four CSTRs in Series 60°C	6-7-77	1	12	4080	944	1366	133	6523	
		2		4960	1024	1387	325	7696	
		3		3840	768	1088	-	5696	
		4		3200	1152	1088	464	5904	
	6-29-77	1	4	2000	544	734	93	2711	
		2		1960	592	680	1624	4856	
		3		2320	640	680	1763	5403	
		4		2560	736	843	2018	6157	

				380					

TABLE B15

Electric Consumption in Pilot Scale Digester

Month	Room Temperature		Daily Electricity Consumption
	Average	Range	Average (Kilowatt-Hours)
December	57.4	55-60	2.24
January	57.6	55-59	2.21
February	57.6	55-60	2.22
March	58.6	57-60	2.26
April	61.1	52-72	1.94
May	63.5	52-76	1.72
June	65.2	57-76	1.63

TABLE B16

CHROMIUM AND COBALT ANALYSIS IN DIGESTER EFFLUENTS

Principle

A sample of digester effluent is dried, pulverized, digested by refluxing in nitric acid, filtered through a glass fiber filter and analyzed for Cr and Co by atomic absorption. In the digestion process, chromium is released from the Cr-tagged tracer material and solubilized in the trivalent form. Cobalt added to the fermentor as cobalt lithium EDTA should remain soluble in the effluent sample throughout the acid digestion procedure.

Procedure

1. Place an aliquot of effluent sample in a dish and dry to a constant weight at 103°C.
2. Weigh out 5g of dry sample material into a refluxing flask and add 20 ml concentrated nitric acid (analytical reagent grade) and 20 ml distilled water.
3. Reflux the mixture for 4 hours.
4. Filter through a glass fiber filter, measure the final volume of the filtrate and store filtrate in plastic containers.
5. Analyze the filtrate for Cr and Co by atomic absorption spectroscopy.
6. Determine the total solids concentration of the effluent sample according to Standard Methods. This may be obtained by simply measuring the volume of the sample aliquot used in step 1 and the weight of dry material yielded upon drying (103°C).
7. Samples analyzed for cobalt and chromium by this method should be run in duplicate.

Calculations

$$\text{Total solids g/l} = \frac{\text{g dry sample}}{\text{ml sample}} \times 1000$$

$$[\text{Me}]_s = \frac{[\text{Me}]_A \times V_A \times \text{Total Solids (g/l)}}{W \times 1000}$$

where: $[\text{Me}]_s$ = metal concentration in the effluent sample

W = weight of dry effluent sample used (5g)

$[\text{Me}]_A$ = metal tracer concentration in the filtered refluxed mixture

V_A = volume of filtered refluxed mixture.

Precision and Accuracy

The overall standard deviation of this method of Cr analysis is about 14 percent. Much of this variance may have been due to the atomic absorption step which alone was found to have a standard deviation of about 9 percent in the analysis of chromium in a nitric acid mixture. Quantitative recovery of chromium was obtained with the nitric acid digestion method of analysis when effluent samples containing .36 mg/l background chromium were spiked with enough tagged Cr fiber to increase sample chromium levels to 10.9 mg/l.

The standard deviation for cobalt analysis is virtually equal to the precision of atomic absorption methods (5 to 6 percent standard deviation). Since cobalt remained soluble in the effluent sample before and after the acid digestion step, recoveries were quantitative.

Appendix C
Semi-Solid Fermentation

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #1

Day	Date 1977	Gas Production ℓ/day ℓ/reactor/day		Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
Period 1: 20 day SRT, 25% TS Feed							
1	2/6	-	-	-	6.9	-	Start-up 10% TS mixture { 50% manure 50% "seed"
2	2/7	4.9	0.88	-	6.8	-	Volume 5.5ℓ
3	2/8	6.4	1.16	-	7.1	-	
4	2/9	7.2	1.31	-	7.4	-	Commence feeding every other day { 250 ml manure 50 g straw
5	2/10	7.4	1.35	-	-	-	
6	2/11	7.7	1.40	-	7.5	-	
7	2/12	-	-	-	7.0	-	
8	2/13	8.5	1.55	-	7.1	-	
9	2/14	-	-	-	-	-	
10	2/15	7.2	1.31	-	7.4	-	Volume reached to 4.5ℓ, feed every other day { 250 ml manure 50 g straw
11	2/16	9.0	1.80	-	7.2	-	
12	2/17	9.4	1.88	-	-	-	
13	2/18	8.5	1.70	-	-	-	
14	2/19	9.5	2.11	-	7.4	-	
15	2/20	7.6	1.69	-	-	-	
16	2/21	7.6	1.69	-	7.5	-	
17	2/22	5.9	1.31	-	-	-	
18	2/23	5.9	1.31	-	7.8	-	
19	2/24	7.4	1.64	-	-	-	
20	2/25	6.9	1.53	-	-	-	
21	2/26	-	-	-	7.6	-	
22	2/27	4.9	1.09	-	-	-	

DATA FROM A SEMI-SOLID FERMENTOR. (35°) - REACTOR #1 (cont.)

Day	Date 1977	Gas Production l/day	Gas Production l/reactor/day	Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
23	2/28	3.8	0.84	-	-	-	
24	3/1	6.2	1.38	-	7.2	-	
25	3/2	4.8	1.07	-	7.2	-	
26	3/3	7.5	1.67	-	-	-	
27	3/4	5.9	1.32	-	7.2	-	
28	3/5	8.4	1.86	-	-	-	
29	3/6	6.2	1.38	-	7.4	-	
30	3/7	8.3	1.84	-	-	-	
31	3/8	1.12	2.48	-	7.2	-	
32	3/9	8.8	1.56	-	-	-	
33	3/10	8.3	1.54	-	7.4	-	
34	3/11	7.0	1.56	45/55	-	-	
35	3/12	-	-	-	7.6	-	
36	3/13	7.4	1.54	-	-	-	
Period 2: 10 day SRT, 25% TE Feed							
37	3/14	5.7	1.26	49/51	7.1	-	daily feed { 250 ml manure 50 g straw
38	3/15	7.7	1.72	-	7.2	-	
39	3/16	8.5	1.38	-	7.1	-	
40	3/17	8.7	1.94	49/51	7.2	-	
41	3/18	7.2	1.60	-	7.1	-	
42	3/19	8.0	1.78	-	7.2	-	
43	3/20	6.6	1.46	-	7.3	-	
44	3/21	7.2	1.60	-	7.1	-	
45	3/22	8.2	1.82	-	7.2	-	

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #1 (cont.)

Day	Date 1977	Gas Production l/day	Gas Production l/reactor/day	Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
46	3/23	7.6	1.68	47/53	7.1	-	
47	3/24	6.7	1.48	-	7.1	20.0	
48	3/25	5.6	1.24	-	7.1	-	
49	3/26	6.4	1.42	-	-	-	
50	3/27	5.9	1.32	-	7.1	-	
51	3/28	5.9	1.32	-	7.2	-	
52	3/29	-	-	46/54	7.2	-	
53	3/30	5.6	1.24	-	7.1	-	
54	3/31	4.5	1.00	-	-	-	
55	4/1	4.6	1.02	-	7.4	-	
56	4/2	4.5	1.00	-	7.5	21.3	
57	4/3	4.4	0.98	-	-	-	
58	4/4	3.3	0.84	-	7.4	-	
59	4/5	4.1	0.92	48/52	-	-	
60	4/6	4.3	0.96	-	-	-	
61	4/7	4.1	0.92	-	7.2	-	
62	4/8	3.5	0.78	-	-	-	
Period 3: 20 day SRT, 25% TS Feed							
63	4/9	5.0	1.12	-	-	-	Feed every other day { 250 ml manure 50 g straw
64	4/10	-	-	-	7.3	-	
65	4/11	-	-	-	-	-	
66	4/12	4.0	0.88	-	-	-	
67	4/13	3.8	0.84	-	7.2	-	

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #1 (cont).

Day	Date 1977	Gas Production l/day l/reactor/day		Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
68	4/14	4.3	0.96	47/53	7.4	-	
69	4/15	3.8	0.84	-	-	-	
70	4/16	4.2	0.94	-	-	-	
71	4/17	4.3	0.96	-	7.1	-	
72	4/18	4.1	0.92	-	7.1	-	
73	4/19	4.3	0.96	52/48	7.2	-	Start daily feeding { 125 ml manure 25 g straw
74	4/20	4.1	0.92	-	7.3	-	
75	4/21	4.6	1.02	-	7.2	-	
76	4/22	4.4	0.98	51/49	7.5	-	
77	4/23	5.0	1.12	-	7.3	-	
78	4/24	5.5	1.22	-	7.5	-	
79	4/25	6.1	1.36	-	7.6	20.6	
80	4/26	6.4	1.42	-	7.6	-	
81	4/27	6.8	1.50	-	7.5	-	
82	4/28	6.3	1.40	-	-	-	
83	4/29	7.5	1.67	-	7.7	-	
84	4/30	6.8	1.50	-	7.5	-	
85	5/1	6.0	1.34	-	7.7	-	
86	5/2	5.9	1.32	-	7.7	-	
87	5/3	6.7	1.48	-	7.8	-	
88	5/4	6.8	1.50	52/48	7.8	-	
89	5/5	7.0	1.56	-	7.7	-	
90	5/6	6.6	1.46	-	7.9	-	

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #1 (cont.)

Day	Date 1977	Gas Production l/day	Gas Production l/reactor/day	Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
91	5/7	6.5	1.44	-	7.7	-	
92	5/8	7.0	1.56	-	7.7	-	
93	5/9	7.0	1.56	-	7.9	-	
94	5/10	6.2	1.38	-	7.8	-	
95	5/11	6.5	1.44	-	8.0	-	
96	5/12	6.9	1.54	52/48	7.9	-	
97	5/13	6.8	1.52	-	8.0	-	
Period 4: 20 day SRT, 35% T.S. Feed							
98	5/14	7.3	1.62	-	8.1	-	Daily feed { 90 ml manure 35 g straw
99	5/15	6.6	1.46	-	8.1	-	
100	5/16	7.2	1.60	-	8.0	-	
101	5/17	6.7	1.48	-	7.9	-	
102	5/18	7.0	1.56	-	7.9	-	
103	5/19	6.2	1.38	49/51	8.0	-	
104	5/20	6.7	1.48	-	7.9	-	
105	5/21	5.9	1.32	-	7.6	-	
106	5/22	5.8	1.28	-	7.7	-	
107	5/23	5.1	1.14	-	7.6	27.4	
108	5/24	5.0	1.12	-	7.2	-	
109	5/25	5.0	1.10	49/51	7.1	-	
110	5/26	-	-	-	7.1	-	
111	5/27	3.6	0.89	-	6.7	-	
112	5/28	3.0	0.66	-	6.3	-	
113	5/29	2.1	0.46	-	-	-	Stop feeding due to low pH

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #1 (cont.)

Day	Date 1977	Gas Production		Gas Composition	pH	Effluent Solids	Remarks
		g/day	g/reactor/day	% CH ₄ /CO ₂		% T.S.	
114	5/30	2.1	0.46	-	6.1	-	
115	5/31	1.9	0.42	-	6.0	-	
116	6/1	1.2	0.26	-	6.1	-	
117	6/2	1.1	0.24	-	5.9	-	
118	6/3	1.8	0.40	-	6.0	-	
119	6/4	1.3	0.28	-	5.8	-	
120	6/5	1.4	0.30	-	5.8	-	
121	6/6	0.5	0.10	-	5.9	-	

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #2

Day	Date 1977	Gas Production l/day l/reactor/day		Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
Period 1: 20 day SRT, 25% T.S. Feed							
1	2/6	-	-	-	6.7	-	Start up 10% T.S. mixture 50%{manure 50%{"seed"
2	2/7	3.1	0.68	-	6.8	-	
3	2/8	3.2	0.72	-	6.9	-	
4	2/9	5.4	1.20	-	6.9	-	
5	2/10	5.6	1.24	-	-	-	Commence feeding every other day { 370 ml manure 80 ml plastic balls
6	2/11	6.2	1.38	-	7.6	-	
391 7	2/12	6.3	1.40	-	-	-	
8	2/13	6.7	1.48	-	7.3	-	
9	2/14	6.5	1.44	-	-	-	
10	2/15	6.7	1.48	-	7.2	-	
11	2/16	6.5	1.44	-	-	-	
12	2/17	-	-	-	7.3	-	
13	2/18	5.1	1.14	-	-	-	
14	2/19	6.3	1.40	-	7.3	-	
15	2/20	-	-	-	-	-	
16	2/21	5.1	1.14	-	7.4	-	
17	2/22	4.6	1.02	-	-	-	
18	2/23	4.1	0.92	-	-	-	
19	2/24	4.4	0.98	-	-	-	
20	2/25	5.0	1.12	-	-	-	
21	2/26	-	-	-	7.4	-	
22	2/27	3.7	0.82	-	-	-	
23	2/28	4.5	1.00	-	7.4	-	

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #2 (Cont.)

Day	Date 1977	Gas Production ℓ/day	Gas Production ℓ/reactor/day	Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
24	3/1	2.6	1.00	-	-	-	
25	3/2	4.0	0.53	-	7.4	-	
26	3/3	-	-	-	-	-	
27	3/4	4.7	1.04	-	7.2	-	
28	3/5	4.9	1.09	-	-	-	
29	3/6	-	-	-	7.4	-	
30	3/7	4.6	1.02	-	-	-	
31	3/8	4.3	0.96	-	7.0	-	
32	3/9	3.9	0.87	-	-	-	
33	3/10	3.3	0.73	-	7.2	-	
34	3/11	3.1	0.69	-	-	-	
35	3/12	4.1	0.91	-	7.2	-	
36	3/13	-	-	-	-	-	
37	3/14	2.2	0.55	-	7.2	-	
Period 2: 10 day SRT, 25% T.S.							
38	3/15	2.4	0.60	-	-	-	Daily feed { 370 ml manure 80 ml plastic balls
39	3/16	2.7	0.68	-	7.2	-	
40	3/17	3.2	0.80	-	7.4	-	
41	3/18	2.8	0.70	-	7.3	-	
42	3/19	2.3	0.58	-	7.3	-	
43	3/20	2.4	0.50	-	7.2	-	
44	3/21	2.0	0.50	-	7.1	-	
45	3/22	2.7	0.58	-	-	-	Operation terminated

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DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #2 (Cont.)

Day	Date 1977	Gas Production		Gas Composition	pH	Effluent Solids	Remarks
		l/day	l/reactor/day	% CH ₄ /CO ₂		% T.S.	
45	6/15	2.6	0.65	-	6.6	-	
46	6/16	2.3	0.59	-	6.7	-	
Period 3: 10 day SRT, 5% T.S.							
47	6/17	2.6	0.65	-	6.7	-	Daily feed { 280 ml tap water 80 ml manure 10 g straw
48	6/18	2.7	0.68	-	6.6	-	
49	6/19	2.8	0.70	-	6.6	-	
50	6/20	3.0	0.74	-	6.6	-	
51	6/21	2.6	0.65	-	6.4	-	
52	6/22	2.5	0.63	-	6.6	-	
53	6/23	2.7	0.68	-	6.8	-	
54	6/24	2.9	0.72	-	6.8	-	
55	6/25	3.3	0.83	-	6.7	-	
56	6/26	3.4	0.86	-	6.7	-	
57	6/27	3.6	0.90	-	6.7	-	
58	6/28	3.6	0.90	-	6.8	-	
59	6/29	3.6	0.90	-	6.7	-	
60	6/30	3.4	0.86	-	6.8	-	
61	7/1	3.3	0.83	-	6.8	-	Stop feeding reactor

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #2 (Cont.)

Day	Date 1977	Gas Production l/day	Gas Production l/reactor/day	Gas Composition % CH ₄ /CO ₂	pH	Effluent Solids % T.S.	Remarks
Period 1: 20 day SRT, 2.5% T.S. Feed							
1	5/2	0.7	0.18	-	-	-	Start up 25% T.S. mixture { 50% manure & straw 50% "seed"
2	5/3	1.7	0.43	-	6.3	-	
3	5/4	2.1	0.52	-	6.5	-	
4	5/5	2.6	0.55	-	6.7	-	
5	5/6	2.7	0.58	-	6.6	-	Commence daily feeding { 170 ml tap water 20 ml manure 2.5 g straw
6	5/7	2.1	0.52	-	6.6	-	
7	5/8	2.1	0.53	-	-	-	
8	5/9	2.1	0.53	-	6.7	-	
9	5/10	2.0	0.50	-	5.6	-	Sample to Albany microbial analysis
10	5/11	1.8	0.45	-	6.6	-	
11	5/12	1.4	0.36	-	6.6	-	
12	5/13	2.2	0.54	-	5.5	-	
13	5/14	1.6	0.41	-	5.6	-	
14	5/15	1.6	0.41	-	5.6	-	
15	5/16	1.5	0.38	-	6.6	-	
16	5/17	1.4	0.36	-	6.6	-	
17	5/18	1.6	0.41	-	6.7	-	
18	5/19	1.4	0.34	-	6.5	-	
19	5/20	1.5	0.38	-	6.5	-	
20	5/21	1.4	0.34	-	6.5	-	
21	5/22	1.4	0.36	-	6.6	-	
22	5/23	1.4	0.36	-	6.4	-	
23	5/24	1.4	0.34	-	6.5	-	

DATA FROM A SEMI-SOLID FERMENTOR (35°) - REACTOR #2 (Cont.)

Day	Date 1977	Gas Production		Gas Composition	pH	Effluent Solids	Remarks
		l/day	l/reactor/day	% CH ₄ /CO ₂		% T.S.	
Period 2: 20 day SRT, 5% T.S. feed							
24	5/25	1.4	0.36	-	6.6	-	Daily feed { 140 ml tap water 40 ml manure 5 g straw
25	5/26	1.4	0.36	-	6.6	-	
26	5/27	1.7	0.43	-	6.5	-	
27	5/28	1.9	0.47	-	6.4	-	
28	5/29	1.7	0.43	-	-	-	
29	5/30	1.7	0.43	-	6.4	-	Gas composition: 55% CH ₄ 45% CO ₂
30	5/31	1.7	0.43	-	6.5	-	
31	6/1	2.2	0.54	-	6.5	-	
32	6/2	2.1	0.52	-	6.5	-	
33	6/3	2.1	0.52	-	6.5	-	
34	6/4	2.3	0.59	-	6.6	-	
35	6/5	2.5	0.63	-	6.6	-	
36	6/6	2.6	0.65	-	6.7	-	
37	6/7	-	-	-	6.6	-	
38	6/8	2.7	0.68	-	6.7	-	
39	6/9	2.7	0.68	-	6.7	-	
40	6/10	2.6	0.65	-	6.7	-	
41	6/11	2.5	0.63	-	6.8	-	
42	6/12	2.3	0.70	-	6.7	-	
43	6/13	-	-	-	6.7	-	
44	6/14	2.7	0.68	-	6.6	-	

APPENDIX D

THERMOPHILIC INTERNAL

LIQUID-SOLID SEPARATION REACTOR DATA

TABLE D1 STEADY STATE DATA FROM THE INTERNAL
LIQUID-SOLID SEPARATION REACTOR OPERATED
AT A 5 DAY HRT, 5 DAY SRT, 5 PERCENT TS
FEED; FLOAT LAYER REMIXED AFTER WASTING
EXCESS FLOAT SOLIDS.

Influent									
Date	Flow ℓ/day	T.S. %	T.V.S. %	pH	Alk mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	T.V.A. mg/ℓ	Temp °C
Average	36.0	5.0	89.76	7.0	2696	198	2066	-	-
Liquid Effluent									
2-7-77	33.0	3.11	85.22	7.0	-	-	-	-	49
3-8-77	32.0	2.94	84.35	7.0	-	-	-	-	49
3-9-77	32.5	2.90	83.45	6.9	-	-	-	-	50
3-10-77	32.5	3.05	83.93	7.2	-	-	-	-	51
Average	32.5	3.00	84.24	7.0	-	-	-	-	50
Solids Effluent									
3-7-77	3.0	14.02	91.92	7.0	-	-	-	-	54
3-8-77	4.0	13.54	91.46	7.1	-	-	-	-	52
3-9-77	3.5	13.51	91.82	7.0	-	-	-	-	53
3-10-77	3.5	13.73	92.27	7.0	-	-	-	-	53
Gas Production									
Date		ℓ/day		CH ₄ CO ₂					
3-7-77		325.77		-					
3-8-77		314.66		-					
3-9-77		323.23		-					
3-10-77		315.91		-					
Average		321.1		-					

TABLE D2 . STEADY STATE DATA FROM THE INTERNAL
LIQUID-SOLID SEPARATION REACTOR OPERATED
AT A 5 DAY HRT, 4.62 SRT, 5 PERCENT TS
FEED; ALL FLOAT SOLIDS REMOVED EACH DAY.

Influent									
Date	Flow ℓ/day	T.S. %	T.V.S. %	pH	Alk mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	36.0	5.0	88.32	6.9	2884	216	1470	-	-
Liquid Effluent									
2-13-77	31.0	2.95	82.71	7.1	-	-	-	-	49
2-14-77	32.0	2.88	82.84	7.1	-	-	-	-	50
3-15-77	32.0	2.80	82.14	7.0	-	-	-	-	50
2-16-77	33.0	2.69	83.27	7.0	-	-	-	-	50
2-17-77	31.0	2.81	82.92	7.0	-	-	-	-	50
2-18-77	31.0	2.65	82.26	7.1	-	-	-	-	50
2-19-77	31.5	2.79	83.51	7.1	-	-	-	-	49
2-20-77	31.0	2.57	81.32	7.0	-	-	-	-	50
2-21-77	31.0	2.90	83.10	7.1	-	-	-	-	49
2-22-77	31.5	2.81	82.56	7.1	-	-	-	-	51
2-23-77	31.0	2.76	81.88	7.1	-	-	-	-	50
2-24-77	32.0	2.84	82.75	7.0	-	-	-	-	50
2-25-77	31.0	2.73	81.68	7.0	-	-	-	-	50
Average	31.5	2.78	82.52	7.1	-	-	-	-	50
Solids Effluent									
2-13-77	5.0	11.58	92.93	6.9	-	-	-	-	56
2-14-77	4.0	11.83	93.20	7.1	-	-	-	-	54
2-15-77	4.0	12.12	93.10	7.0	-	-	-	-	54
2-16-77	3.0	10.98	93.01	6.9	-	-	-	-	53
2-17-77	5.0	11.40	92.33	7.0	-	-	-	-	51
2-18-77	5.0	11.16	92.55	6.9	-	-	-	-	50
2-19-77	4.5	11.66	92.77	6.9	-	-	-	-	50
2-20-77	5.0	10.72	88.90	7.1	-	-	-	-	51
2-21-77	4.5	11.67	92.29	6.9	-	-	-	-	51
2-22-77	4.5	11.64	92.17	7.0	-	-	-	-	52
2-23-77	5.0	10.94	91.58	7.0	-	-	-	-	51
2-24-77	4.0	12.10	91.30	6.9	-	-	-	-	52

Solids Effluent (Cont.)

2-25-77	5.0	11.54	90.84	6.9	-	-	-	-	51
Average	4.5	11.50	92.07	7.0	-	-	-	-	52

Gas Production

Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂
2-13-77	152.89	-	2-20-77	-	-
2-14-77	-	-	2-21-77	-	-
2-15-77	208.13	-	2-22-77	-	-
2-16-77	-	-	2-23-77	240.22	-
2-17-77	-	-	2-24-77	-	-
2-18-77	241.29	-	2-25-77	191.08	-
2-19-77	-	-	Average	206.7	-

TABLE D3 . STEADY STATE DATA FROM THE INTERNAL
LIQUID-SOLID SEPARATION REACTOR OPERATED
AT A 5-DAY HRT, 5 DAY SRT, 5 PERCENT TS
FEED; NO REMIXING OF FLOAT LAYER AFTER
WASTING EXCESS FLOAT SOLIDS.

Influent									
Date	Flow ℓ/day	T.S. %	T.V.S. %	pH	Alk mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	36.0	5.0	88.73	6.8	2750	248	1547	-	-
Liquid Effluent									
3-29-77	32.0	2.75	81.82	7.0	-	-	-	-	50
3-30-77	33.0	2.68	82.07	7.1	-	-	-	-	51
3-31-77	32.5	2.90	88.60	7.0	-	-	-	-	50
4-01-77	32.0	2.50	80.86	7.0	-	-	-	-	50
4-02-77	32.0	2.91	82.13	7.1	-	-	-	-	51
4-03-77	32.5	2.72	81.27	7.0	-	-	-	-	49
4-04-77	31.0	3.06	83.83	6.9	-	-	-	-	49
4-05-77	33.0	2.78	81.16	7.1	-	-	-	-	49
4-06-77	32.5	2.71	80.69	7.2	2500	448	-	-	49
4-07-77	32.5	2.83	81.82	7.0	-	-	-	-	48
6-19-77	34.0	3.44	85.47	6.8	2200	336	-	398	48
6-20-77	33.0	3.45	85.90	6.8	-	-	-	-	50
6-21-77	34.0	3.10	84.84	6.8	-	-	-	-	48
6-22-77	32.5	3.55	85.92	6.8	-	-	-	-	51
6-23-77	33.5	3.41	86.51	6.9	-	-	-	-	50
6-24-77	34.0	-	-	6.9	-	-	-	-	50
6-25-77	34.0	3.02	84.77	7.0	-	-	-	-	50
6-26-77	34.0	2.58	81.01	6.9	-	-	-	-	50
6-27-77	34.0	3.02	93.71	7.0	-	-	-	-	50
6-28-77	34.0	3.50	85.28	7.0	2100	476	-	199	49
Average	33.0	3.00	83.29	7.0	2267	420	-	299	50
Solids Effluent									
3-29-77	4.0	14.41	91.71	7.1	-	-	-	-	51
3-30-77	3.0	13.31	71.71	7.1	-	-	-	-	53
3-31-77	3.5	14.20	91.96	7.0	-	-	-	-	53
4-01-77	4.0	14.55	91.12	6.9	-	-	-	-	53
4-02-77	4.0	14.15	91.58	6.9	-	-	-	-	54
4-03-77	3.5	13.74	91.48	7.1	-	-	-	-	51

Solids Effluent (Cont.)

4-04-77	5.0	12.61	91.28	6.8	-	-	-	-	52
4-05-77	3.0	13.78	91.33	7.0	-	-	-	-	52
4-06-77	3.5	13.25	91.50	7.0	-	-	-	-	52
4-07-77	3.5	13.74	91.15	7.1	-	-	-	-	51
6-19-77	2.0	13.16	90.75	7.0	-	-	-	-	50
6-20-77	3.0	12.81	88.59	6.8	-	-	-	-	59
6-21-77	2.0	12.84	89.34	6.8	-	-	-	-	53
6-22-77	3.5	13.06	89.44	6.8	-	-	-	-	54
6-23-77	2.5	13.37	90.48	6.7	-	-	-	-	53
6-24-77	2.0	13.86	88.63	6.8	-	-	-	-	54
6-25-77	2.0	13.33	88.85	6.8	-	-	-	-	53
6-26-77	2.0	13.70	88.65	7.0	-	-	-	-	54
6-27-77	2.0	13.96	87.66	7.1	-	-	-	-	53
6-28-77	2.0	13.66	89.27	7.0	-	-	-	-	51
Average	3.0	13.61	90.18	6.7	-	-	-	-	53

Gas Production

Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂
3-29-77	340.85	-	4-03-77	314.95	-
3-30-77	348.73	-	4-04-77	291.88	-
3-31-77	303.75	-	4-05-77	401.73	-
4-01-77	287.80	-	4-06-77	413.66	-
4-02-77	309.05	-	4-07-77	361.06	-
6-14-77	453.51	53.1/46.9	6-24-77	525.47	-
6-20-77	487.96	-	6-25-77	474.77	-
6-21-77	554.04	54.8/45.2	6-26-77	408.60	-
6-22-77	535.51	-	6-27-77	493.75	-
6-23-77	508.80	-	6-28-77	505.28	59.0/91.0
			Average	416.06	55.6/44.4

TABLE D4 . STEADY STATE DATA FROM THE INTERNAL
LIQUID-SOLID SEPARATION REACTOR OPERATED
AT A 5 DAY HRT, 7.6 DAY SRT, FLOAT SOLIDS
REMOVED ON A 10 DAY RETENTION TIME BASIS,
5 PERCENT TS FEED; NO REMIXING OF FLOAT
LAYER.

Influent									
Date	Flow ℓ/day	T.S. %	T.V.S. %	pH	Alk mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	36.0	5.0	88.76	7.0	302.6	342	1921	-	-
Liquid Effluent									
4-23-77	32.0	2.95	81.69	7.2	2800	677	-	563	43
4-24-77	32.0	2.83	80.91	7.2	-	-	-	-	49
4-25-77	33.0	2.95	81.78	7.3	2800	616	-	-	49
4-26-77	33.0	2.78	82.46	7.0	-	-	-	-	50
4-27-77	32.5	2.95	83.25	7.1	-	-	-	-	41
4-28-77	33.0	2.05	83.31	7.2	-	-	-	-	50
4-29-77	33.0	2.79	83.45	7.0	2300	476	-	-	49
4-30-77	33.0	2.76	83.39	7.0	-	-	-	-	49
5-01-77	32.5	2.59	79.92	7.0	-	-	-	-	48
5-02-77	33.0	2.46	80.33	7.0	2600	546	-	352	50
Average	32.5	2.76	82.05	7.1	2625	579	-	458	49
Solids Effluent									
4-23-77	4.0	13.72	91.25	7.3	-	-	-	-	53
4-24-77	4.0	13.40	89.93	7.3	-	-	-	-	52
4-25-77	3.0	14.20	91.25	7.0	-	-	-	-	53
4-26-77	3.0	13.90	90.16	7.1	-	-	-	-	53
4-27-77	3.5	13.36	89.76	7.1	-	-	-	-	53
4-28-77	3.0	14.34	90.08	7.3	-	-	-	-	53
4-29-77	3.0	13.81	90.00	7.2	-	-	-	-	51
4-30-77	3.0	13.40	89.78	7.1	-	-	-	-	51
5-01-77	3.5	13.58	89.55	7.1	-	-	-	-	51
5-02-77	3.0	10.39	89.22	7.2	-	-	-	-	54
Average	3.5	13.47	90.10	7.2	-	-	-	-	52

Gas Production

Date	ℓ/day	CH ₄ CO ₂
4-23-77	491.72	-
4-24-77	482.22	-
4-25-77	463.03	-
4-26-77	568.00	-
4-27-77	530.99	-
4-28-77	461.32	-
4-29-77	382.14	57.6/42.4
4-30-77	401.43	-
5-01-77	421.53	-
5-02-77	453.00	54/46
Average	465.54	55.8/44.2

TABLE D5 . STEADY STATE DATA FROM THE INTERNAL
LIQUID-SOLID SEPARATION REACTOR OPERATED
AT A 5 DAY HRT, 225 DAY SRT, FLOAT SOLIDS
REMOVED ON A 10 DAY RETENTION TIME BASIS,
5 PERCENT TS FEED, RECYCLED EFFLUENT USED
FOR DILUTION, NO REMIXING OF FLOAT LAYER.

Influent									
Date	Flow l/day	T.S. %	T.V.S. %	pH	Alk mg/l	NH ₄ N mg/l	TKN mg/l	T.V.A. mg/l	Temp. °C
Average	36.0	5.0	89.78	6.3	1725	123	1365	-	-
Liquid Effluent									
5-22-77	35.5	3.98	82.16	7.0	3000	700	-	-	50
5-23-77	35.5	4.03	82.66	7.2	3300	532	-	-	51
5-24-77	35.0	4.19	82.93	7.1	-	-	-	505	52
5-26-77	35.5	4.07	82.49	7.1	-	-	-	-	51
5-27-77	35.5	4.42	83.55	7.2	-	-	-	-	51
5-28-77	35.5	4.21	82.80	7.1	-	-	-	-	52
5-29-77	35.5	4.37	82.29	7.0	-	-	-	-	49
5-30-77	35.5	4.30	82.33	7.0	-	-	-	-	51
5-31-77	35.5	4.28	82.98	7.1	-	-	-	-	50
6-01-77	35.5	4.31	82.68	7.1	3400	672	-	359	50
6-02-77	35.5	4.32	83.80	7.1	-	-	-	-	50
Average	35.5	4.21	82.69	7.1	3233	635	-	391	51
Solids Effluent									
5-22-77	0.5	11.65	87.66	7.3	-	-	-	-	51
5-23-77	0.5	11.75	87.78	7.3	-	-	-	-	52
5-24-77	1.0	12.23	87.68	7.2	-	-	-	-	52
5-25-77	0.5	11.75	86.88	7.3	-	-	-	-	51
5-26-77	0.5	11.62	88.39	7.3	-	-	-	-	51
5-27-77	0.5	12.11	87.95	7.3	-	-	-	-	54
5-28-77	0.5	11.90	87.57	7.4	-	-	-	-	52
5-29-77	0.5	11.92	87.20	7.1	-	-	-	-	56
5-30-77	0.5	11.76	88.20	7.2	-	-	-	-	52
5-31-77	0.5	11.47	87.70	7.1	-	-	-	-	52
6-01-77	0.5	11.33	89.13	7.2	-	-	-	-	52
6-02-77	0.5	12.01	88.51	7.1	-	-	-	-	52
Average	0.5	11.79	87.89	7.2	-	-	-	-	52

Gas Production					
Date	ℓ/day	CH ₄ CO ₂	Date	ℓ/day	CH ₄ CO ₂
5-22-77	171.28	51/49	5-28-77	171.39	-
5-23-77	190.40	-	5-29-77	109.09	-
5-24-77	173.61	-	5-30-77	129.21	-
5-25-77	185.70	-	5-31-77	151.02	-
5-26-77	149.14	-	6-01-77	139.00	-
5-27-77	126.13	52/48	6-02-77	139.90	50.1/46.9
Average				152.99	52/48

APPENDIX E
Anaerobic AFEB Fermentation Data

TABLE E1. STEADY STATE DATA FOR THE ANAEROBIC AFEB
OPERATED AT A 1.2-DAY HRT, 1% T.S. FEED

Influent											
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	Tot. COD mg/ℓ	Sol. COD mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	3.58	1.0	84.48	6.7	803	40	412	11520	2304	--	--
Effluent											
3-10-77	3.60	--	--	6.6	--	--	--	--	--	--	35
3-11-77	3.60	--	--	6.6	--	--	--	--	--	--	35
3-12-77	3.60	--	--	6.5	--	--	--	--	--	--	35
3-13-77	3.60	--	--	6.5	--	--	--	--	--	--	36
3-14-77	3.70	--	--	6.5	--	--	--	--	--	--	34
3-15-77	3.50	--	--	6.6	--	--	--	--	--	--	35
3-16-77	3.55	1.07	85.98	6.6	1100	112	546	14938	1230	587	34
3-17-77	3.50	0.61	80.33	6.5	900	94	392	9351	2061	480	35
Average	3.58	0.84	83.16	6.6	1000	103	469	12145	1646	532	35

Gas Production									
Date	3-10-77	3-11-77	3-12-77	3-13-77	3-14-77	3-15-77	3-16-77	3-17-77	Average
ℓ/day	3.71	3.20	3.31	2.86	2.86	2.32	2.96	3.14	3.05

TABLE E2. STEADY STATE DATA FOR THE ANAEROBIC AFEB
OPERATED AT A 1.2-DAY HRT, 2% T.S. FEED

Influent											
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	Tot. COD mg/ℓ	Sol. COD mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	3.52	2.0	82.97	6.8	1321	207	878	22565	6068	--	--
Effluent											
4-26-77	3.50	--	--	6.7	--	--	--	--	--	--	36
4-27-77	3.55	--	--	6.6	--	--	--	--	--	--	36
4-28-77	3.50	--	--	6.8	--	--	--	--	--	--	36
4-29-77	3.60	--	--	6.7	--	--	--	--	--	--	36
4-30-77	3.50	--	--	6.7	--	--	--	--	--	--	35
5-01-77	3.50	1.31	76.34	6.8	1900	470	1204	14962	2235	491	35
5-02-77	3.50	1.30	76.92	6.8	1800	479	1134	14394	2159	520	35
Average	3.52	1.31	76.63	6.7	1850	475	1169	14678	2197	506	36
Gas Production											
Date	4-26-77	4-27-77	4-28-77	4-29-77	4-30-77	5-01-77	5-02-77	Average			
ℓ/day	12.48	--	--	--	11.51	12.25	12.75	12.25			

TABLE E3. STEADY STATE DATA FOR THE ANAEROBIC AFEB
OPERATED AT A 0.6-DAY HRT, 2% T.S. FEED

Influent											
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	Tot. COD mg/ℓ	Sol. COD mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	7.08	2.0	83.97	6.6	861	64	875	19715	5833	--	--
Effluent											
5-23-77	7.0	--	--	6.5	--	--	--	--	--	--	34
5-24-77	7.0	--	--	6.5	--	--	--	--	--	--	34
5-24-77	7.0	--	--	6.5	--	--	--	--	--	--	35
5-25-77	7.0	1.73	81.50	6.5	1400	210	1015	19669	3419	612	35
5-25-77	6.8	--	--	6.5	--	--	--	--	--	--	34
5-26-77	7.0	1.76	82.29	6.5	1400	224	1078	23897	3622	714	35
5-26-77	7.0	--	--	6.5	--	--	--	--	--	--	35
5-27-77	7.8	1.71	83.63	6.5	1300	203	896	17963	3500	831	35
Average	7.08	1.73	82.51	6.5	1367	212	996	20510	3514	719	35
Gas Production											
Date	5-23-77	5-24-77	5-24-77	5-25-77	5-25-77	5-26-77	5-26-77	5-27-77	Average		
ℓ/day	15.80	16.42	15.34	--	--	14.40	15.84	14.86	15.44		

TABLE E4. STEADY STATE DATA FOR THE ANAEROBIC AFEB
OPERATED AT A 0.3-DAY HRT, 2% T.S. FEED

Influent											
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ H mg/ℓ	TKN mg/ℓ	Tot. COD mg/ℓ	So1. COD mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	14.09	2.0	83.22	5.8	1029	103	814	17833	5840	--	--
Effluent											
6-6-77	14.0	--	--	6.5	--	--	--	--	--	--	35
6-7-77	14.0	--	--	6.5	1200	189	--	--	--	--	35
6-7-77	14.4	--	--	6.5	--	--	--	--	--	--	35
6-7-77	14.0	--	--	6.5	1300	196	--	--	--	--	34
6-7-77	14.0	--	--	6.5	--	--	--	--	--	--	34
6-8-77	14.4	--	--	6.5	--	--	--	--	--	--	35
6-8-88	14.0	1.70	--	6.5	1200	182	1260	20644	3655	419	35
6-8-77	14.0	1.75	82.86	6.5	1200	196	1120	19886	3558	378	34
6-8-77	14.0	1.85	83.78	6.5	1200	182	1260	24053	3996	409	35
Average	14.09	1.77	83.32	6.5	1220	189	1213	21528	3870	402	35
Gas Production											
Date	6-6-77	6-7-77	6-7-77	6-7-77	6-7-77	6-8-77	6-8-77	6-3-77	6-8-77	Average	
ℓ/day	21.28	21.88	20.00	23.04	21.52	22.24	20.80	20.92	21.88	21.51	

TABLE E5. STEADY STATE DATA FOR THE ANAEROBIC AFEB
OPERATED AT A 0.15-DAY HRT, 2% T.S. FEED

Influent											
Date	Flow l/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/l	NH ₄ H mg/l	TKN mg/l	Tot. COD mg/l	Sol. COD mg/l	T.V.A. mg/l	Temp. °C
Average	28.0	2.0	82.25	6.9	1169	145	868	21525	6284	--	--
Effluent											
6-10-77	28.0	--	--	6.6	--	--	--	--	--	--	34
6-10-77	28.0	--	--	6.5	1300	189	--	--	--	--	34
6-10-77	28.0	--	--	6.6	--	--	--	--	--	--	34
6-10-77	28.0	--	--	6.5	--	--	--	--	--	--	33
6-11-77	28.0	--	--	6.5	--	--	--	--	--	--	33
6-11-77	28.0	1.81	81.77	6.5	1300	196	1120	16985	3321	354	33
6-11-77	28.0	--	--	6.5	--	--	--	--	--	--	34
6-11-77	28.0	1.85	82.16	6.5	1300	196	1050	20992	3531	263	33
6-11-77	28.0	--	--	6.5	--	--	--	--	--	--	33
6-11-77	28.0	--	--	6.5	--	--	--	--	--	--	33
6-11-77	28.0	1.93	82.38	6.5	1200	210	1120	16031	3493	228	33
Average	28.0	1.86	82.10	6.5	1275	198	1097	18003	3448	282	33

Gas Production						
Date	6-10-77	6-10-77	6-10-77	6-10-77	6-11-77	6-11-77
l/day	32.40	25.92	--	27.20	--	27.44
Date	6-11-77	6-11-77	6-11-77	6-11-77	6-11-77	Average
l/day	27.44	26.96	27.44	27.20	27.68	27.74

TABLE E6. STEADY STATE DATA FOR AN ANAEROBIC CONTROL DIGESTER
OPERATED AT A 10-DAY HRT, 1% T.S. FEED

Influent											
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	Tot. COD mg/ℓ	Sol. COD mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	0.35	1.0	84.51	6.7	567	52	451	12028	2608	--	--
Effluent											
3-13-77	0.35	--	--	6.5	--	--	--	--	--	--	35
3-14-77	0.35	--	--	6.5	--	--	--	--	--	--	34
3-15-77	0.35	0.97	82.47	6.5	1000	112	637	13708	1511	550	34
3-16-77	0.35	0.44	70.45	6.5	1000	108	336	6679	1527	302	34
3-17-77	0.35	--	--	6.6	--	--	--	--	--	--	35
3-18-77	0.35	--	--	6.5	--	--	--	--	--	--	34
3-19-77	0.35	--	--	6.4	--	--	--	--	--	--	35
3-20-77	0.35	--	--	6.4	--	--	--	--	--	--	34
3-21-77	0.35	--	--	6.6	--	--	--	--	--	--	35
3-22-77	0.35	--	--	6.6	--	--	--	--	--	--	34
Average	0.35	0.71	76.46	6.5	1000	110	487	10194	1519	426	34
Gas Production											
Date	3-13-77	3-14-77	3-15-77	3-16-77	3-17-77	3-18-77					
ℓ/day	0.56	0.22	0.31	0.31	0.49	0.74					
Date	3-19-77	3-20-77	3-21-77	3-22-77	Average						
ℓ/day	0.40	0.66	0.64	0.63	0.50						

TABLE E7. STEADY STATE DATA FOR AN ANAEROBIC CONTROL DIGESTER
OPERATED AT A 10-DAY HRT, 2% T.S. FEED

Influent											
Date	Flow ℓ/day	T.S. %	T.V.S. % of T.S.	pH	Alk. mg/ℓ	NH ₄ N mg/ℓ	TKN mg/ℓ	Tot. COD mg/ℓ	Sol. COD mg/ℓ	T.V.A. mg/ℓ	Temp. °C
Average	0.35	2.0	82.97	6.8	1321	207	878	22565	6068	--	--
Effluent											
4-23-77	0.35	--	--	6.9	--	--	--	--	--	--	34
4-24-77	0.35	--	--	6.9	--	--	--	--	--	--	33
4-25-77	0.35	--	--	6.8	--	--	--	--	--	--	33
4-26-77	0.35	--	--	6.8	--	--	--	--	--	--	35
4-27-77	0.35	--	--	6.8	--	--	--	--	--	--	36
4-28-77	0.35	--	--	6.8	--	--	--	--	--	--	35
4-29-77	0.35	--	--	6.8	--	--	--	--	--	--	35
4-30-77	0.35	--	--	6.8	--	--	--	--	--	--	34
5-01-77	0.35	1.61	77.07	6.9	2000	445	1218	12500	2349	296	34
5-02-77	0.35	1.35	75.56	6.9	2000	454	1057	16288	2519	454	35
Average	0.35	1.48	76.32	6.8	2000	450	1138	14394	2434	375	34

Gas Production

Date	4-23-77	4-24-77	4-25-77	4-26-77	4-27-77	4-28-77
ℓ/day	1.22	1.08	1.40	1.62	1.96	1.78
Date	4-29-77	4-30-77	5-01-77	5-02-77	Average	
ℓ/day	2.11	1.63	1.46	2.10	1.64	

APPENDIX F

MICROBIOLOGICAL ANALYSIS

TABLE F1

MICROBIOLOGICAL ANALYSIS OF ANAEROBIC FERMENTOR SAMPLES

Most Probable Number Analysis of Methanogens

Dilution Solution:

<u>Component</u>	<u>Amt/Liter</u>
Mineral Solution #1 ^a	75 ml
Mineral Solution #2 ^b	75 ml
0.10% Resazurin	1 ml
d H ₂ O	659 ml

Boil, cool rapidly and bubble with CO₂ for 10 minutes, then add:

8% Na₂CO₃ (CO₂ equilibrated) 50 ml

Continue bubbling with CO₂, and dispense as follows into CO₂ flushed serum bottles:

1. 4.3 ml per 5 ml bottle
2. 8.6 ml per 20 ml bottle

Crimp-autoclave 15 minutes at 121°C. Then 12-24 hours before use, add 1.25% cysteine-sulfide (sterile) as indicated:

1. 4.3 ml--add 0.2 ml aseptically = final volume 4.5 ml
2. 8.6 ml--add 0.4 ml aseptically = final volume 9.0 ml

Autoclave 15 minutes at 121°C

Reduce with sterile 1.25% cysteine-sulfide

a Mineral solution #1: 6 gm K_2HPO_4 /liter--autoclave

b Mineral solution #2:

<u>Component</u>	<u>Amt/Liter</u>
KH_2PO_4	6 gm
$(NH_4)_2SO_4$	6 gm
NaCl	12 gm
$MgSO_4 \cdot 7H_2O$	2.45 gm

Autoclave

Procedure:

1. Serially dilute the samples from 10^{-1} to 10^{-8} (10^{-1} dilution of the sample is ground under gas with a tissue homogenizer until the sample is evenly distributed).
2. Inoculate 3 bottles of each medium at each of 5 successive 10-fold dilutions (10^{-4} to 10^{-8}).
3. Incubate 14 days at the temperature of the reactor. Then measure methane by GC in all bottles in 0.5 ml samples, using aseptic technique. Analyze methane standards.
4. Record the number of tubes of each dilution which are positive for methane. Determine the MPN from a most probable number index table (Standard Methods, 13th ed., 1971).

Methane Production Rate Study

Procedure:

1. Preflush two 5 ml serum bottles with CO₂ and one 5 ml bottle with 50% H₂:50% CO₂. (Determine the volume of each serum bottle before the experiment and record on the bottle with a glass etcher.)
2. Add two grams of sample to each bottle.
3. Make the following additions to the bottles:

<u>Bottle</u>	<u>Addition</u>	<u>Label</u>
1	1 ml boiled & cooled d H ₂ O	no substrate
2	1 ml 0.1 M Na Acetate	+ Acetate
3	1 ml 0.1 M Na Formate	+ Formate

4. Incubate the sample at the temperature of the reactor for 3 hours.
5. Determine methane by GC analysis of 0.5 ml gas sample. Analyze methane standards.
6. Calculate $\mu\text{mole CH}_4$ per hour per gram of sample on the data sheet.

Total Count

Medium:

<u>Component</u>	<u>Amt/Liter</u>
Mineral Solution #1	40 ml
Mineral Solution #2	40 ml
1% Ammonium sulfate	76 ml
0.1% resazurin	1.0 ml
trypticase	5 gm

<u>Component, continued</u>	<u>Amt/Liter, continued</u>
yeast extract	1.0 gm
0.02% Hemin stock ^a	10 ml
volatile fatty acid stock ^b	1 ml
glucose	1 gm
soluble starch	1 gm
cellobiose	1 gm
Difco agar	20 gm
d H ₂ O	to volume
8% Na ₂ CO ₃ (CO ₂ equilibrated)	50 ml
Gas with 100% CO ₂	Dispense 9.6 ml/tube
Autoclave 15 minutes at 121°C	
Reduce with 0.4 ml sterile 1.25% cysteine-sulfide prior to inoculation	

^a 0.02% Hemin Stock

1. 0.28 g KOH in beaker
2. Add 25 ml of 95% ethanol
3. Add 10 mg Hemin
4. Bring to 100 ml

^b Volatile Fatty Acid Stock

<u>Constituent</u>	<u>Amount</u>
n-butyric acid	31.8 ml
z-methyl butyric acid	6.0 ml
n-valeric acid	6.0 ml
isovaleric acid	6.0 ml

Procedure:

1. Inoculate 1 tube each of total count medium with 0.1 ml of the 10^{-4} to 10^{-8} dilutions.
2. Prepare roll tubes and incubate at the temperature of the reactor for 14 days.
3. Count the number of colonies in each tube and calculate the total count per gram sample.

TABLE F2. MICROBIOLOGICAL ANALYSIS RAW DATA

Fermentor Type	HRT	% TS Feed	Serum Bottle Tests m.l CH ₄ /g sample/day			Methanogens x 10 ⁻⁵		Total Count x 10 ⁻⁵	Date
			No Add	Ac ^a	F ^b	No Add	H ₂ :CO ₂ -F		
Thermophilic Control	10	4	0.80	0.82	1.3	2.3	460.0	--	11/17
	10	4	0.30	0.45	0.90	--	--	--	12/7
	10	4	0.27	0.54	0.81	<0.8	<0.3	200.0	12/14
			Average		0.60				
	10	6	--	--	--	9.3	14.0	--	11/17
	10	6	0.40	0.80	1.30	2.3	0.9	600.0	12/7
	10	6	0.36	0.72	1.17	<0.3	<0.3	400.0	12/14
	10	8	1.80	2.80	2.90	46.0	75.0	--	11/17
	10	8	0.85	1.30	1.50	--	--	--	12/7
	10	8	0.54	0.72	1.35	<0.3	<0.3	600.0	12/14
	10	10	--	--	--	390.0	1,500.0	--	11/17
	10	10	1.40	2.30	2.40	11.0	0.9	700.0	12/7
	10	10	0.90	1.62	2.52	<0.3	0.4	1,200.0	12/14
	5	4	0.05	0.05	0.08	460.0	460.0	3,100.0	1/19
	5	4	0.07	0.13	0.16	0.4	110.0	1,300.0	1/21
	5	4	0.06	0.12	0.20	110.0	240.0	3,000.0	2/4
	5	4	0.12	0.30	0.13	--	--	--	2/9
	5	4	0.03	0.08	0.05	<0.3	<0.3	160.0	2/15
	5	6	0.05	0.10	0.13	110.0	110.0	3,100.0	1/19
	5	6	0.10	0.20	0.23	0.3	15.0	2,000.0	1/21
	5	6	0.08	0.20	0.42	110.0	240.0	6,000.0	2/4

^a Acetate carbon source added^b Formate carbon source added

(continued)

TABLE F2, continued

Fermentor Type	HRT	% TS Feed	Serum Bottle Tests ml CH ₄ /g sample/day			Methanogens x 10 ⁻⁵		Total Count x 10 ⁻⁵	Date
			No Add	Ac ^a	F ^b	No Add	H ₂ :CO ₂ -F		
Thermophilic Control	5	6	0.17	0.59	0.28	--	--	--	2/9
	5	6	0.12	0.30	0.27	<0.3	<0.3	280.0	2/15
	5	8	0.20	0.45	0.80	46.0	46.0	9,300.0	1/19
	5	8	0.28	0.62	0.96	0.3	15.0	1,200.0	1/21
	5	8	0.22	0.56	0.78	110.0	240.0	10,000.0	2/4
	5	8	0.36	1.26	0.78	--	--	--	2/9
	5	8	0.25	1.30	0.45	<0.3	<0.3	600.0	2/15
	5	10	0.25	0.50	1.00	0.3	39.0	13,800.0	1/19
	5	10	0.44	0.67	1.37	0.3	75.0	1,550.0	1/21
	5	10	0.36	0.84	1.48	390.0	460.0	12,000.0	2/4
	5	10	0.42	1.34	1.46	--	--	--	2/9
	5	10	0.50	1.75	1.20	<0.3	<0.3	620.0	2/15
	3-1/2	4	0	0	0	<0.3	4.3	5,900.0	3/25
	3-1/2	4	0	0	0	<0.3	3.9	6,000.0	3/29
	3-1/2	6	0	0.03	0.09	<0.3	150.0	6,500.0	3/25
	3-1/2	6	0	0	0.02	<0.3	4.3	1,000.0	3/29
	2-1/2	8	0.24	0.84	0.48	24.0	110.0	8,000.0	3/31
	2-1/2	8	0.15	0.47	0.27	110.0	1,110.0	1,200.0	4/1
						Add Acetate			
	2-1/2	10	0.40	0.54	1.20	2.3	1,110.0	12,500.0	3/25
	2-1/2	10	0.30	0.98	1.70	1.5	1,110.0	26,000.0	3/29

(continued)

TABLE F2, continued

Fermentor Type	FRT	% TS Feed	Serum Bottle Tests ml CH ₄ /g sample/day			Methanogens x 10 ⁻⁵		Total Count x 10 ⁻⁵	Date
			No Add	Ac ^a	F ^b	No Add	H ₂ :CO ₂ -F		
Thermophilic Control	20	6	0.02	0.02	0.06	23.0	210.0	1,900.0	7/8
	20	6	0.02	0.06	0.06	43.0	240.0	1,500.0	7/12
	20	6	0.02	0.08	0.06	--	--	--	7/14
	30	10	0.10	0.20	0.46	93.0	150.0	5,600.0	7/8
	30	10	0.12	0.76	0.23	93.0	460.0	3,400.0	7/12
	30	10	0.15	0.92	0.32	--	--	--	7/14
	30	8	0.05	0.09	0.28	110.0	460.0	2,200.0	1/14
	30	8	0.03	0.23	0.61	7.5	240.0	4,700.0	3/31
	30	8	0.03	0.06	0.36	24.0	24.0	1,000.0	4/1
Pilot-Scale Plug Flow	30	8	0	0.17	0.74	2.1	110.0	4,900.0	4/6
	30	8	0.03	0.57	1.07	46.0	210.0	4,000.0	4/8
	30	8	--	0.36	0.69	24.0	2.3	6,200.0	4/14
	30	12	0.06	0.32	0.86	--	--	--	7/19
	30	12	0.06	0.44	0.72	--	--	--	7/19
	30	12	0.056	0.28	0.64	--	--	--	7/19
Section Z1C	30	12	0.058	0.30	0.66	--	--	--	7/19
Effluent	30	12	0.06	0.24	0.80	--	--	--	7/19
Effluent	30	12	0.06	0.35	1.11	--	--	--	7/21
Control, 35°C	30	12	0.08	0.32	0.93	--	--	--	7/21
Bench Plug Flow	10	8	0.24	0.42	0.49	46.0	46.0	1,500.0	1/7
	10	8	0.56	0.75	1.40	110.0	110.0	1,500.0	1/12

(continued)

TABLE F2, continued

Fermentor Type	HRT	% TS Feed	Serum Bottle Tests ml CH ₄ /g sample/day			Methanogens x 10 ⁻⁵		Total Count x 10 ⁻⁵	Date	
			No Add	Ac ^a	F ^b	No Add	H ₂ :CO ₂ -F			
4 CSTR's in Series 35°C	12	8	0.04	0.05	0.14	<0.3	24.0	1,100.0	3/25	
	12	8	0.09	0.16	0.26	<0.3	46.0	1,000.0	3/29	
	10	8	0.10	0.18	0.40	40.0	110.0	7,800.0	4/27	
	10	8	0.05	0.12	0.37	110.0	110.0	5,100.0	4/29	
	with Recycle	10	0.05	0.10	0.54	110.0	110.0	2,800.0	6/9	
	with Recycle	10	0.04	0.08	0.94	110.0	110.0	2,000.0	6/10	
	4 CSTR's in Series 22°C	30	8	0	0.04	0.08	230.0	46.0	1,800.0	6/2
		30	8	0.02	0.04	0.06	21.0	46.0	1,300.0	6/7
		30	8	0	0	0	46.0	4.3	150.0	6/9
	4 CSTR's in Series 60°C			- - - N O D A T A - - -						
Controls:	22°C	12	0.07	0.43	0.53	<0.3	24.0	4,000.0	4/14	
	35°C	12	0	0.16	0.51	4.3	4.3	4,600.0	4/14	
	22°C	30	0.06	0.07	0.18	240.0	21.0	3,000.0	6/7	
	35°C	10	0.07	0.10	0.17	110.0	110.0	3,000.0	6/9	
Fluidized Bed 37°C		2	0	0	0	7.5	110.0	1,500.0	6/10	
		2	0.11	0.22	1.12	2.0	1,100.0	3,300.0	6/11	
Liquid-Solid Separation 60°C	5	5	0.06	0.21	0.68	240.0	46.0	5,500.0	6/2	
Feed Material Alone: 35°C			0.03	0.03	0.03	24.0	110.0	--	3/17	
			0.08	0.16	0.34	<0.3	110.0	800.0	3/31	

TABLE F3. SUMMARY OF MICROBIOLOGICAL AND ANAEROBIC FERMENTOR PARAMETER DATA FROM THE BENCH SCALE THERMOPHILIC SEMICONTINUOUS CONTROL FERMENTORS

Date	HRT Days	Influent		Effluent TVS g/l	TVS Destruct. Rate gm/l-day	CH ₄ Gas Prod. Rate l/l-day	Loading lbs vs/ft ³ -day	Serum Bottle Test		Total Count x10 ⁻⁵
		TS g/l	TVS g/l					ml CH ₄ /g-day		
								Ac	F	
11/17	10	40	35.6	32.4	.32	.50	.22	.82	1.3	
12/7	10	40	35.6	26.3	.93	.63	.22	.45	.90	
12/14	10	40	35.6	24.8	1.08	.42	.22	.54	.81	200
12/7	10	60	53.4	38.0	1.54	.79	.33	.80	1.30	600
12/14	10	60	53.4	37.0	1.64	.53	.33	.72	1.17	400
11/17	10	80	71.1	49.8	2.13		.44	2.8	2.9	
12/7	10	80	71.1	53.8	1.73	.75	.44	1.3	1.5	
12/14	10	80	71.1	49.2	2.19	.63	.44	.72	1.35	600
12/7	10	100	88.9	65.9	2.30	1.03	.55	2.3	2.4	700
12/14	10	100	88.9	61.8	2.71	.90	.55	1.62	2.52	1,200
1/19	5	40	35.5	27.1	1.68	.77	.44	.05	.08	3,100
1/21	5	40	35.5	29.0	1.30	.67	.44	.13	.16	1,300
2/4	5	40	35.5			.61	.44	.12	.20	3,000
2/9	5	40	35.5	28.8	1.34	.52	.44	.30	.13	
2/15	5	40	35.5	29.5	1.20	.57	.44	.08	.05	160
1/19	5	60	53.2	44.5	1.74	1.22	.66	.10	.13	3,100
1/21	5	60	53.2	39.9	2.66	1.20	.66	.20	.23	2,000
2/4	5	60	53.2	36.9	3.26	1.08	.66	.20	.42	6,000
2/9	5	60	53.2	41.7	2.30	.89	.66	.59	.28	
2/15	5	60	53.2	43.0	2.04			.30	.27	280
1/19	5	80	71.0	55.2	3.16	1.73	.89	.45	.80	9,300
1/21	5	80	71.0	51.4	3.92	1.55	.89	.62	.96	1,200
2/4	5	80	71.0	56.9	2.82	1.51	.89	.56	.78	10,000
2/9	5	80	71.0	53.4	3.62	1.23	.89	1.26	.78	
2/15	5	80	71.0	54.9	3.22	1.65	.89	1.30	.45	600
1/19	5	100	88.7	63.0	5.14	2.32	1.11	.50	1.0	13,800
1/21	5	100	88.7	62.5	5.24	2.01	1.11	.67	1.37	1,550
2/4	5	100	88.7			1.81	1.11	.84	1.48	12,000
2/9	5	100	88.7	62.4	5.26	1.54	1.11	1.34	1.46	
2/15	6	100	88.7	65.1	4.72	2.14	1.11	1.75	1.20	620
3/25	3.33	40	35.6	31.7	1.17		.66	0	0	5,900
3/29	3.33	40	35.6	32.5	.93		.66	0	0	6,000
3/25	3.33	60	53.4	47.9	1.65	.31	1.00	.03	.09	6,500
3/29	3.33	60	53.4	50.4	.90	.21	1.00	0	.02	1,000
3/31	2.50	80	71.1	53.2	7.16	2.30	1.77	.84	.48	8,000
4/1	2.50	80	71.1				1.77	.47	.27	1,200
3/25	2.50	100	88.9	72.2	6.68	2.69	2.22	.54	1.2	12,500
3/29	2.50	100	88.9	64.9	9.6	3.24	2.22	.98	1.7	26,000
7/8	20	60	53.3			.63	.17	.02	.06	1,900
7/12	20	60	53.3			.54	.17	.06	.06	1,500
7/14	20	60	53.3			.80	.17	.08	.06	
7/8	30	100	88.9			.67	.66	.20	.46	5,600
7/12	30	100	88.9			.59	.66	.76	.23	3,400
7/14	30	100	88.9			.79	.66	.92	.32	

GLOSSARY OF TERMS

- Alkalinity -** A measure of a fluid's buffering capacity or capacity to neutralize acids.
- Anaerobic Fermentation -** A bacteria - mediated process whereby organic matter is metabolized in an oxygen-free environment with a net release of energy, resulting in end products of carbon dioxide and methane gas.
- Ash Content -** All matter that remains after oxidation in an oven at 600°C.
- Attached Film Expanded Bed Reactor (AFEB) -** An upflow partially expanded filter bed composed of small inert particles that are coated with a film of microorganisms capable of metabolizing organic materials as the feedstock is passed through the reactor.
- Batch Reactor -** A stirred or unstirred reactor which has been initially filled with feed material, seeded with ample culture of microorganisms and allowed to operate without feeding for a long period of time.
- Bench Scale Reactor -** A reactor of any design and feeding regime (batch, semi-continuous, etc.) which is conveniently sized for laboratory studies.
- Biodegradable Volatile Solids -** Those volatile solids which have been stabilized by microorganisms after an infinite amount of time.
- Capillary Suction Time (CST) -** A test that determines the rate at which capillary water can be removed from a mixture of liquids and solids.
- Continuous Stirred Tank Reactor (CSTR) -** A continuously fed well-mixed reactor in which the influent stream, upon entering the fermentor, instantaneously takes on the physical and concentration characteristics of the reactor

	contents and from which the effluent stream has the same composition as the reactor contents.
Empirical Modeling Approach to Biological Processes -	An approach which assumes the rate of substrate removal to be proportional only to the substrate concentration surrounding the biomass, hence a first order mass transfer kinetic model.
Food Value -	A number composed of weighted protein and carbohydrate values incorporated into an index that has meaning only on a comparative basis.
Mesophilic Fermentation -	Fermentation conducted within the temperature range of 20°C to 35°C.
Most Probable Number (MPN) -	A bacteria enumeration test in which a sample is successively diluted and placed in tubes of enrichment broth. By observing which and how many tubes support bacterial growth, the MPN can be calculated from a statistical table.
Rational Fundamental Modeling Approach to Biological Processes -	An approach which is based upon pure bacterial culture enzyme kinetics and assumes that the microbial growth rate is a function of substrate concentration in contact with the microorganisms and the concentration of microorganisms in the reactor.
Semi-Continuous Fed Reactor -	A well-mixed reactor which is fed periodically but allowed to ferment as a batch reactor between feedings.
Thermophilic Fermentation -	Fermentation conducted in the temperature range of 50°C to 60°C.
Thermophilic Liquid Solid Separation Reactor -	A reactor that accomplishes partial phase separation of solids from the liquid due to a flotation phenomenon that occurs when gaseous bubbles attach to solid particles within the reactor.
Total Solids -	All matter that remains as residue upon evaporation of a sample at 103°C to 105°C.
Total Viable Count -	A test by which to determine the number of cells in a bacterial population capable of dividing.

Unmixed Horizontal
Displacement Reactor (UHD) - A elongated unmixed reactor which, in theory, performs by having axially uniform fluid segments flow through the system with no longitudinal mixing.

Volatile Solids - All matter which will oxidize and be driven off as gas at 600°C.