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ANALYSIS OF DIGESTER DESIGN CONCEPTS

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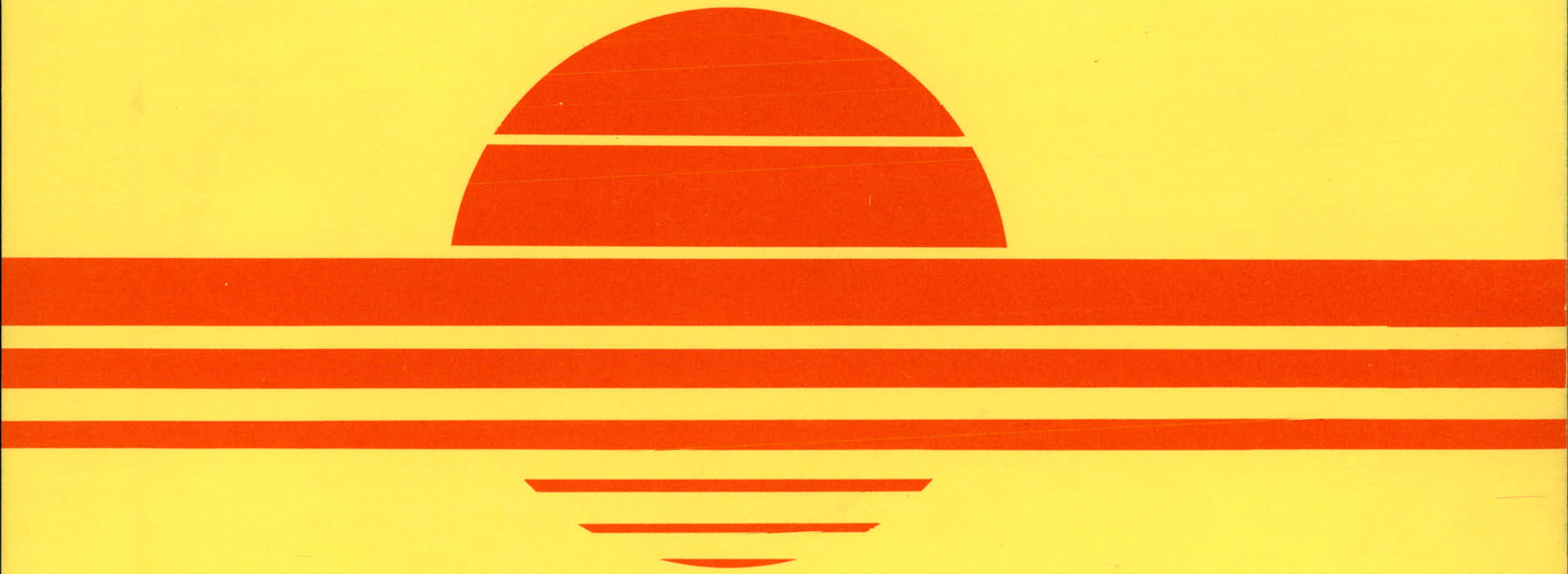
By
Edward Ashare
Elizabeth H. Wilson

January 29, 1979

Work Performed Under Contract No. EY-76-C-02-2991

Dynatech R/D Company
Cambridge, Massachusetts



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ANALYSIS OF DIGESTER DESIGN CONCEPTS

- Engineering Report -

Prepared by:
Edward Ashare
Elizabeth H. Wilson

Dynatech Report No. 1855

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ABSTRACT

Engineering economic analyses were performed on various digester design concepts to determine the relative performance for various biomass feedstocks. A comprehensive literature survey describing the state-of-the-art of the various digestion designs is included. The digester designs included in the analyses are CSTR, plug flow, batch, CSTR in series, multi-stage digestion and biomethanation. Other process options investigated included pretreatment processes such as shredding, degritting, and chemical pretreatment, and post-digestion processes, such as dewatering and gas purification. The biomass sources considered include feedlot manure, rice straw, and bagasse.

The results of the analysis indicate that the most economical (on a unit gas cost basis) digester design concept is the plug flow reactor. This conclusion results from this system providing a high gas production rate combined with a low capital "hole-in-the-ground" digester design concept. The costs determined in this analysis do not include any credits or penalties for feedstock or by-products, but present the costs only for conversion of biomass to methane. The batch land-fill type digester design was shown to have a unit gas cost comparable to that for a conventional stirred tank digester, with the potential of reducing the cost if a land-fill site were available for a lower cost per unit volume.

The use of chemical pretreatment resulted in a higher unit gas cost, primarily due to the cost of pretreatment chemical. A sensitivity analysis indicated that the use of chemical pretreatment could improve the economics provided a process could be developed which utilized either less pretreatment chemical or a less costly chemical.

The use of other process options resulted in higher unit gas costs. These options should only be used when necessary for proper process performance, or to result in production of a valuable by-product.

TABLE OF CONTENTS

TABLE OF CONTENTS (Continued)

TABLE OF CONTENTS (Continued)

TABLE OF CONTENTS (Concluded)

LIST OF FIGURES

LIST OF FIGURES (Continued)

LIST OF FIGURES (Concluded)

LIST OF TABLES

Section 1

INTRODUCTION

As a result of the energy crisis, a considerable effort has been expended over the last few years to develop alternate renewable sources of energy. One such source is biomass, which can be considered as solar energy being collected and stored by plants. Biomass can be grown specifically for use as an energy crop, or it can be obtained as residues, both from crops and animals. Studies have been undertaken by various organizations under contract with the U.S. Department of Energy (and its predecesosr, ERDA) to investigate the feasibility of various biomass sources as potential alternate renewable energy supplies. These include grains and grasses, silviculture, aquatic plants, animal residues, and other crops such as corn and sugar cane.

Once there is an available source of biomass, the next step is to convert the biomass to a more easily utilized form of energy. Both thermochemical and biological processes are being considered as conversion processes for production of liquid and/or gaseous fuels. One such process is the biological anaerobic digestion of biomass which is used for production of methane gas.

There have been many approaches suggested for operation of an anaerobic digester, some of which are still in the laboratory or small pilot-scale stage. It is the purpose of this report to present an engineering economic analysis of some of the conceptual designs for the anaerobic digestion of biomass, including the advanced as well as the new technology. Table 1.1 presents some of these processes, with an indication of some of the advantages of each.

A comprehensive literature survey of the many aspects of the overall digestion process, including pretreatment, digestion, and post-treatment, was performed. This is presented in Section 2 of the report. In Section 3 is presented the procedure utilized for the engineering economic analysis of the

Table 1.1

ECONOMIC ANALYSIS AND ASSESSMENT OF ANAEROBIC DIGESTER DESIGN CONCEPTS FOR BIOMASS CONVERSION

Digester Concept	Pretreatment (if integral to the digestion concept)	Remarks
Landfill	None	As practiced at Palos Verdes, City of Mountain View, and other sites, methane is withdrawn from an existing landfill of solid waste. This makes this digestion concept the "most simple" for biomass. Note that no pretreatment or admixtures of inoculum/buffer/nutrients are implied.
Controlled Landfill	Admixture of inoculum, nutrient, and buffer essential to process functioning. Pretreatment alternatives: a) None b) Shredding c) Shredding plus mild alkali	This concept implies admixing of the biomass source with nutrients, inoculum and buffer. Note that shredding and alkali pretreatment are additions to the digestion concept, i.e., they are not required, although they may enhance the rate.
Multi-Stage Digestion	Shredding/comminution to some degree appears essential. Chemical pretreatment is not required in that the first digestion "stage" is a microbially induced (acetic acid) treatment.	This system, the separation of acid formers and methane formers, has received the attention of a number of workers. CNG, for example, funded a multi-stage plug flow front end digestion system followed by a CSTR back end. IGT and University of Pennsylvania have also investigated this digestion concept.
"Channel Digester" (addressed by Biogas of Colorado, Ludington, MI group and Cornell University)	Dilution of the biomass and a substantial degree of comminution appears to be required. Mild alkali treatment may be included.	This appears to be a standard type of innovation for groups seeking low capital cost biomass digestion systems.
Packed Bed Digester	Minimal dilution mild alkali pretreatment may be an advantage.	Packed bed concept implies that the digester liquid is regularly drained and recycled for the conversion to occur, rather than mix the solids. Thus higher solids conversion is possible. <u>Note:</u> Development work to make the process continuous will be required.
Conventional CSTR Digestion (Traditional to Sewage Treatment)	Pretreatment alternatives: a) Comminution required b) Comminution and mild alkali treatment	Base-line for comparison. Prof. Pfeffer, for example, is accumulating a data base for biomass conversion.
"McCarty Process"	High temperature/high pressure alkali pretreatment developed recently by McCarty/Gossett, et al.	Organic compounds from pretreatment to be digested in a McCarty type anaerobic filter (extensions of process by Jeris, Converse, and Jewell).
"Porteous Saeman Process"	Acid hydrolysis of largely cellulosic material to wood sugars carried out. Subsequent conversion to ethyl alcohol practiced, but higher conversion efficiency to CH_4 may be expected.	Process used during WWII to produce ethyl alcohol. Initiated again by Porteous at Dartmouth for use with solid waste. Conversion of all products of acid hydrolysis to CH_4 is the objective (for ethyl alcohol production, only the cellulose fraction is utilized).
"Kukharenko Process"	Pretreatment similar to McCarty but other chemicals selected (K_2CO_3). Conversion to benzene-carboxylic acids is target goal for lignaceous fraction.	CSTR digester or anaerobic filter follows the pretreatment step. Suitable for other than annually harvested biomass sources.
Gasification plus Biomethanation	Gasification of total carbon in biomass converted to CO, CO_2 and H_2. Thus, the total lignin and refractory organic matter is utilized.	Anaerobic fermentation at high pressures of the gases CO, CO_2, and H_2 to CH_4 takes place. Perhaps the highest conversion rates possible with this system.

various conceptual designs. The process conditions for the analysis are presented in Section 4. The results of the analysis are presented in Section 5. This section includes comparisons of the various processes for several different types of biomass. An economic sensitivity analysis to determine the effects of some important processing variables was also performed. This analysis has led to conclusions and recommendations regarding the anaerobic digestion of biomass to methane programs. These are presented in Section 6. Appendices are included to present details of the analysis procedure as well as the numerical results of the analysis.

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Section 2

ANAEROBIC DIGESTION TECHNOLOGY

2.1 Introduction

The decomposition of organic matter and subsequent formation of methane by microorganisms is indigenous to nature. In fact, anaerobic microorganisms have functioned to digest vegetative foodstuffs in the alimentary tracts of herbivorous animals throughout evolutionary history. Anaerobic microbial metabolism may take place whenever the ingress of oxygen is stopped completely or limited to such an extent that microbes remove the oxygen. For the production of methane to occur however, there must be growth of methanogenic bacteria, which require highly specific environmental conditions. An understanding of biological processes occurring in nature coupled with numerous achievements in sanitary and chemical engineering have made methane production from biomass economically viable.

Traditionally, anaerobic digestion has been used as a treatment process for industrial and municipal wastes. Fundamental knowledge of the process has grown out of the waste treatment industry. Advancement in anaerobic digestion technology has led to an adaptation of the concept for the production of methane from biomass. A flow diagram illustrating the general system for anaerobic digestion of biomass is shown in Figure 1. The digestion process is preceded by pretreatment and followed by gas treatment and solids disposal.

The extensive list of anaerobic digestion processes discussed in Section 2.2 emphasize the numerous alternative anaerobic digestion processes. Selection of the proper process is dependent on the quantity of feedstock and on the composition of the substrate, in particular the moisture content and flow of material through the system. The processes also differ in complexity and efficiency.

Pretreatment of the feedstock has proved to be valuable in accelerating the anaerobic digestion process by increasing degradability of the substrate. Section 2.3 discusses alternative pretreatment processses which assist in the fractionation and decomposition of various substances contained within biomass feedstocks. Some of these pretreatment processes have not yet been utilized specifically for biomass bioconversion, but are applicable to the technology.

Utilization and disposal of solids remaining after anaerobic decomposition are discussed in Section 2.4. Treatment of gases released from the system to meet the standards of pipeline quality natural gas,as discussed in Section 2.5, involves removal of carbon dioxide, hydrogen sulfide and moisture.

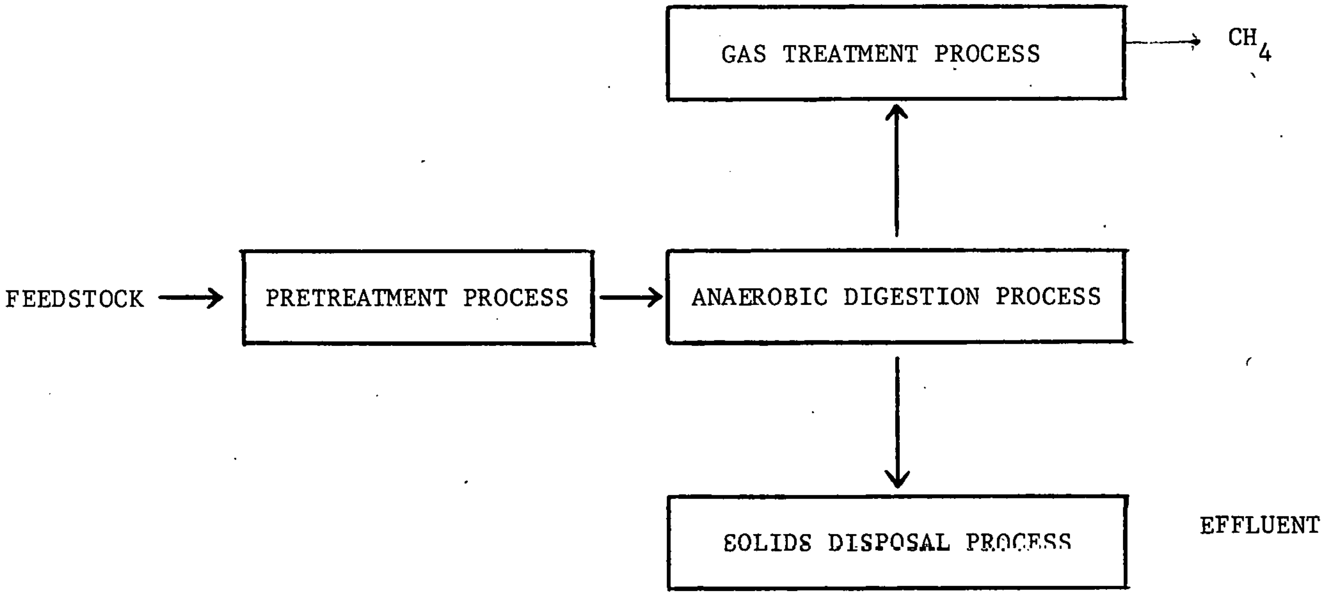

Fig. 1 A Flow Diagram Illustrating the System for Anaerobic Digestion of Biomass

2.2 Anaerobic Digestion Processes

2.2.1 Single-stage Processes

Lagoons

The Italian physicist, Volta, is often given credit for first observing the formation of a combustible gas in 1776 (Refs. 1, 2) and relating it to the quantity of plant material in bottom sediments, ponds and streams. Lagooning, which is the simplest type of anaerobic digestion process used for waste-treatment of high moisture materials, utilizes the same phenomenon that Volta observed in nature. A lagoon is merely a pond lined with concrete or other water proof material and open to the atmosphere. Waste material moves slowly into the lagoon so that solids settle out and partially purified liquids leave the system. Aerobic metabolism takes place in the top layer of the liquid, but in the liquids and solids below, anaerobic microorganisms break down and stabilize organic matter.

Lagooning is a common practice for stabilization of sewage sludge as a secondary treatment in rural areas. Although they serve only about 7 percent of the population, there are approximately 3500 lagoon installations in the United States with 90 percent in communities with populations of less than 10,000 (Ref. 3). These lagoons are termed facultative stabilization ponds since their operation involves both aerobic and anaerobic decomposition of organic matter. As depicted in Figure 2, waste organics in suspension are broken down by bacteria releasing inorganic nitrogen, phosphorus and carbon dioxide, which are used by algae for growth. Dissolved oxygen released by the algae and from the atmosphere is in turn taken up by the bacteria. Anaerobic decomposition of solids on the bottom of the lagoon yields inorganic nutrients, hydrogen sulfide and inorganic acids. These lagoons are usually operated at a maximum depth of 5 ft, so as a result, the anaerobic cycle is limited and methane gas production is minimal.

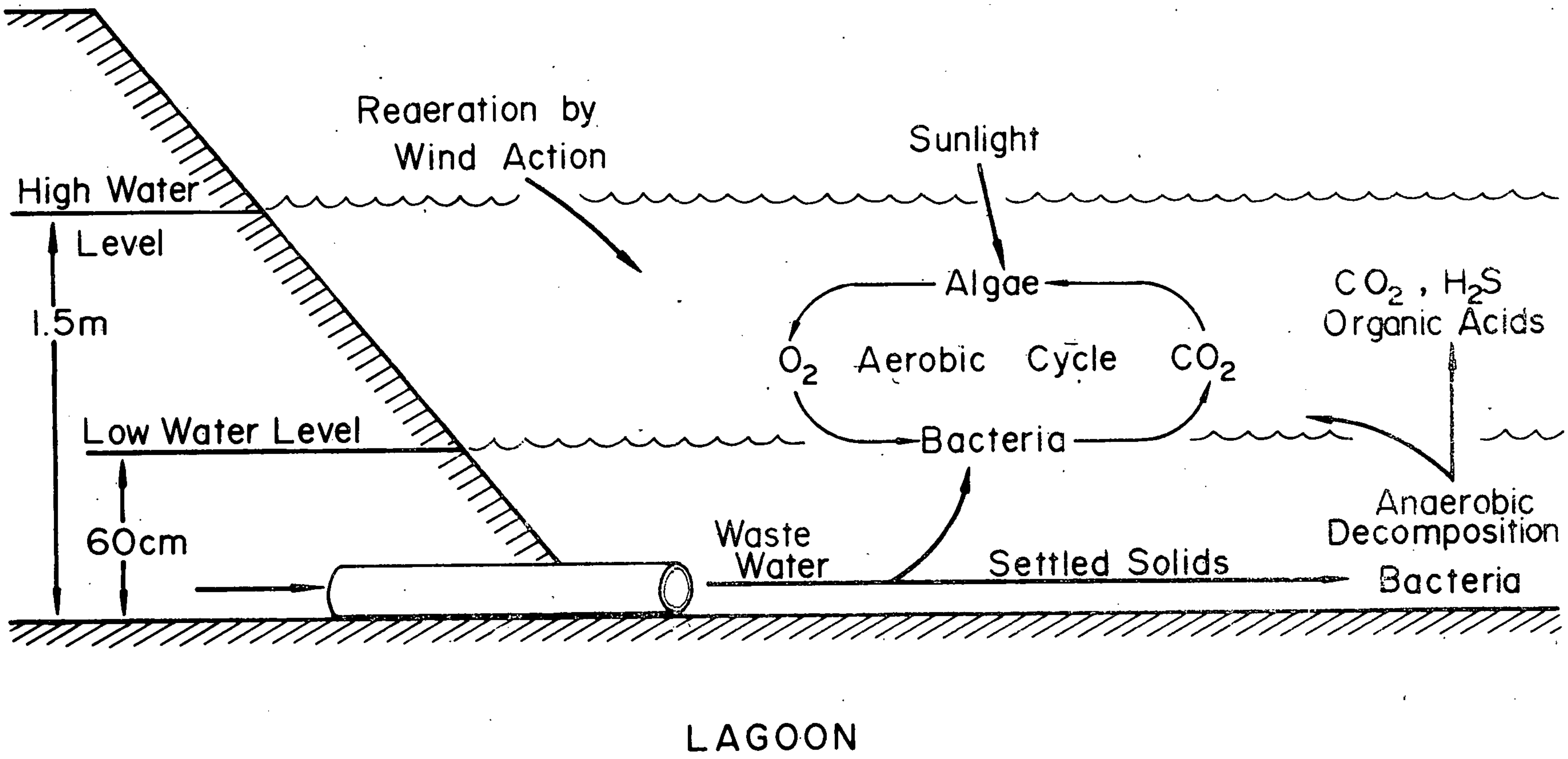

Figure 2 Schematic diagram of a facultative stabilization pond showing basic biological reactions of bacteria and algae.

With the growth of intensive livestock and poultry production in the past 30 years it has become impossible to continue the practice of recycling animal wastes to pasturelands. In a review of alternative systems for the treatment and disposal of animal residues (Ref. 4), Loehr suggested a lagooning system involving water flushing followed by anaerobic treatment followed by aerobic treatment and land disposal. The aerobic treatment process is recommended since the anaerobic lagoons in common use do not reduce odor or purify the manure adequately for immediate disposal.

Treatment of swine wastes by lagooning has been practiced since about 1959 (Ref. 5). Similar to facultative stabilization ponds, the first swine waste lagoons were maintained at a depth of 0.9 - 1.5 m (3 - 5 ft) and the loading rates were based on square feet of lagoon surface per animal (Ref. 6). Studies conducted using model lagoons, however, indicated that depths deeper than 1.5 m (5 ft) with consistent daily loadings at a rate of 5 cu m/hog (180 cu ft/hog) were more efficient (Ref. 7). Miner (Ref. 8) recommended a length to width ratio of 3 to 1 or less or as deep as economically possible without the bottom of the lagoon reaching below the water table. He also recommended that the waste enter the middle of the lagoon through submerged inlets if the flow were large enough to keep the inlet from becoming clogged by solids. The outlet should remove the liquid from near the surface but should be beneath the scum layer.

One of the major problems with animal waste disposal in lagoons is odor production and considerable research has been conducted for its control. The odorous gases from manure under anaerobic conditions are composed of hydrogen sulfide, methyl and ethyl mercaptan, indole, skatole and amines (Refs. 6, 9). Gumerman and Carlson (Ref. 9) found that by passing air containing hydrogen sulfide through columns of soil, dry soil removed high concentrations of hydrogen sulfide more efficiently than did wet soil. They advocated that small lagoons should be covered by a framework roof covered by polyethylene to catch the odorous gas and divert it into the soil column. The large lagoons, they contended, could use floating polystyrene foam covers with soil filters located on a grid pattern throughout the cover.

A covered lagoon has been developed for the treatment of meat-processing wastes (Ref. 3). The cover is not a synthetic material, however, but is made of 500 mg/ℓ of grease which normally occurs in the waste. A schematic diagram of a first-stage anaerobic lagoon for treatment of slaughterhouse waste water in rural locations is shown in Figure 3. Minimum pretreatment of the raw wastes includes: blood recovery for a salable by-product, screening to remove coarse solids (paunch manure), and skimming to reduce the grease content. The lagoon is constructed with steep sides and a depth of 4.6 m (15 ft) to minimize the surface area relative to total volume. This construction allows several inches of grease accumulation to form a natural cover for retaining heat, suppressing odors and maintaining anaerobic conditions. Operation of lagoons in a series is not recommended because it is difficult to maintain an adequate grease cover in the second stage lagoon. Influent waste water enters near the bottom so that it mixes with active microbial solids in the sludge blanket. The discharge pipe is located on the opposite end and is submerged below the grease cover. Upward flow of the discharge allows settling of the bacterial floc so that the anaerobic mixed liquor is retained in the lagoon. Sludge recirculation is not necessary, since gasification and the inlet-outlet flow pattern provides adequate mixing. The normal operating standards to achieve a BOD removal efficiency of 75 percent are a loading of 324 g BOD/cu m /day (20 lb per 1000 cu ft per day), a minimum detention time of four days, and a minimum temperature of 24°C (75°F). This lagoon system has proved highly successful for the treatment of meat processing wastes.

Landfills

Disposal of municipal solid refuse has traditionally been accomplished in sanitary landfills. Recently interest in this process has extended to the production of methane from biomass with a high solids content, such as food processing wastes and agricultural residues.

The first formalized description of the sanitary landfill method of disposal was in the form of an ASCE Committee report published in 1959. The operational procedures drawn up in the report are summarized in the ASCE definition

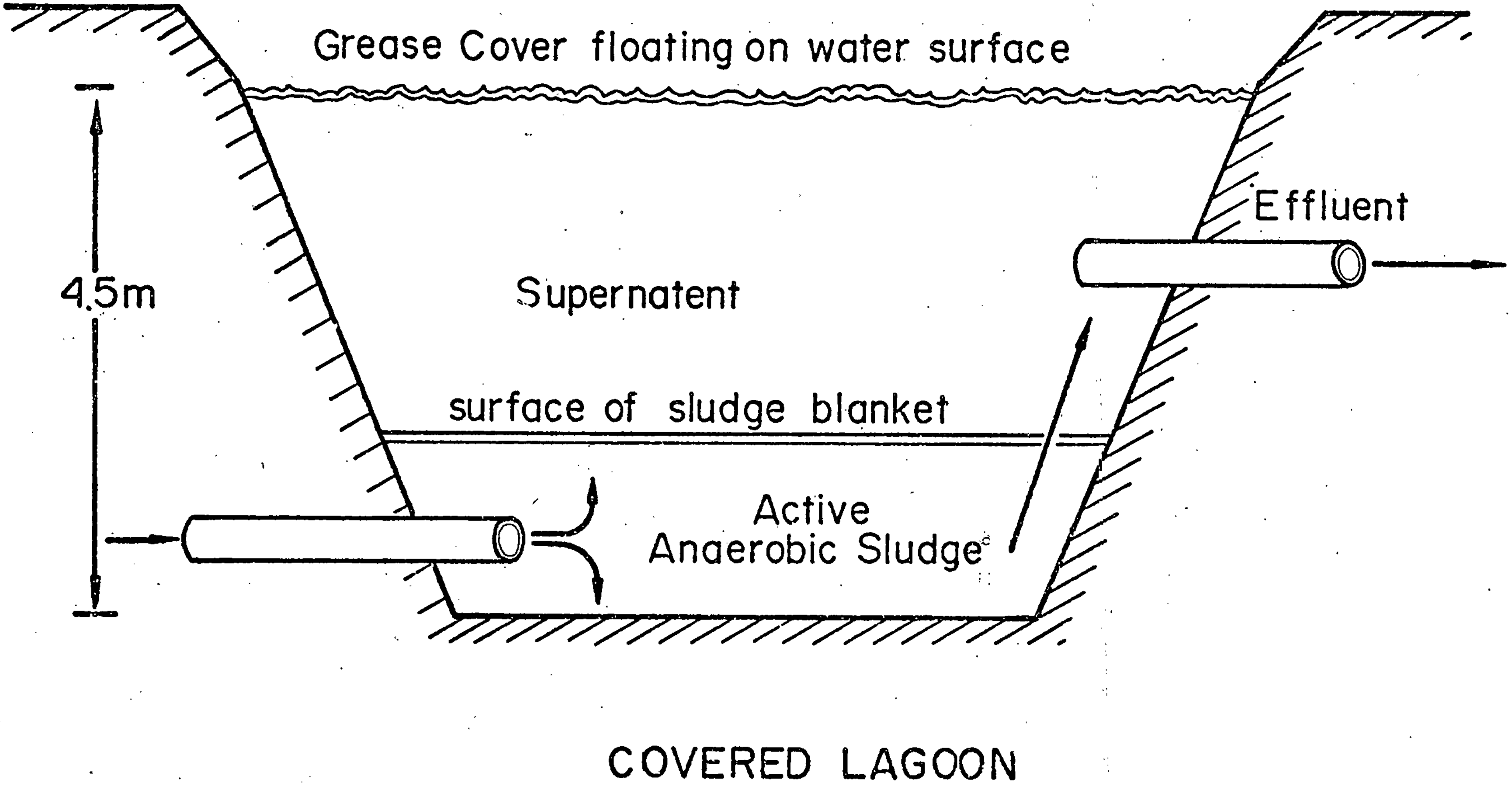

Figure 3 Schematic diagram of an anaerobic lagoon for treating meat-processing wastewater.

of a sanitary landfill, which is: "Sanitary landfilling is a method of disposing refuse on land without creating nuisances or hazards to public health or safety, by utilizing the principles of engineering to confine the refuse to the smallest practical volume, and to cover it with a layer of earth at the conclusion of each day's operation or at much more frequent intervals than may be necessary." (Ref. 10). The guidelines were geared toward aesthetics and did not take into consideration the possibility of pollution of the surrounding atmosphere and groundwater.

As in lagoons, anaerobic conditions exist in all but the top 3 m of a landfill resulting in the production of gases, and leachates. Lack of control of these contaminants has resulted in fires (Refs. 11, 12), suffocation (Ref. 13), ground water degradation affecting acidity, alkalinity and hardness (Refs. 14, 15), destruction of vegetation (Ref. 16), and discharge of micro-organisms (Ref. 15). Consequently a great deal of research on the production of gases and leachates from decomposing refuse has been carried out with the objective of identifying methods for their control.

Investigations on the formation of gases in landfills began in the late 1930's. A gas sampling device which prevented contamination with air was developed for use in New York landfills. The rigidly constructed device was driven into the refuse with a mallet and gases were extracted for further study (Ref. 17). One of the first reported "scientific" attempts to collect and identify gases from several depths within a landfill was made in 1940 (Ref. 18). Since that time numerous gases have been identified in landfills including carbon dioxide, methane, hydrogen, nitrogen, carbon monoxide, oxygen and sulphuretted hydrogen (Ref. 19), but the former two usually make up about 90% of the total gas produced (Ref. 20) after a steady state is reached.

A pattern of biogas production in relation to the age of a landfill has been categorized into four phases (Refs. 21, 22). The initial aerobic phase is short, leading into the second phase of high carbon dioxide production at approximately a molar equivalent to the oxygen consumed so that little nitrogen is displaced. In the anaerobic non-methanogenic phase, a carbon dioxide "bloom" occurs as organic acid production proceeds. These blooms which may produce as much as 90% CO_2 have been reported to occur after 11 days (70% CO_2) and 23 days (50% CO_2) (Ref. 23) and after 40 days (90% CO_2) (Ref. 24). Hydrogen production increases and nitrogen displacement usually increases dramatically. In the third phase, methanogenesis activity begins and methane concentrations increase as carbon dioxide and hydrogen levels decrease. This steady-state phase has been found to occur as early as 180 days after filling operations began (Ref. 25), 250 days (Ref. 26) or as late as 500 days (Ref. 24). The fourth steady-state phase of methane production occurs considerably later and ranges from 50 - 70% (Refs. 19, 24, 25).

Gases occurring in landfills move both horizontally and vertically at velocities of about 7 cm/day (0.23 ft/day) (Ref. 27). Movement downwards results from molecular diffusion plus density differences, whereas movement to the sides and upward to the atmosphere results from diffusion alone (Refs. 28, 29). Carbon dioxide usually moves into soil beneath a landfill if the soil is homogeneous and not impervious as well as upward; whereas methane tends to rise through the landfill and not diffuse into groundwater (Ref. 29).

Leachates consisting of solid matter and microbial waste products are released when groundwater or infiltrating surface water moves through decomposing solid wastes. Since the composition of the leachate is important in determining its potential effects on the quality of nearby surface water and groundwater, it has been the object of many laboratory and field studies (Refs. 30 to 35), resulting in the identification of significant amounts of Fe, Zn, Cl, Na, Ni, Cu and $CaCO_3$, plus many trace elements.

Some contaminants are removed naturally from leachates percolating through the soilds underlying and surrounding the solid waste, by ion exchange, filtration, adsorption, complexing, precipitation, and biodegradation. The mechanism of purification is affected by the moisture content of voids in the soil through which the leachate moves. Purification of contaminants flowing in the unsaturated zone is generally greater than in the saturated zone because there is more potential for aerobic degradation,adsorption, complexing and ion exchange of organics, inorganics, and microbes. In highly saturated areas anaerobic degradation prevails. Adsorption and ion exchange are highly dependent on the surface area of the liquid and solid interface.

The distance that contaminants travel depends on the composition of the soil, its permeability and the type of contaminant. Organic materials that are biodegradable do not travel far, but inorganic ions and refractive organics can travel appreciable distances. Some inorganic contaminants from a dump located in an abandoned gravel pit have been traced for 365.7 m (1,200 ft) (Ref. 31) through a highly porous glacial alluvium. Rates of movement can be so slow through some solids, however, that the full impact of the contaminant is not realized for many years (Ref. 36).

Since natural purification processes have only a limited ability to remove contaminants, it is necessary to construct sanitary landfills with an impermeable barrier to prevent water from entering the fill and subsequent leaching. The same impermeable barrier may also block the flow of gases. The most common method to retard diffusion downward and horizontally is to place a 46 - 122 cm (18 - 48 inch) layer of compacted clay between the refuse and adjacent soils (Ref. 37). The clay is placed as a liner during excavation or is installed as a curtain wall to block underground gas flow (Ref. 14). The liner is usually constructed during daily operation as the fill progresses since prolonged exposure to air dries the clay and causes it to shrink and crack.

If clay is not available, synthetic liners are used in solid waste disposal operations. They are usually made of butyl rubber, polyethylene, Hypalon[*] or polyvinyl chloride (Ref. 38) and are installed in multiple layers. Any of these lining materials will control the flow of leachates, but since polyvinyl chloride is less permeable by gas than polyethylene, the former material is preferable to prevent the flow of gases. The membranes are laid down carefully to avoid punctures and layers of sand are placed on both sides of them.

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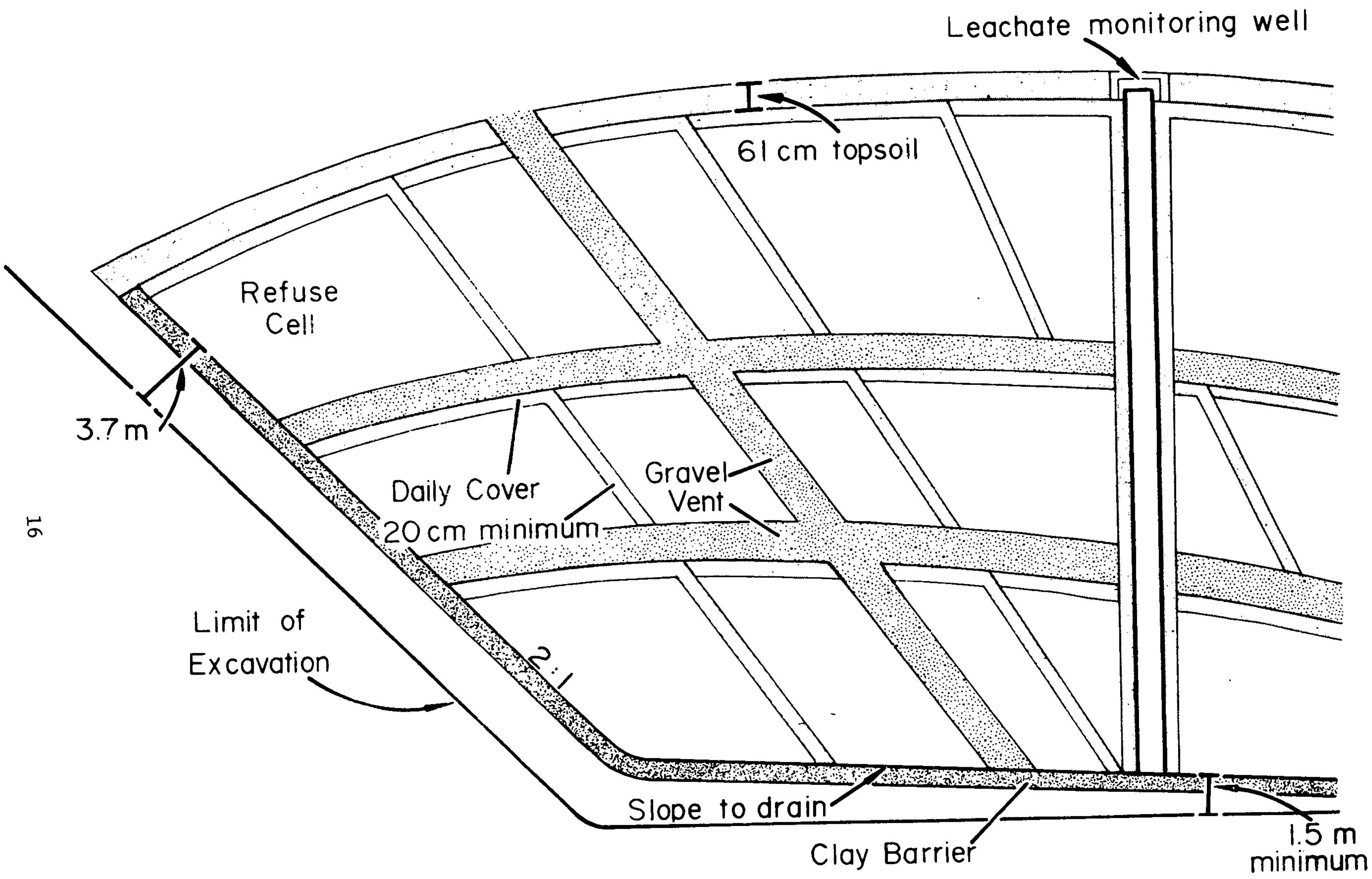
[*] Hypalon is a registered trademark of DuPont.

When impermeable barriers are used, some method for the removal of the contained fluid is usually provided. The fluid in a bowl-shaped liner can be pumped by a well or series of wells or it can exit through outlets in the bottom of the liner. When a natural ravine or canyon is involved, the removal point is at the downstream end of the filled area.

Impermeable barriers have also been used to prevent the flow of gases upward. Concrete slabs laid on top of a landfill have been used to prevent gases from getting into buildings (Ref. 39). The use of asphaltic or bituminous materials as barriers is questionable, however, since they tend to crack or rupture when the decomposed refuse settles (Ref. 40). In heavily populated areas such as New York City, buildings erected on completed landfills are protected from methane gas leakage by using gas tight construction and adequate ventilation (Ref. 41).

Permeable methods are also used to control the movement of gases. Lateral gas movement can be prevented by using a material that is more permeable than the surrounding soil, such as gravel. Wells filled with gravel are commonly used to vent gases to the atmosphere (Refs. 20, 42, 43, 44). Gravel trenches are constructed during normal daily operation after a layer of cover has been placed over the refuse. Existing landfills have been modified for gas control by inserting perforated pipes into the fill (Refs. 45, 46).

A typical completed sanitary landfill is shown in Figure 4. The bottom and sides are lined with a clay barrier to control leachate and gas movement laterally and downward. Solid waste is spread and compacted into confined areas and covered with a thin 20 cm continuous layer of soil resulting in the formation of refuse cells. The dimensions of the cell are determined by the volume of the compacted waste and this, in turn, depends on the density of the in-place solid waste. Usually the cell approaches a square shape with side slopes of 20° to 30° (Ref. 37). Gravel is laid vertically and horizontally for the escape of gases and sometimes a well is provided to monitor leachate contaminants during landfill construction and long after the landfill has been



SANITARY LANDFILL

Figure 4 A cross-sectional diagram of a typical sanitary landfill. Refuse cells are formed by compaction of the solid waste and daily cover with a layer of soil. A well is provided for monitoring leachate.

completed. Upon completion, the sanitary landfill is covered with top soil and graded after settlement. Sanitary landfills vary in depth from about 100 m in deep canyons to about 3 m in flat land areas.

In the past, gases (primarily CO_2 and CH_4) leaking from landfills were considered a nuisance and were oxidized either by burning (Refs. 43, 47, 48) or by spraying with commercial oxidizing agents (Ref. 49) to control odors. Recently, however, the sanitary landfill has been regarded as a natural factory for the production of these gases as a renewable energy resource and a number of research projects on the extraction of gases from existing landfills for commercial use are underway in the United States.

Pumping of gases from existing landfills is accomplished by the same principle that is involved in the extraction of groundwater. Operation of pumps attached to perforated pipes extending into the landfill, establishes a pressure gradient within the landfill causing the gas to flow into the pipes and at the same time, create enough negative pressure to allow atmospheric gases to penetrate the soil. If the landfill is overpumped the infiltration of atmospheric gases will augment aerobic conditions and result in a decrease in methanogenesis and an increase in CO_2 concentrations of the extracted biogas. In order to determine the most efficient operating parameters for a particular landfill, a test well and supporting wells are drilled and a series of tests are run to evaluate landfill pressure profiles, biogas composition and heating values, and stability of withdrawal rates (Refs. 22, 50, 51).

Typically a gas recovery well is sunk into a 75 - 90 cm diameter drilled shaft which is filled with clean crushed rock in the lower portion and backfilled with clay in the upper portion. The lower 80% of the well is perforated utilizing alternating 10 - 15 cm diameter PVC pipe coupled with burlap joints. Collection manifolds connect the wells, bringing the biogas to a common collection site (Ref. 22).

The economic success of gas extraction from existing landfills has led to current interest in increasing landfill gas yields. Attempts are being made to accelerate biodegradation rates of landfill wastes and hence, gas production rates, by pretreating the refuse before deposition into the landfill, and enhancing microbial metabolism during decomposition. The concept of a "controlled landfill" implies construction and daily operation of the landfill for the purpose of attaining high methane yields.

Pretreatment operations which affect the rate of gas production in controlled landfills, such as particle size reduction, classification and alkaline pretreatment will be discussed in Section 2.3. These operations are performed before deposition of the organic matter into the landfill to control physical and chemical reactions taking place within the system.

It has recently been proposed that enhancement of microbial activity can be achieved by recycling leachate through the decomposing refuse. The leachate which accumulates along an impermeable barrier, such as a clay substratum, is collected and recycled with the refuse cells. A comprehensive laboratory study of this technique began in 1971 at the Georgia Institute of Technology under the direction of Dr. Fred Pohland, for the purpose of accelerating stabilization processes within sanitary landfills and removing readily degradable pollutants from the leachate (Ref. 52). Simulated landfills constructed of corrugated steel sections of pipe 91.4 cm (36 inches) in diameter and joined to form a column 4.3 m (14 ft) deep were used for experimentation. A conical concrete bottom with a 3.8 cm (1.5 inch) drain was formed in each simulated fill to seal the bottom of the pipe section and allow for drainage of leachate. The fill receiving recirculated leachate was pumped back through a distributor unit buried between the top of the compacted shredded refuse and the soil cover and allowed to percolate through the refuse. Refuse removed from the fill after 11 weeks indicated that decomposition had proceeded at a more rapid pace in the fill receiving recycled leachate than in the control fill where the leachate was wasted. Carbon reduction and volatile solids reduction was greater in the experimental fill than in the control (Ref. 53).

Subsequent experimentation on the effects of leachate recycle plus nutrient addition and pH control on stabilization of sanitary landfills have been conducted using the same apparatus. Seeding with raw primary sewage sludge accelerated the biological stabilization processes initially with a more rapid and larger production of volatile acids and organic pollutants in the leachate, but also with an eventual reduction in stabilization time for the readily decomposed organic materials in the leachate when compared to the fill without recycle. Pretreatment with sodium hydroxide promoted more rapid development of methane formers with a concomitant increase in the rate of stabilization and removal of pollutant concentrations from the leachate (Ref. 52). Although Pohland (Ref. 52) did not measure gas production in the simulated landfill, it follows that an increase in stabilization rates would result in an increase in methane formation. In experiments on the anaerobic digestion of leachate in a completely mixed continuous flow reactor system with sludge recycle a gas production of 0.81 cu m CH_4/kg with 74.6 percent methane (17.4 cu ft/lb) at a 15 day retention time was reported (Ref. 52).

Septic Tanks

The oldest and simplest enclosed anaerobic digester is the common septic tank for treating domestic sewage. It is believed that from this type of digester the potential use of gases from the fermentation process was first realized. Gas from a "carefully designed" septic tank was used for street lighting in Exeter, England in 1895 (Ref. 54). The experience must have been successful enough to encourage others, because in the 1920's several devices were built and used in England, specifically for the purpose of generating flammable methane gas (Ref. 55). Today septic tanks are used primarily for single-house domestic sewage, although some storage pits for farm wastes of a similar design are in use.

Sewage is fed into the specially designed chamber of a septic tank by gravitational forces. The solids settle and are anaerobically metabolized. The partially purified liquids are drained off, the gases are vented to the atmosphere and the gradual accumulation of solids must be removed periodically.

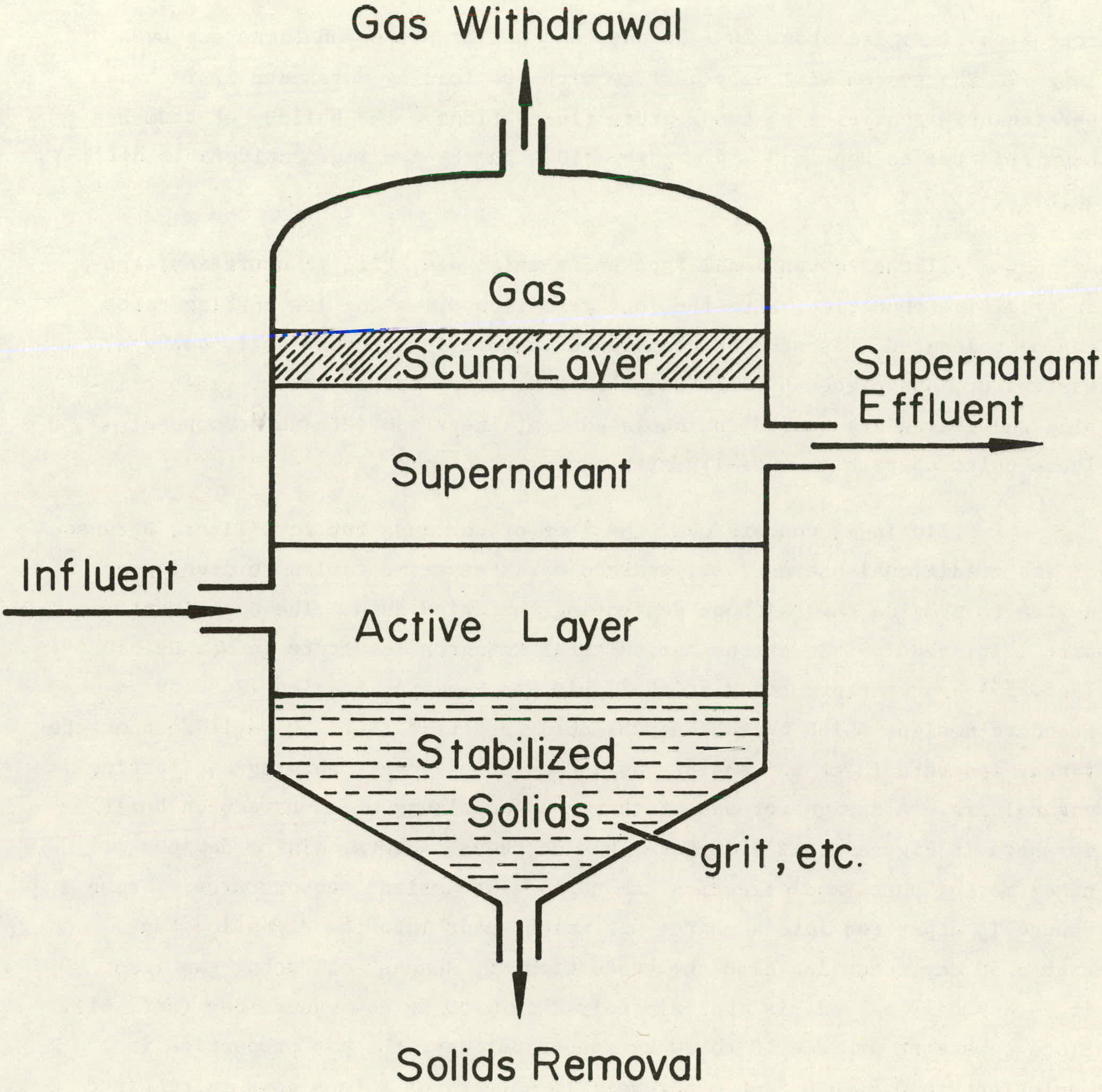
One of the major problems encountered with early septic tanks was the lifting of solids into the clarification zone by rising gas bubbles. This problem was solved by using a two-story Imhoff tank. This process also involved only one vessel; however the design was such that the sedimentation of solids from dilute wastes and the biological action of digestion were separated physically. The separation of functions in the Imhoff tank increased the quality of the effluent over that obtained in the septic tank (Ref. 56).

The septic tank principles were the basis for the design of one of the first primary digestion plants which was capable of handling sewage from a very large population and from which electricity was generated in the city of Birmingham, England in 1911 (Ref. 57). This plant is one from which the present systems of anaerobic sewage digestion grew (Ref. 58).

Conventional Single-Stage Tank Digesters

The conventional single-stage anaerobic digestion unit is merely a holding tank for wastes, in which both biological stabilization and solid-liquid separation take place. The unit is not heated or stirred, so that charged material undergoes decomposition in a quiescent state and highly liquid substrate tends to stratify (Figure 5) into a stabilized solids/grit layer, an actively digesting layer, a supernatant layer and a scum layer. Sludge enters the center of the active zone where digestion takes place and water is released to form a supernatant zone. The decomposed solids are heavier than the liquid and settle to the bottom. As gases are formed, they rise to the surface, pass through the scum layer and are removed from above. The rising gases carry lighter sludge particles to the surface above the supernatant and form a dense layer of scum. This scum layer, in time, becomes thick enough to walk on.

This type of digester is rarely built any longer because of the long retention times and inefficiency of the system. Quiescent conditions for satisfactory gravitational solids separation will not allow adequate contact between the substrate and microorganisms for optimum biological action. Liquid



CONVENTIONAL SINGLE-STAGE DIGESTER

Figure 5 The stratification of highly liquid wastes within a conventional single stage anaerobic digestion unit. Standard operation characteristics:

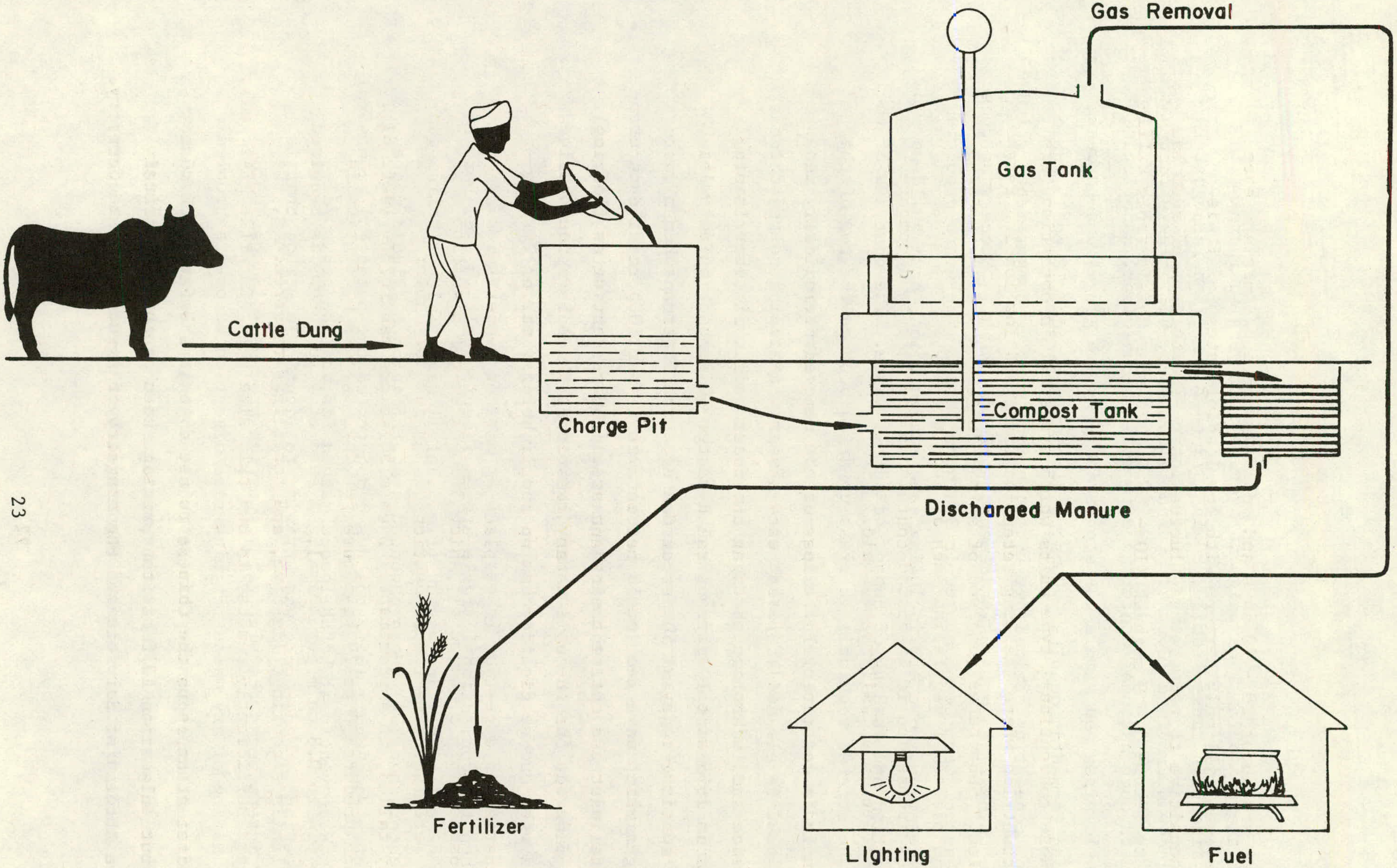
- Unheated
- Loading rate 0.64 - 1.60 kg volatile solids/m^3/day
- Detention time 30 - 60 days
- Intermittent feeding and withdrawal
- Stratification

retention times are about 20 - 30 days and solids retention times are even longer. The system will only perform with low loading rates and it is particularly sensitive to temperature fluctuations. The buildup of scum has been reported to become 3 - 9 m thick (10 - 30 ft) causing considerable difficulties.

Those conventional type units which are still in operation, are in primitive countries where the long retention times and low loading rates can be tolerated. Substrates composed of animal dung, night soil, garbage, wastewater, and pieces of vegetation such as crop stalks, straw, grass clipping and leaves are sealed in insulated containers and left to decompose. These units operate on a family-size scale.

In India concern over the loss of cow dung for fertilizer, because of its traditional use as fuel, sparked early experimentation to develop a system to provide fuel without destroying the dried dung. The experiments were initiated in 1939 at the Agricultural Research Institute in New Delhi (Ref. 55) and continue today at the Gobar Gas Research Station (Ref. 59). Standard designs which have been published in a text (Ref. 60) utilize concrete tanks, concrete inlet and outlet basins and steel covers serving as floating gas holders. A design for one of the many gas plants which have been built is shown in Figure 6. The digester has no pumps, motors, mixing devices or other moving parts and digestion takes place at ambient temperatures. Fresh manure is deposited into a charge pit which leads into the digestion tank. With a 50 day retention time the production of "dungas" or "gobar gas", as it is commonly called, is approximately 2 cu m/50 kg cow manure/day (Ref. 61). Since cows each produce 10 to 20 kg manure per day, the gas production is equivalent to 0.4 to 0.8 cu m per animal. A herd of 4 to 5 cows is required to provide a family's cooking, lighting, and fertilizer requirements, but few families are sufficiently wealthy to own this size herd of cattle (Ref. 62).

First attempts by the Chinese to use digesters began in the Great Leap period, but only since 1970 has the practice been promoted rigorously by the Chinese Academy of Services and the Ministry of Agriculture and Forestry.

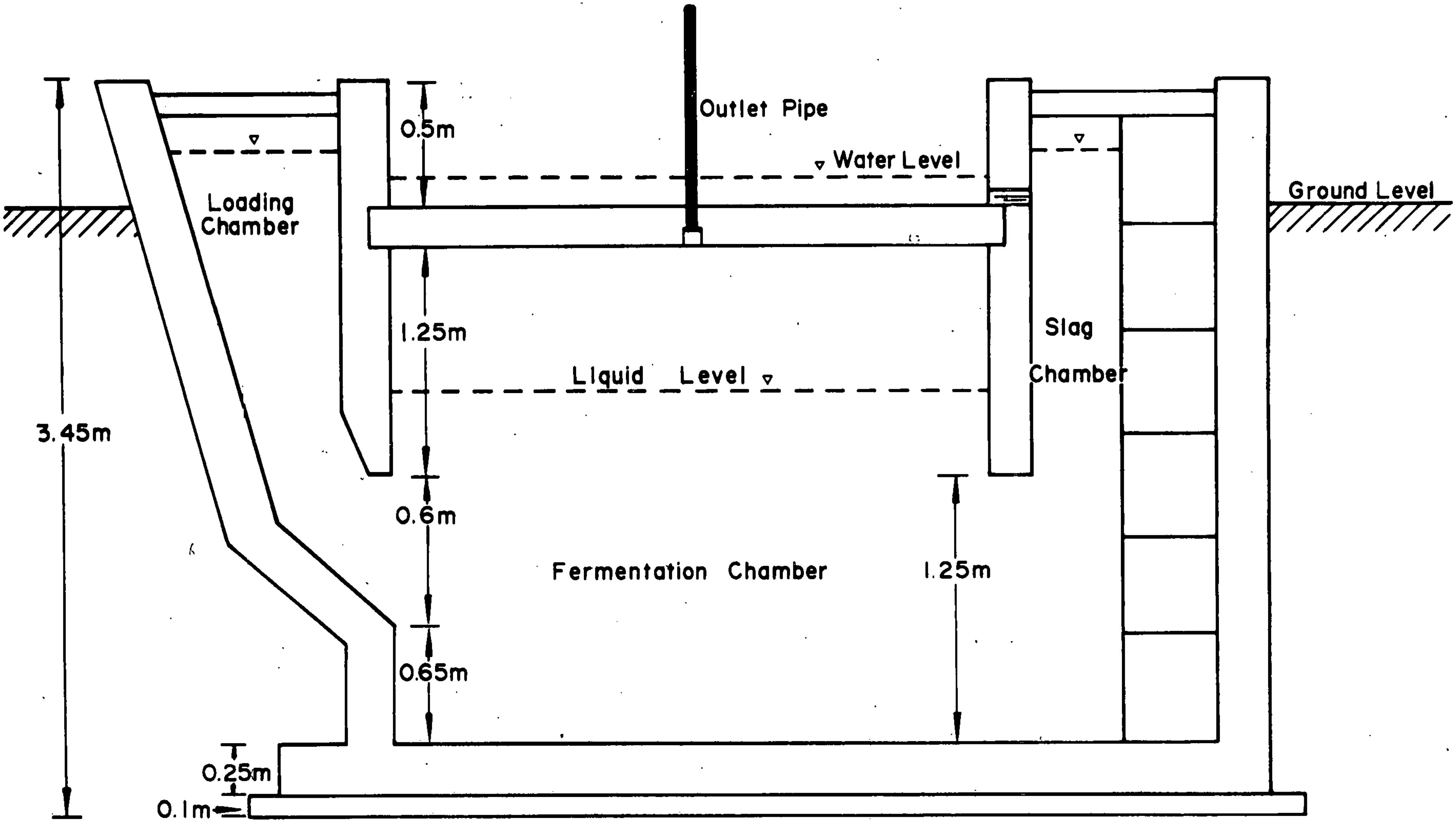


GOBAR GAS PLANT

Figure 6 A schematic diagram of a Gobar Gas Plant designed in India to produce gas and dried manure. (modified from Ref. 59).

The success of the promotional campaign is overwhelming with an increase in the number of digestion tanks from 30,000 in 1973 to 2.8 million in August 1976 just in the Szechwan Province (Southern China) alone (Ref. 63). The effort has been supported through national conferences to train technicians, manufacture rubber or plastic pipes and simple gas stoves and pressure gauges and to design unusually shaped fermentation pits. Construction of a typical 10 to 15 cubic meter digester, consisting of loading, fermentation and slag compartments (Figure 7) is done by the Chinese peasants using cement, rocks or bricks. Success of these digesters depends on warm temperatures in Southern China. There are problems with biogas generation, however, including sand accumulation, scum build-up and removal, collection and loading of human and animal waste. The contents of the digester must be stirred regularly to break up dried scum on the top of the liquid and to mix loaded materials. Slag must be removed periodically and the fermenter must be cleaned at least twice a year. According to the Chinese Office of Science and Technology the best combinations for digester loading for high gas production are 10 percent human waste, 30 percent animal waste, 10 percent straw and grass, and 50 percent water; or 20 percent human waste, 30 percent hog manure and urine and 50 percent water; or 10 percent each of human and animal waste, 30 percent marsh grass and 50 percent water (Ref. 64). When properly managed, the 10 cubic meter digester will supply a South Chinese family of five with enough gas for cooking and lighting (Ref. 63).

When the conventional single-stage digestion unit is used for the decomposition of relatively dry wastes (i.e., animal wastes with a solids concentration greater than 20%) the pronounced stratification of liquid wastes is not apparent. Wong-Chong (Ref. 65) has demonstrated the potential economical value of using a conventional batch-loaded method for what he terms "dry anaerobic digestion". Wong-Chong was striving to identify a method that is not as energy intensive as the high rate digestion process (discussion to follow). The mixing of dry wastes in high rate digestion becomes a problem when the solids content approaches 6% (Ref. 66), so that the feedstock must be diluted before digestion to facilitate mixing and then thickened after digestion to control disposal volume. Encouraged by the successful recovery



SZECHWANESE BIOGAS DIGESTER

Figure 7 A cross-section of a typical Szechwanese biogas digester, not drawn to scale. Flow of the water through a small hole (5 mm in diameter) in the wall separating the gas and the slag compartment keeps a relatively stable pressure inside the digester. The outlet pipe (metal or hard plastic) is about one meter long with an inside diameter less than 2 mm (modified from Ref. 63).

of methane gas from municipal solid waste landfills (Refs. 26, 67-69) Wong-Chong digested fresh and aged dairy manure and poultry manure in a conventional digester at ambient temperature. Although ammonia inhibited methanogenesis of the highly nitrogenous poultry waste, gas was successfully produced from aged and fresh dairy manure. From fresh manure 0.70 to 0.79 cu m of gas (60 - 65% methane) was generated per kg of volatile solids destroyed (11.3 - 13.0 cu ft/lb) and the aged manure produced about one half that amount. Economic problems associated with energy intensive high rate digestion were solved by "dry anaerobic digestion".

Bag Type Digesters

The use of bag digesters, which are known for their low cost, mass production capabilities and easy transportability, is reported to have originated in Taiwan (Ref. 70). The bag digester shown in Figure 8, consists of a sausage shaped bag made of 0.55 mm thick Hypalon which is laminated with neoprene and reinforced with nylon. The bag is held in an excavated hole and equipped with PVC inlet and outlet pipes and a gas release. A similar digester made of Red Mud-Plastic material has also been used in recent years for the digestion of hog manure in Taiwan (Ref. 71). This material exhibits extraordinary weather resistant characteristics and strong mechanical strength and according to the Union Industrial Research Institute where it was developed in 1975, the life of the red mud-plastic digester is expected to be about 8 years (Ref. 72).

Light weight, compact plastic digesters have been designed for field and laboratory use. One such system is a disposable multilayer plastic bag with a unique sealing device. A collapsible cardboard carton contained within a gas-impermeable plastic film is fitted with a catalyst and holders for a disposable hydrogen generator and an anaerobic indicator. There are no gas cylinders, vacuum pumps, valves and gauges. This system lends itself readily to compact storage, quick assembly and ease of operation and is also disposable after use (Ref. 73).

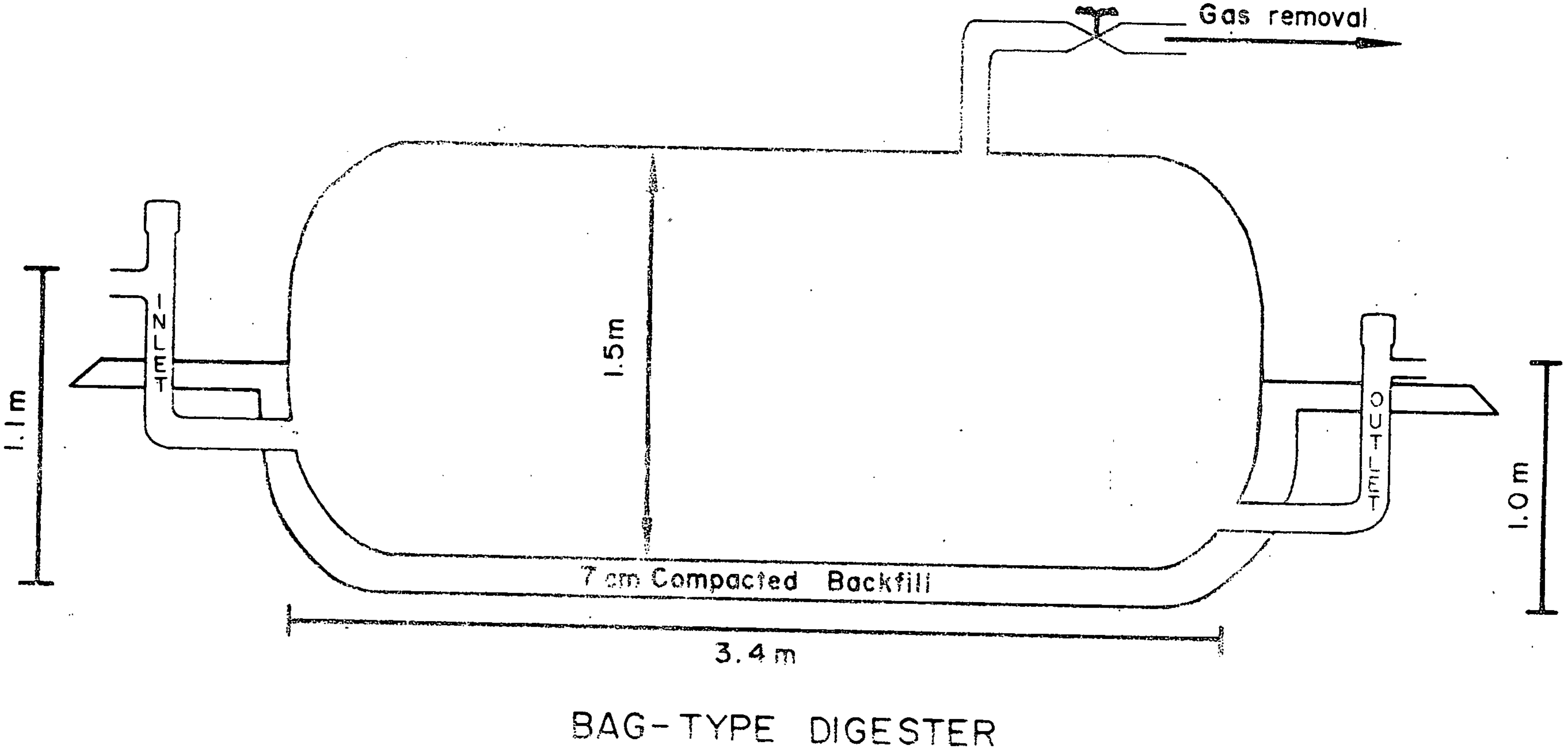


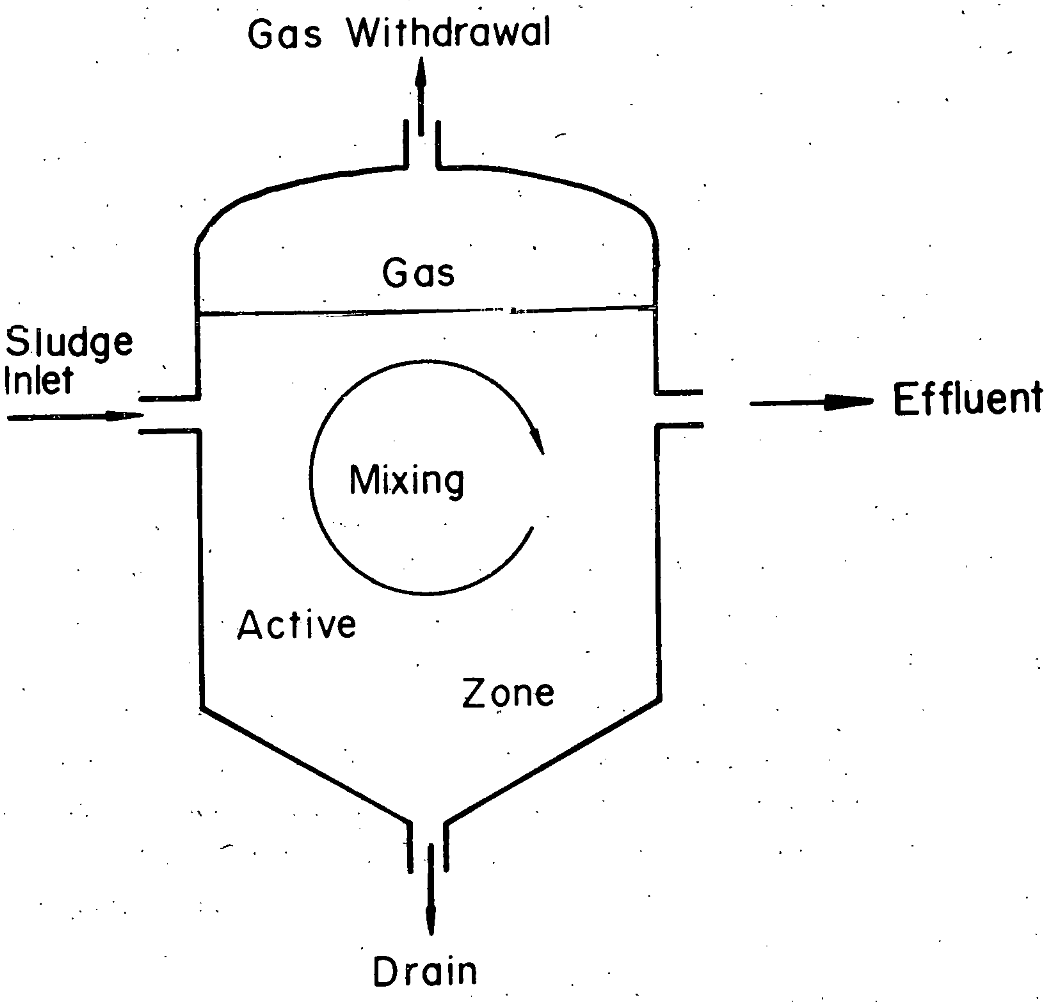
Figure 8 An illustration of a 5 m^3 Hypalon digester used in Taiwan (modified from Ref. 74).

High Rate Digesters

In most waste purification and gas production systems, the low loading rates and long detention times required for conventional digestion can not be tolerated. When the loading rates are increased and the detention times increased, however, there are insurmountable problems with scum buildup, "souring" due to localized acid buildup, and decreased contact between the substrate and microorganisms resulting in inefficiency of the system. To ratify these problems a more controlled energy intensive type of digestion was developed, known as high rate digestion. A high rate process (Figure 9) requires complete mixing of the digester contents by either a continuous or periodic system to maintain a homogeneous "mixed liquor" stage in all parts of the tank. If this criterion is met, the digester can function at loading rates of 2.4 - 6.5 kg volatile solids/cu m/day (0.15 - 0.40 lb/cu ft/day) and detention times of 15 days or less.

The key parameter which controls the efficiency of a high rate digester is increased contact between the microorganisms and substrate by mixing. Basic biological considerations assume that bacterial growth and metabolism require a source of energy or food, and it follows that if a maximum rate of metabolism and reproduction is to occur, the food must be available continuously and must be in intimate contact with microorganisms. Contact is accomplished by mixing the contents of the digester and thus bringing acid-forming bacteria into contact with organic matter and methane-forming bacteria in contact with acid intermediates.

Besides this major function of mixing, several other benefits are derived (Refs. 75 - 77) including 1) prevention of a scum blanket, 2) dispersion of potential metabolic inhibitors, such as volatile acids, throughout the unit avoiding localized conditions, 3) assistance with the disintegration of coarser organic particles into smaller particles with a greater net area, 4) maintenance of uniform temperatures throughout the digestion tank, 5) inhibition of large particle settling, 6) escape of gas bubbles on the liquid surface, and 7) improved process control.



HIGH RATE DIGESTION

Figure 9 Diagrammatic representation of a high rate digester. Standard operation characteristics:

1. Heated to 30 - 35°C.
2. Detention time 15 days or less.
3. Leaching rate 2.4 - 6.4 kg volatile solids/cu m/day (0.15 - 0.40 lb/cu ft/day).
4. Continuous or intermittent feeding and withdrawal.
5. Homogeneity.

As early as 1931, Bach (Ref. 78) discussed the advantages of mixing in digester design and operation. According to Sawyer and Schmidt (Ref. 79), the success of any high rate digestion system depends upon maintaining the contents of the digester in constant turbulence and of uniform character from top to bottom. Trubnick (Ref. 80) in comparing mixed and unmixed systems indicated that with equal loading and detention times, the mixed system produced a 30 percent increase in BOD reduction and a 20 percent increase in gas production. In a comparative study, Rufolds and Trubnick (Ref. 81) found that approximately 85 percent of the BOD was reduced in less than 3 days with agitation as compared to 25 days without mixing.

Mixing efficiency is influenced by the shape of the digestion tank. Different shapes of tanks have developed in Great Britain and the United States as compared with those in European countries. In the English-speaking countries cylindrical tanks with a level or slightly inclined bottom and cover are popular, while in Germany a more spherical form with a cone shaped top and bottom is used. The larger digestion tanks (about 12,000 cu m) in Germany approach more of an egg-shape. Although the Anglo-American shape is less expensive, the European conical shape allows more efficient circulation avoiding the accumulation of unmixed sludge pockets. (Ref. 77) Investigations have shown that in most cases, active digestion tank volume is only about 60 percent of the reactor volume. Methods which have been described to achieve efficient circulation within digesters include continuous pumping of the entire contents (Refs. 77, 82-86), gas recirculation (Refs. 77, 87-93), and mechanical circulation by means of screw impelled pumps, bottom scrapers and other agitators (Refs. 77, 94-98), and rotating drums (Refs. 2, 99, 100).

The value of temperature control for high rate digestion is well documented (Refs. 101 - 107). As early as 1934 Fair and Moore (Ref. 108) reported that four temperature ranges exist for anaerobic digestion bacteria: a thermophilic zone above 42°C, an intermediate zone 28 - 42°C, a temperate zone - below 10-28°C and a cryophilic zone below 10°C. Most investigators claim that digestion in the thermophilic range provided in typical high-rate digesters is a more rapid process than digestion in the mesophilic range. Rudolfs and Heukelekian (Ref. 109), Ardern and Hockett (Ref. 110) and Goleuke (Ref. 111) have shown significant increases in gas production rates and volatile solids destruction with increasing temperatures provided either sufficient time was allowed for appropriate types of organisms to become established or that digesters were appropriately seeded with temperature acclimated inoculum.

2.2.2 Two-Stage Processes

Two-Stage Digesters

The development of the next concept, two-stage digestion, originated with the realization that mixing was incompatible with solids separation in sewage treatment. The biological action of digestion had to be physically separated from the thickening process as described by Buswell et al. (Ref. 112). Sewage sludge containing about 5% solids was pumped from a sedimentation tank into an anaerobic digestion tank where it remained for about 4 - 8 days. Mixing and subsequent scum control were achieved by recirculating liquor from just below the scum layer. During the stage of decomposition about 50% of the solid matter was given off in the form of gas which was collected. The sludge was then pumped to a second unmixed tank where it was left to "ripen" to a consistency that would drain on a sand bed. The small quantity of gases produced in this second stage were allowed to escape. The apparatus patented by Fischer et al. (Ref. 113) for two-stage digestion incorporated several modifications in the process including gas collection in the second stage and a sludge rake in the second stage to direct the sedimented sludge to the centrally located sump at the bottom of the digester.

Two-stage digestion used currently for sewage treatment and for gas production from low solids organic materials is basically a combination of high rate digestion and conventional unmixed digestion as illustrated in Figure 10. The first stage is usually a continuously fed, completely mixed,

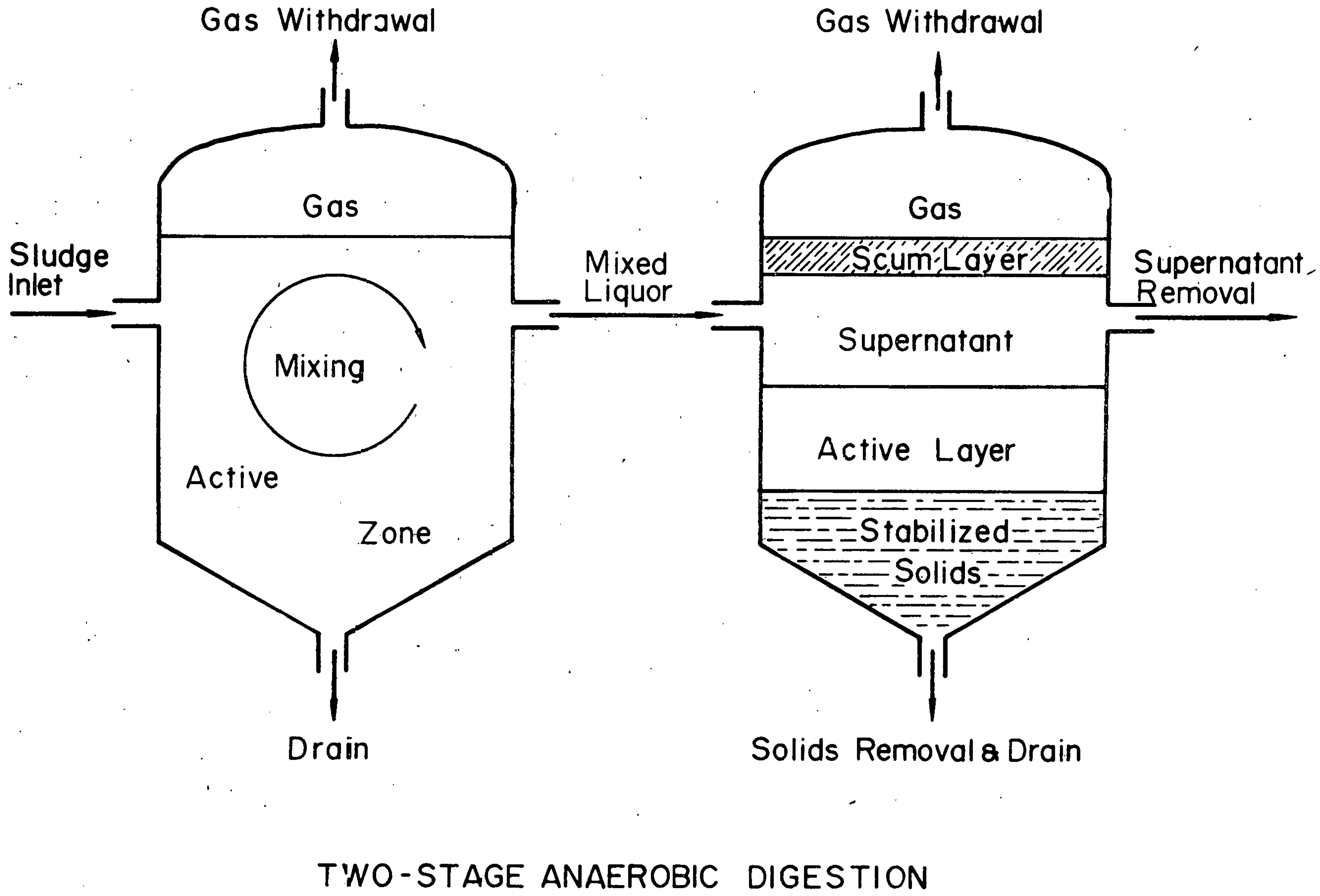


Figure 10 A schematic diagram of two-stage digestion consisting of high rate digestion in the first stage and conventional unmixed digestion in the second stage.

temperature controlled biological unit. In developing countries, however, the first stage is heated only by the warm climate and agitated 2 - 3 times daily 15 - 20 minutes at a time (Ref. 59). The second stage is a quiescent unit for solids separation and concentration. The modifications incorporated into the two-stage digester provide significant improvements over the high rate and conventional digester when used individually. Most important is the reduction in retention time from a 30- to 60-day range as used in the one-stage digester to a 10- to 20-day range in the two-stage digester (Refs. 3, 56, 75, 114).

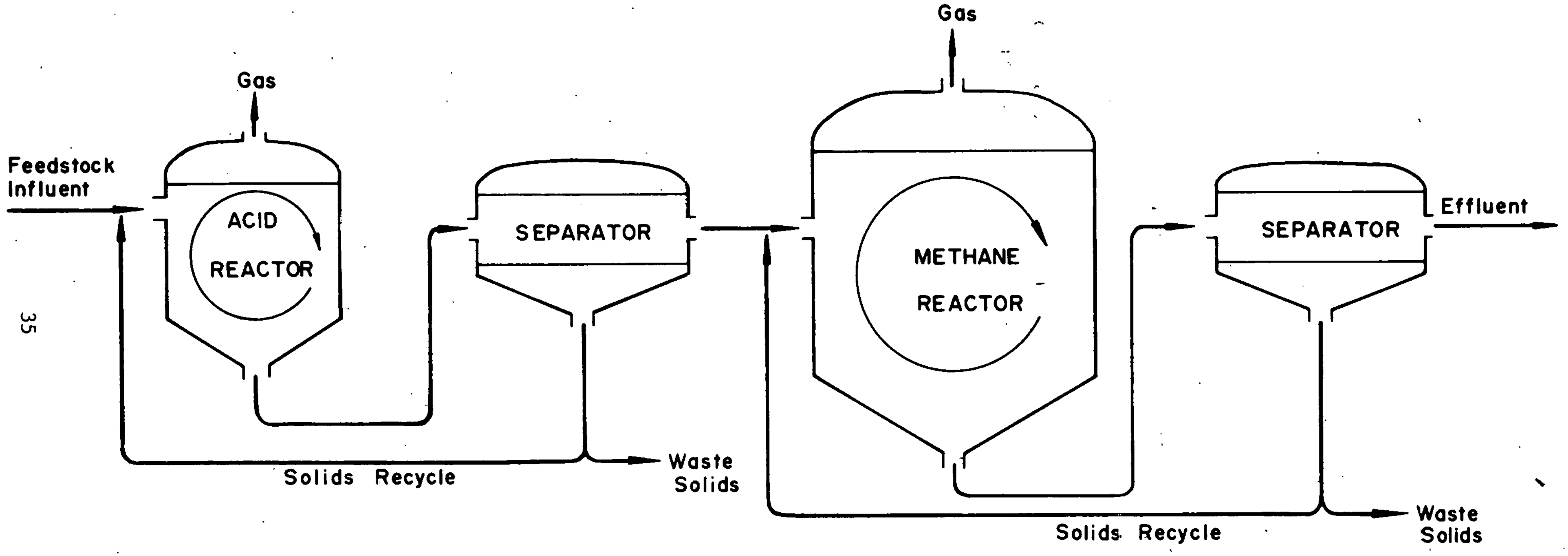
Two-Phase Digesters

Control of the rate of organic assimilation is limited to the first-stage in the original two-stage digestion process, but further research indicated that waste stabilization and methane production rates could be increased by providing optimum conditions for microorganisms in both the acid forming and methane forming stages. This concept is often termed "two-phase digestion".

In the first stage, the complex organics such as fats, protein and carbohydrates, are hydrolized, fermented and biologically assimilated by facultative and anaerobic bacteria. For the most part, the end products of the first-stage conversion are organic fatty acids. Acid forming bacteria bring about these initial conversions to obtain the small amounts of energy released for growth, and a small portion of the organic waste is converted to cells. Although no waste stabilization occurs during the first stage of treatment, the first stage functions to phase the organic matter in a form suitable for the second stage of treatment. During the second-stage organic acids are converted by several different bacteria to produce the gaseous end products, methane and carbon dioxide. Since the survival of the methane formers is dependent on that of the acid formers - a strict balance between the two is essential to the success of the system. When the system is in balance, the methane bacteria use the acid intermediates as rapidly as they are produced by the acid forming bacteria. If the methane bacteria are not present in suitable numbers, or are slowed down to unfavorable environmental conditions, the volatile acids will not be utilized.

In contrast to the original concept of two-stage digestion, two-phase digestion involves two high rate digesters in series, one for acid production and the other for methane production, as illustrated in Figure 11. Ideally, control of temperature, mixing, pH and loading rates can be achieved depending on the particular substrate and microorganisms present in each digester. The acid formers and methane formers are very different in terms of physiology, nutritional requirements, growth kinetic capabilities and sensitivity to environmental stresses (Ref. 115). By separating them into two vessels and controlling physical conditions within the two systems to meet the needs of the microorganisms, higher growth rates and methane yields can be attained.

The loading rates of each group can be controlled, resulting in enhanced process efficiency and reliability. The research of Ghosh and Pohland (Ref. 116) has indicated that under controlled conditions, the acid and methane formers have the potential of growing at minimum generation times of 34 min. and 5 hrs respectively. Another study has provided evidence that a two-phase digestion process may be operated at one-half the detention time of a high-rate digester and still exhibit higher rates of solids stabilization and methane production (Ref. 117). The first phase of a two-phase anaerobic digestion process patented by Ghosh and Klass (Ref. 118) is described as receiving an organic feed at a loading rate ranging from 16.2 - 162 kg total organics/cu m/day (1 - 10 lb/cu ft/day). For short detention times of less than two days the material is digested under conditions which efficiently liquefy and breakdown the feed to lower molecular weight acids and other intermediates for conversion to methane. The succeeding phase is loaded at a rate ranging from 1.6 - 8.1 kg total organics/cu m/day (0.1 - 0.5 lb/cu ft/day), and is operated to treat lower molecular weight acids and intermediates for detention times of about 2 - 7 days under conditions which lead to the production of methane.



TWO-PHASE ANAEROBIC DIGESTION

Figure 11 A schematic diagram of two-phase digestion involving two high rate digesters in series. In this case, the digesters are followed by separators to achieve efficient solids recycle.

Ghosh et al. (Ref. 117) have outlined the following advantages of two-phase digestion over single-stage digestion: 1) capability of maintaining the optimum environment for each group of digester organisms, 2) substantial reduction in total reactor volume and the consequent savings in capital and operating costs, 3) higher rates of solids stabilization and increased production rates and methane content of the final product gases, 4) decreased heat requirement and increased thermal efficiency, 5) suitability for incorporation into existing treatment plants with minimum capital investment and 6) reduction of the nitrogen content of the system effluent by simultaneous liquefaction and denitrification of waste feeds in the acid digester. The same authors state the need for skilled operation and increased instrumentation for monitoring and control as disadvantages.

A number of methods to achieve two-phase separation have been suggested including: 1) selective inhibition of each group of organisms through the addition of inhibitors, such as oxygen, nitrates, sulfates, metals, etc., 2) potential poisoning (Ref. 119), 3) kinetic control (Ref. 115), and 4) membrane dialysis separation. The latter of these separation techniques has received increased attention for laboratory studies. A simple dialysis cell, compartmented by semi-permeable membranes was developed in 1955 (Ref. 120) to observe associated and symbiotic relationships among selected species of aerobically grown microorganisms. The concept of applying dialysis separation between bacterial cultures in two fermentation tanks was conceived from the work of Gallup and Gerhardt (Ref. 121). These researchers used the dialysis unit to separate a nutrient reservoir from a pure culture of microorganisms under aerobic conditions. In anaerobic digestion studies the acid bacteria are separated from the methane bacteria even though rapid interchange of nutrients and metabolic waste products occurs throughout the unit (Refs. 122, 123). Laboratory use of dialysis units has proved successful, but due to operational difficulties with dialysis membranes, the potential is limited for large-scale anaerobic digestion (Refs. 115, 124).

2.2.3 Anaerobic Contact Processes

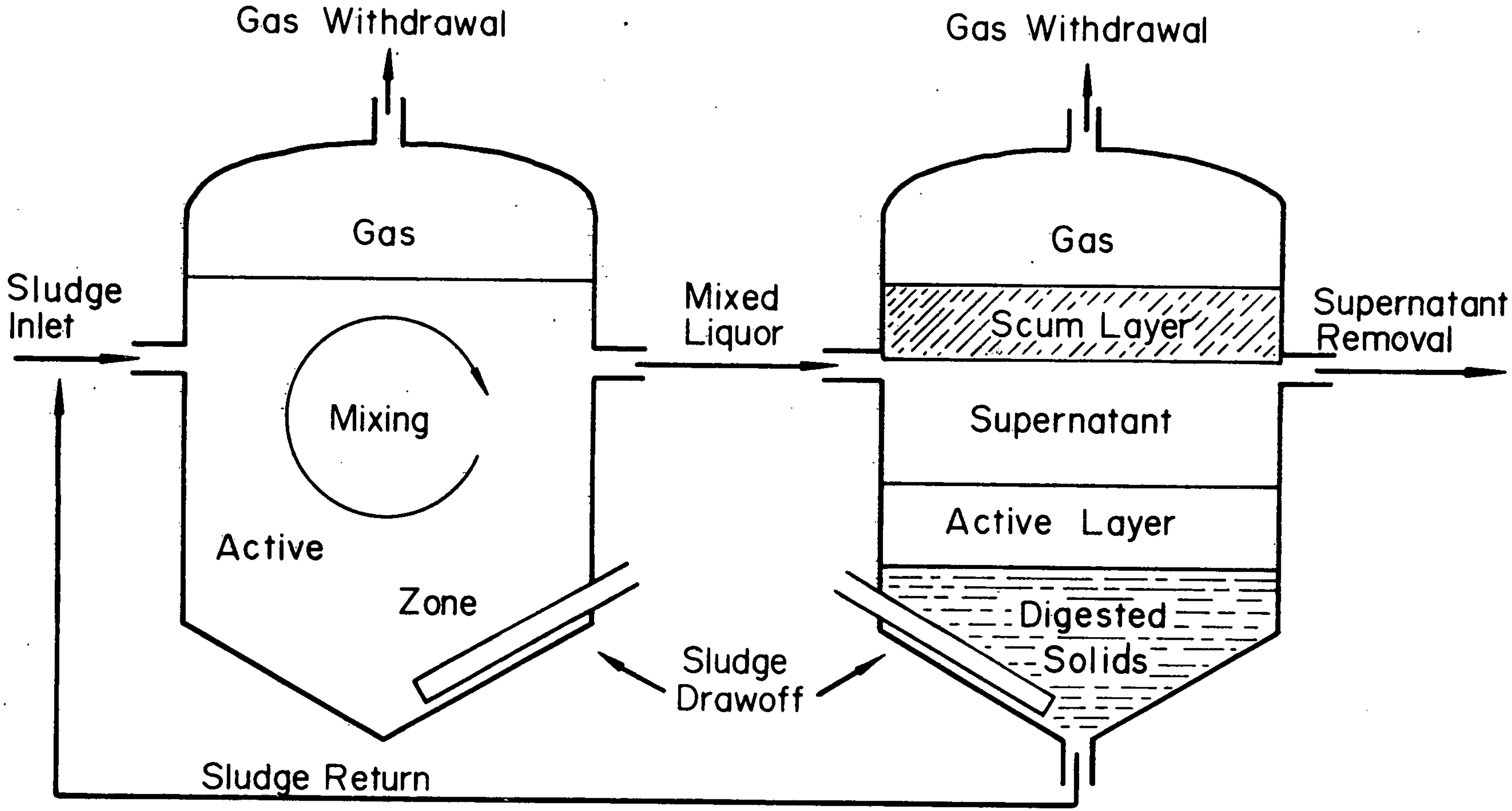
Digestion with Solids Recycle

The two-stage and two-phase digestion processes, as previously described, work well for substrates with high solids concentration, but when used for dilute substrates, such as industrial wastes, problems arise accommodating the large volume of organics in the second vessel. Since the retention time in the first vessel is shorter than that in the second vessel, the latter container must be as much as 15 - 20 times the size of the former, and a sludge thickening process is required. A method recycling the solid portion of the second vessel and associated microorganisms, known as the contact process, has been implemented to significantly reduce the required digester volume and at the same time to give a high solids retention time for stable operation (Refs. 125-133). As illustrated in Figure 12 the anaerobic contact process involves a high rate digester and a separation tank with solids recycle. The long solids retention time with short liquid retention time is achieved by recirculating fermenter liquid through the solids separation or concentration unit which minimizes waste of solids in the effluent. The design for solids retention times, as outlined by McCarty (Ref. 134) is shown in Table 2.1. In systems without solids recycle, dilute wastes may not support an adequate active microbial mass, but with sludge return, the microbial population can be maintained at levels to obtain optimum efficiency in a system having a greater flexibility than found in single and two-stage anaerobic units without recycle.

Table 2.1

Design for Solids Retention Times

	Solids Retention Times, Days	
Operating Temperature °F	Minimum	Suggested for Design
65	11	28
75	8	20
85	6	14
95	4	10
105	4	10



ANAEROBIC CONTACT DIGESTION

Figure 12 A schematic diagram of anaerobic contact digestion involving a high rate digester followed by a separation tank with solids recycle.

In an evaluation of the early literature, Gates et al. (Ref. 135) defined the anaerobic contact process as meeting the following criteria: 1) use of sludge recycle, 2) an influent concentration of 4,000 mg/ℓ BOD or less, and 3) a liquid detention time of four days or less based on influent flow. More recently, however, this process has been used for stronger treatment of high-strength wastes containing BOD concentrations in excess of 4,000 mg/ℓ.

The anaerobic contact process has since been reported in the literature for treatment of cannery wastes (Refs. 97, 136, 137) brewery wastes (Refs. 138, 139), distillery wastes (Refs. 140-143), cotton kiering liquor (Ref. 144), synthetic milk wastes (Ref. 135), strawboard waste (Ref. 145), yeast wastes (Refs. 146, 147), maize starch (Ref. 148), sewage (Ref. 149), molasses wastes (Ref. 146), and meat packinghouse wastes (Refs. 125, 126, 144, 150-156). Successful contact process operations have been reported with BOD_5 loadings varying from 1.2 to 11.8 kg/cu m/day (74 - 730 lb/1000 cu ft/day). In two cases (Refs. 152, 157) successful treatment was reported with temperatures of only about 24°C (75°F) and BOD_5 loadings of about 1.6 kg/cu m/day (100 lb/1000 cu ft/day). Often industrial wastes are already preheated to 32 - 35°C (90 - 95°F) and further heating is not necessary in the digester. The hydraulic detention time for successful treatment has been reported ranging from 0.5 days for meat packing wastes (Refs. 150, 152) and 3.8 days for molasses (Ref. 146).

In research using the anaerobic contact process the key emphasis has been on methods to achieve sufficient solids separation to permit continuous solids recycle. High efficiency is necessary to maintain the required long solids retention times while operating at short hydraulic detention times. Utilization of the contact process for sewage treatment is often compared to the return activated sludge step in an activated sludge plant since solids recycle is employed to achieve more efficient BOD reduction. Patents relating to the treatment of sewage sludge by the contact process were described by Torpey (Refs. 158, 159). A thickening stage between clarification and high rate digestion of recycled sludge eliminates the supernatant problem. A later patent by Amero (Ref. 160) described the contact process for sewage treatment in which separation of the acidic fraction of mother liquor low in solids was accomplished by centrifugation. By replacing the removed liquor and by suitably regulating the process undesired acid buildup in the digester was minimized. In an experimental study by Torpey and Melbinger (Ref. 147) digested sludge from a single-stage digestion system was recirculated back to the thickener and mixed with raw sludge at an optimum recycle ratio of 0.5. Recycle of the digested sludge increases volatile solids destruction from an average of 20,000 kg/day (44,000 lb/day) to 24,100 lb/day (53,000 kg/day). The additional solids destruction was also reflected in the reduction of the volatile content of the digested sludge from 59.4 to 53.5% during recycle.

As early as 1936, Buswell and Boruff (Ref. 161) patented a contact process for the treatment of industrial wastes and for gas production. It was emphasized that if accurately regulated the recycled sludge acts as a buffer to control acidity in the high-rate digester and an increase in gas production can be expected.

One of the most successful applications of the contact process, treatment of meat packinghouse wastes, utilized vacuum degasification followed by gravity sedimentation for solids separation. The degasifier is required between the digester and settling tank to remove large quantities of gases which tend to float the solids rather than allowing them to settle. The flow diagram shown in Figure 13 illustrates the process used for full scale treatment at Wilson & Company (Ref. 150). Preheated organic wastes are held in a concrete tank at a detention time of 12 - 13 hours based on the flow of raw wastes. The solids in the mixed liquor (suspended solids concentration 7,000 to 12,000 mg/ℓ) are actively digested as the mixed liquor leaves the digesters. The mixed liquor is then discharged through vacuum degasifiers and to two sludge separation tanks where the solids settle by gravity. The solid sludge is then returned to the digesters as seed to maintain anaerobic culture. The detention time in the separators is about 1.2 hours, based on total flow,

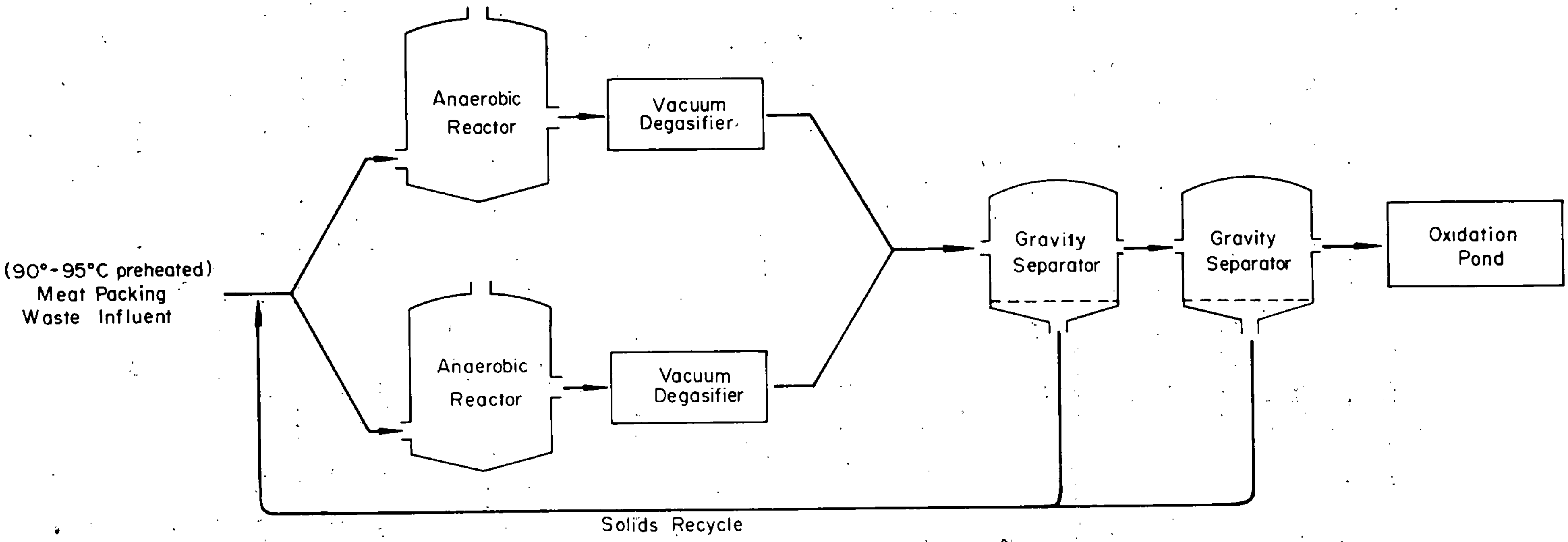


Figure 13 Flow diagram illustrating the process used for full-scale treatment of meat packing waste at Wilson & Company (modified from Ref. 150).

including sludge recirculating through the system at 3 volumes per volume of influent. In spite of the fact that the residual gases are removed in the degasifier, the sludge is still flocculent and must be removed with a suction-type sludge removal mechanism. The treated effluent overflows into weir troughs discharging into two oxidation ponds for final BOD reduction.

A vacuum degasifier placed before the clarifier was also used to reduce gas content in the treatment of starch-gluten wastewaters (Ref. 139) with a BOD of 12,000 mg/ℓ and a total solids of 14,000 mg/ℓ. Volatile solids reductions of 80 and 70% were attained at detention times, including recycle, of 22 and 14 hrs, respectively.

Illustrated in Figure 14 is a pilot plant system used recently for the treatment of rum distillery slops by the contact process with vacuum degasification between digestion and clarification stages (Ref. 143). In the treatment of slops with a COD of 70 - 100 g/ℓ, effluents containing less than 30 g/ℓ COD were produced at solids retention times greater than 40 days. It was demonstrated that methane recovery from this system on a plant-scale resulted in the reduction of unit treatment costs by at least one-third at a design capacity of 190 cu m/day and as much as two-thirds at a design capacity of 1,140 cu m/day.

Although vacuum degasification is efficient, it is no longer popular due to high costs. Researchers are seeking new methods to achieve high solids concentrations in the contact process.

For studies of the treatment of carbonated kiering liquors and slaughterhouse wastes, Pettet _et al_. (Ref. 144) attained solids separation through the use of a spiral-channel clarifier in which the warm reactor (33°C) effluent travelled along the surface. The surface travel was of sufficient duration to allow degasification and subsequent gravity settling. For the kiering liquor, detention times of 12.1 to 1.3 days and loadings ranging from 0.14 - 11.9 kg/day/cu m (0.009 - 0.74 lb/day/cu ft) were employed with continuous solids separation and intermittent recycle. For slaughterhouse wastes a detention time of 30 hr and a loading rate of 1.6 kg BOD/day/cu m (0.1 lb/day/cu ft) were used.

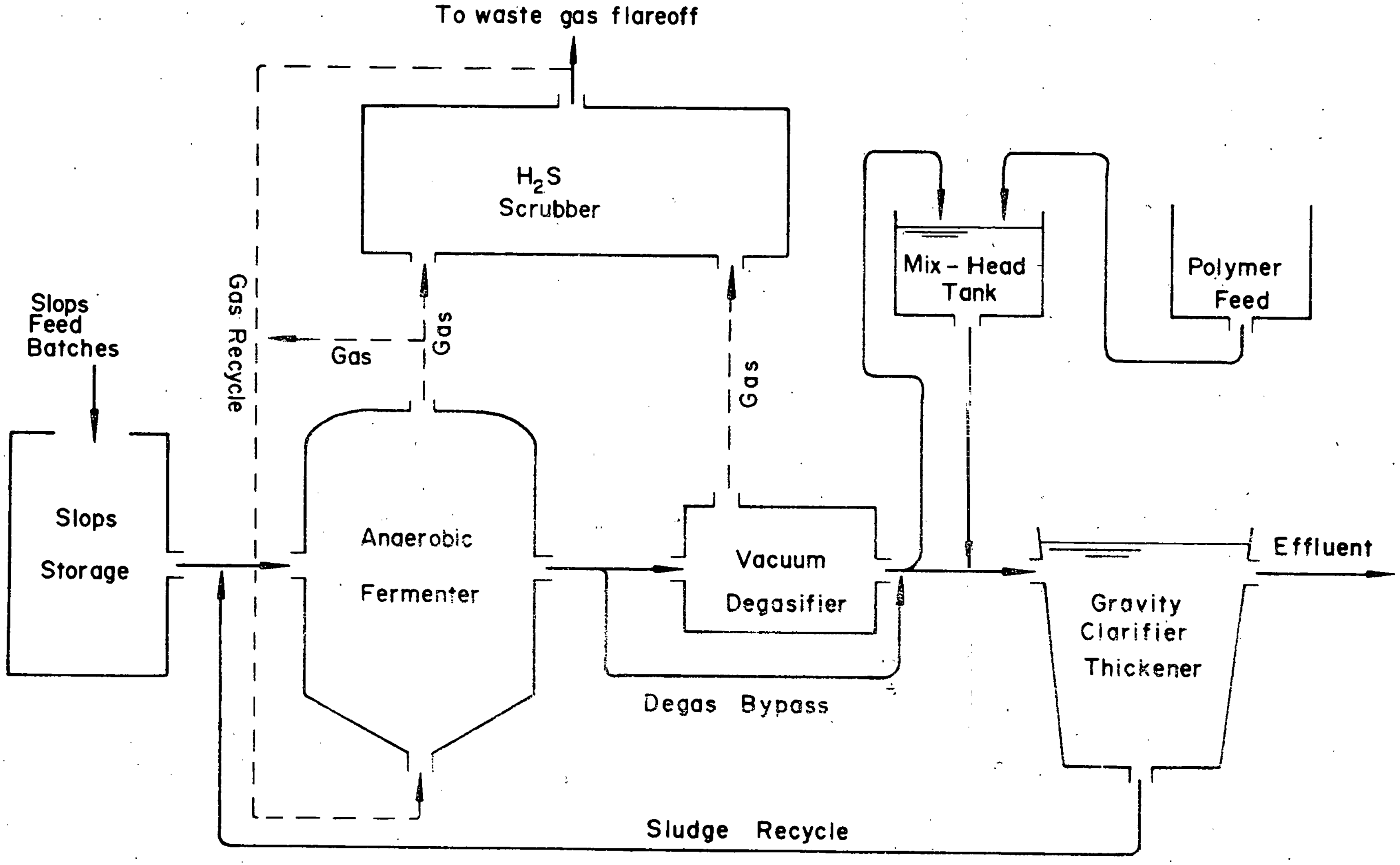


Figure 14 An illustration of the contact process used successfully on a pilot-scale for treatment of rum distilling slops. Fermentation is followed by vacuum degasification and gravity sedimentation to achieve efficient solids separation and recycle.

Newton et al. (Ref. 138) attempted to reduce the gas content of solids in brewing wastewaters by aeration of the mixed liquor before entering the clarifier. This system has been questioned by Gates et al. (Ref. 135), however, due to toxic effects of oxygen.

The treatment of synthetic-milk wastes, conducted on a laboratory-scale using an upflow anaerobic contact unit in which the second stage settler is positioned over the first stage fermenter, has been reported (Ref. 135). The settling tank consisted of a baffle-type separator which showed good potential for accomplishing solid separation in the process. Fermenter hydraulic residence times as low as six hours and loadings varying from 0.8 to 5.4 kg BOD/day/cu m (0.05 - 0.34 lb/day/cu ft) were used successfully.

Van den Berg and Lentz (Refs. 97, 162) chose pear wastes to study the application of the anaerobic contact process for treatment of fruit processing wastes, since like most of these wastes, pear wastes have high carbohydrate contents and are nutritionally imbalanced. Experiments were performed with waste from a mechanical pear peeling line in a cannery which was stored frozen until required. After thawing, the waste was diluted with water to a 9% solids concentration, homogenized in a Waring blender for 3 min and passed through a screen to remove fibers that might plug the tubes in the peristaltic pumps. Water, ammonium and phosphate salts (for an alkalinity of about 3,000 ppm $CaCO_3$ in the fermenter liquid) and nitrogen and phosphate as nutrition for the microorganisms were added to the pear waste. The sedimentation unit was operated on a one-hour draw-and-fill cycle, rather than continuously, with liquid retention times of 12 - 24 hrs and continuous gentle agitation. The settler was allowed to fill and the waste settle for 45 min. Then the effluent was withdrawn and the solids returned to the fermenter. Where high suspended solids content of the fermenter liquid limited gravity settling, low speed centrifugation was used for separation of solids and effluent. The results of the study indicated that the contact process was applicable to pear waste with COD-removal efficiencies of up to 95%, provided ammonium and phosphate salts, as well as other nutrients present in the commercial fertilizer "Milorganite" or in yeast extract, were added.

Bioenergy Process

An anaerobic contact process known as the Bioenergy Process has recently been developed by Biomechanics, Ltd. under the direction of G.M. Rippon for the treatment of wastes on a commercial-scale. The company brochure lists possible applications of the process as treatment of animal wastes, distillery and brewing wastes, fermentation wastes, fruit wastes, milk wastes, fish processing wastes, olive oil processing wastes, paper processing wastes, vegetable processing wastes, wine production wastes and wet separation of starch and gluten from wheat. However, the Bioenergy Process has its main application in the food processing industry where low grade waste heat is available (Refs. 163, 164).

The Bioenergy Process, illustrated in Figure 15, involves a combined heating and mixing system which has been marketed under the name Biomix, a digester and a Bioenergy Separator. In the lower portion of the digester organic wastes are heated and agitated gently and in the upper portion, the gases produced are collected. The digested sludge is regulated through the Bioenergy Separator to remove liquid treated effluent.

The patented Biomix (Ref. 165) consists of a pump which utilizes thermosyphon and gas lift principles to mix the digester contents and control scum formation, and a heat exchanger which warms the digester contents to an optimum temperature of 35°C. Waste heat from the industrial process which requires cooling may be utilized to increase the temperature of the digestion process. Biomechanics, Ltd. claims that Biomix equipment is low in cost and maintenance free over periods of several years of continuous operation (Ref. 164).

The Bioenergy Separator, also patented by Biomechanics (Ref. 166) is a cooler and gravity separator which provides solids separation by subjecting the treated liquid to a thermal shock and allowing solids to settle. The temperature of the liquid is reduced suddenly from 35°C to 25°C within 30 sec in the cooler, causing microbial activity and gas production to cease temporarily. The cooled treated liquid and bacteria in the form of a dispersed suspension are fed from the cooler to a gravity separator. After entering the gravity separator, the bacteria tend to flocculate into relatively dense clumps and

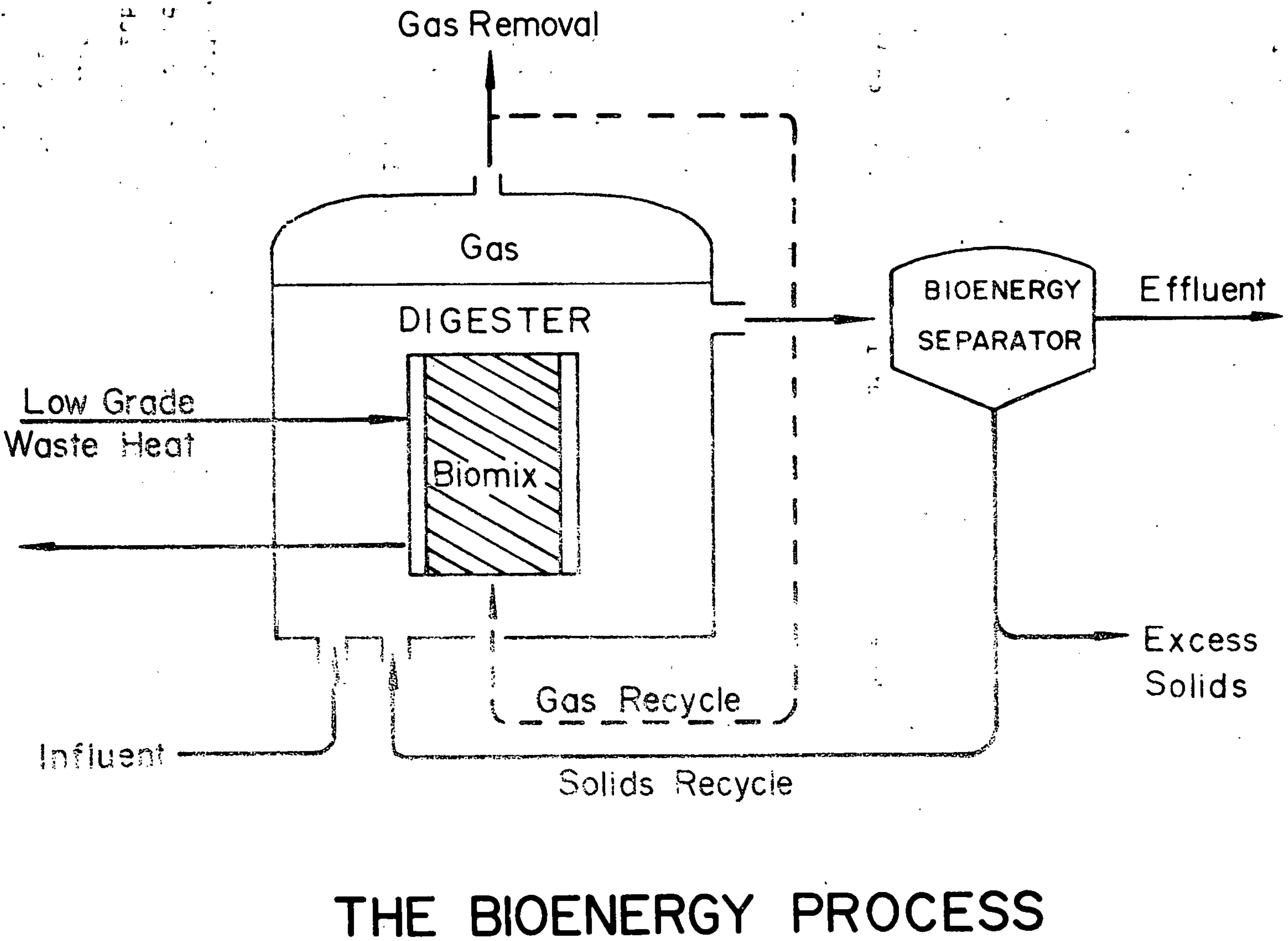


Figure 15 A diagrammatic representation of the Bioenergy Process patented by Biomechanics.

settle to form a sludge at the bottom. After about 10 hrs a relatively clear liquid is removed from the tap and the sludge containing bacteria is removed through the bottom of the separation unit and fed back into the digester.

The effect of the thermal shock on reducing the metabolic rate of the bacteria is only temporary and after returning to the 35°C digester, the bacteria regain at least some of their former activity. Even at the low temperature in the separator some bacteria begin to produce gas before the sludge is removed. As a safeguard against this occurrence, solids are prevented from floating upward into the clearer liquid layer by packing material in the configuration of Raschig rings or sheets of perforated metal. Any solids raised by gas bubbles encounter turbulence as they float around sharp edges of the packing material which is sufficient enough to cause the solids to be dislodged and sink back to the bottom.

The Bioenergy Process has been economically successful for commercial treatment of wastes from wheat starch gluten factories in France (Tenstar Aquitaines in Bordeaux) and in England (Tenstar Products, Ltd. in Ashford, Kent). The success of the process is attributed to the efficient use of waste heat produced during industrial processing of wheat flour (Refs. 163, 164). Interest in the process for treatment of other food processing wastes has increased recently due to the realization that with the improved technology of the Bioenergy Process capital costs are comparable to those of conventional aerobic methods and operating costs are reduced due to the production of fuel gas (Ref. 167).

2.2.4 Attached Film Processes

The technology of an attached film process is based on the principle that actively metabolizing microorganisms will attach to an inert substrate and form a sustained concentration in a defined area around the substrate. Anaerobic digestion by the attached film process involves a column or tower packed with inert material which has a large surface area for microbial attachment and the column from below and move upward coating the media with a film of microorganisms which cling readily, but loosely, to the substrate surface and remain in the media bed to metabolize organic matter. Additional flocculent biological growth is suspended between the filtration particles (Ref. 76). The 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 (Ref. 168) and short liquid retention times (6 - 12 hrs) without danger of biomass washout. Since the microorganisms remain in the filter for a long period of time, solids retention time is long and efficient anaerobic waste treatment is possible with low energy input.

Packed Bed Filter

The attached film concept originated with the anaerobic packed bed filter used by Coulter <u>et al</u>. (Ref. 169) for solids separation in the second stage of an anaerobic contact process for sewage treatment. The anaerobic filter, illustrated in Figure 16, was a submerged column packed with rocks for the attachment of microorganisms. The liquid retention time was 2.5 hr for the filter process (35 hrs for the total system) and suspended solids removal was greater than 90%.

Pretorius (Refs. 170, 171) designed a two-stage system based on that of Coulter <u>et al</u>. (Ref. 169) consisting of a digester for solids concentration and hydrolysis and a packed bed filter termed a biophysical filter for gasification. As illustrated in Figure 17, treated sludge from the first digester flowed into the bottom of the filter and the movement of solids flowing upward was retarded by a series of layers of particles decreasing in size from 2 cm in diameter at the bottom to filter sand (16 - 30 mesh) at the top giving a void ratio of 0.6. Using this system up to 90% COD reduction of raw sewage with an original COD of 500 mg/ℓ was obtained at a 24 hr liquid retention time and a temperature of 20°C.

Treatment of more concentrated industrial waste in the range of 1500 to 3000 mg/ℓ COD were investigated by McCarty (Ref. 172) and Young and McCarty (Ref. 173) using only the anaerobic filter without the first-stage digester. Substrates utilized in early experiments (Ref. 172) included methanol,

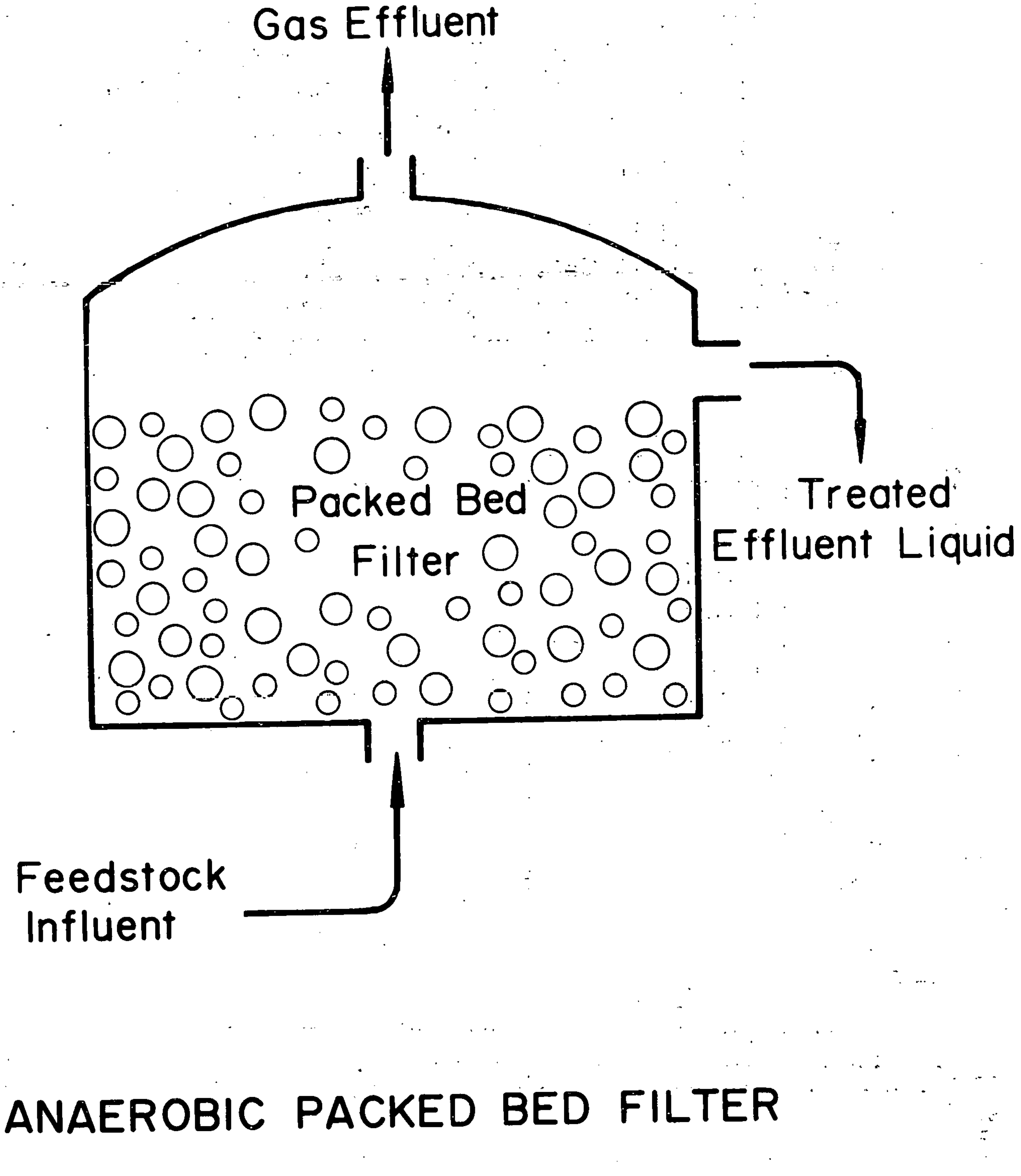
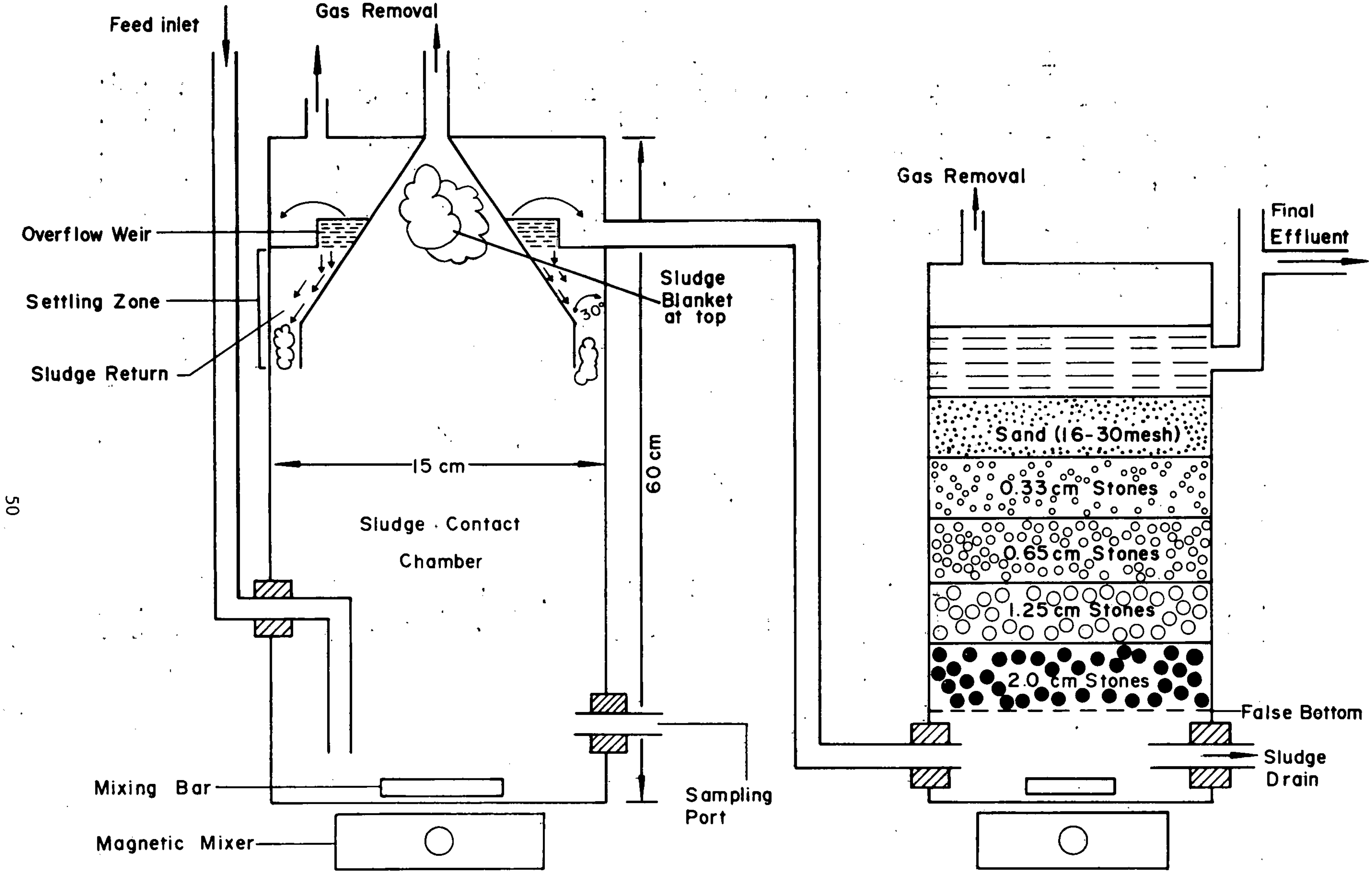


Figure 16 Illustration of an attached film process, the packed bed filter consisting of a submerged column packed with rocks for attachment of microorganisms.



PRETORIUS' SYSTEM

Figure 17 Illustration of a two-stage system consisting of a digester for solids concentration and hydrolysis and a packed bed filter termed a biophysical filter for gasification. The series of layers of particles in the second stage provide attachment for microorganisms.

methanol plus acetate, methanol plus acetate plus propionate, and acetate plus propionate. Removal of 74 - 88% COD was accomplished by wastes originally having strengths of 2,940 - 2,650 mg/ℓ COD at organic loading rates between 1.7 - 3.6 kg COD/cu m/day (107 - 226 lbs/cu ft/day) and liquid retention times of 6 and 12 hrs. Solids retention time was greater than 100 days. In later experiments (Ref. 173) using a protein-carbohydrate synthetic substrate, as high as 90% COD removal was accomplished by treating wastes originally having strengths of 1500 and 3000 mg/ℓ COD at organic loading rates of 0.4 and 0.9 kg COD/cu m/day (26.5 and 53.0 lbs/cu ft/day) and at liquid retention times between 4.5 and 72 hrs. At higher loadings of 3.4 kg COD/cu ft/day (212 lbs/1000 cu ft/day) the COD removal efficiency dropped to 36.7 - 63.0%. Solids retention time in the experiments varied from 84 to 665 days.

The experimental success and kinetic analysis conducted by Young and McCarty (Ref. 173) on the anaerobic filter laid the groundwork for future application of this process. Investigations using the anaerobic filter have been undertaken for treatment of carbohydrates in food processing wastes (Ref. 174), acetic acid (Ref. 175), potato processing wastes (Ref. 176), wheat starch waste on a commercial scale (Refs. 177-179), petrochemicals on a pilot scale (Ref. 180), brewery press liquor (Refs. 181, 182), brewery wastes, paper mill condensate, acetic and glycolic acids (Ref. 183), nonionic detergent (Ref. 184), "Metrecal" (Ref. 185), pharmaceutical waste containing 95% methanol (Refs. 186, 187), and dilute waste sulfite liquor (Ref. 188). These wastes, ranging in strength from 2,000 - 27,000 mg/ℓ COD were treated using packed bed filters composed of stones, intalox saddles, berl saddles, raschig rings or ping pong balls about 2.5 - 3.8 cm (1 - 1-1/2 inches) in diameter. Details on the conditions and significance of these investigations have been summarized in table-form by Mueller and Mancini (Ref. 189).

Fluidized or Expanded Bed Reactor

Due to the static nature of a packed bed filter, organic sludge tends to accumulate in the interstices of particles in the treatment of high-strength wastes. The system becomes clogged and flocculent biological growth is limited. To relieve this problem, substrates composed of small particles about 1 mm in diameter, such as carbon and sand (Refs. 190, 191), ion exchange resin and PVC commercial filter media (Refs. 168, 192, 193) have been used as filters. Unlike the relatively large particles of the packed bed filter, these particles are highly motile while the system is in operation. Feedstock and recirculated liquids enter the reactor at the base at a rate sufficient to achieve the upflow velocity necessary for a slight expansion of the media, so that there is no danger of clogging. Since the reactor volume changes while the unit is in operation, this attached film process has been termed the fluidized process (reactor volume increases up to 100%) or expanded bed process (reactor volume increases up to 10 - 20%) (Ref. 168). A schematic of this process is shown in Figure 18. Other advantages of the expanded bed filter over the packed bed are the increase in surface area available for biological growth per unit of reactor volume and the smaller headloss (Ref. 190).

Secondary sewage has been denitrified successfully in fluidized bed reactors containing either activated carbon or sand, at liquid retention times from 3 - 10 minutes (Refs. 190, 191). Pilot plant investigations resulted in a 99% efficiency of nitrite and nitrate nitorgen removal at a liquid retention time of 6.5 min when methanol was added as a carbon source.

For the anaerobic treatment of domestic sewage strength synthetic substrate, Leuschner (Ref. 193) used an expanded bed filter similar to that designed by Jewell (Ref. 192) for aerobic sewage treatment. Spent ion exchange resin and PVC commercial filter media were provided for the attachment of microorganisms. COD removal efficiencies varied from 50 - 90% when synthetic substrate loading rates of 0.8 - 8.9 kg COD/cu m/day (0.052 - 0.552 lb/cu ft/day) were employed at liquid retention times ranging from 3 - 12 hrs.

Using the same filter composition, Jewell _et al_. (Ref. 168) recently fermented a high strength particulate organic residue, cow manure diluted to 2% total solids, with a total volatile solids destruction efficiency of 39.5% at a liquid retention time of 1.2 days. By comparison, high rate continuous stirred tank reactors operate at equal efficiencies at a liquid retention time

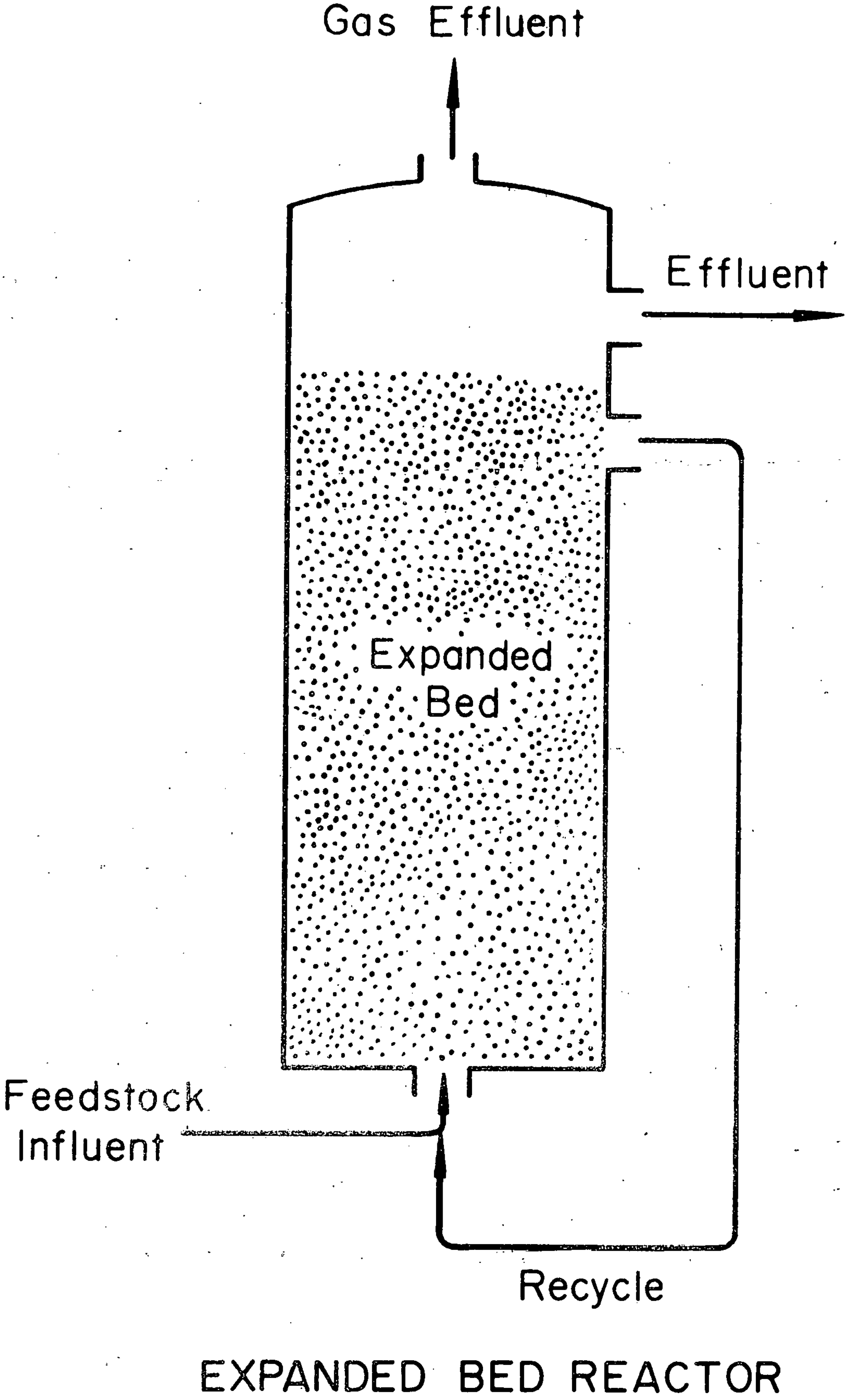


Figure 18 Illustration of an attached film process termed the expanded bed filter which consists of a column packed with very small mobile particles.

of 10 days with the same feedstock. The process was stable and capable of producing gas at liquid retention times as low as 0.15 days (3.6 hrs), although total volatile solids destruction efficiencies decreased to 7.2%. At the highest loading rate of 132.8 kg total solids/cu m /day (8.2 lbs/cu ft/day) and liquid retention time of 0.15 days (3.6 hrs), a minimum gas production of 6.60 volumes of gas per volume of reactor per day were reported.

Application of the fluidized bed has been identified for the fermentation industry as well as in sewage sludge treatment. A "completely mixed microbial film fermentor" has been developed for the continuous fermentation of brewers yeast (Ref. 194). A "tapered fluidized bed" has been used for denitrification and in the immobilized enzyme conversion of lactose to glucose and galactose, and in enzyme-catalyzed hydrogen production (Ref. 195).

2.2.5 Multi-Stage Processes

Some investigators (Refs. 196 - 198) have indicated that anaerobic digestion is not just two-phase production of acid formers and methane formers, but a multi-stage process involving many different bacteria which have a specific function in the step-wise degration of long-chain polymers, such as carbohydrates, fats and proteins. The steps involve: (1) enzymatic hydrolysis to produce amino acids, sugars and fatty acids, (2) further anaerobic decomposition to low molecular weight end-products such as volatile fatty acids, lower alcohols, and aldehydes and a variety of gases, and (3) the production of methane from carbon dioxide and hydrogen and of carbon dioxide and methane from the volatile fatty acids and other acid-stage products. It has been argued that because optimum environmental conditions for the different microorganisms vary from group to group and species to species, the isolation of the biological reactions into separate vessels should enable the environmental conditions to be varied in order to maximize the rate of each reaction (Refs. 83, 199, 200).

Multi-stage processes consist basically of a series of reactors combining any number of digester processes described previously. The most common multi-stage process is a series of continuously stirred tank reactors controlled to provide variations in temperature and pH. A series of stirred reactors simulates the plug flow reactor concept. The system may involve complete recycle or recycle in certain stages. Schematic illustrations of these systems are shown in Figure 19.* Conditions of the system and economic feasibility will vary depending on feedstock characteristics and the overall objective of the system.

In many fermentations the objective is not the production of microorganisms, but of metabolic products, such as antibiotics, enzymes, solvents, alcohols, organic acids, etc. The formation of the products may or may not be associated with the growth phase of the microorganisms since many chemical transformations may be carried out by microorganisms which are not growing. Product formation without growth can arise because of the different environmental conditions (pH, aeration, temperature, etc.) required for growth and product formation, as in the production of penicillin, vitamin B_{12}, ethanol, and butanol, or because the principal substrate cannot support growth, but can be metabolized by the microorganisms with the formation of the desired products. Under these circumstances a single-stage digester can provide only compromise conditions for both growth and product formation, while in a multi-stage system the principal substrate can support growth as well (Ref. 83). Maxon (Ref. 201) has summarized the use of multi-stage systems for the production of ethanol by yeast in which from 2 to 12 reactors in series were used with and without recycle.

Multi-stage fermentation may also be used to advantage in the formation of complex media containing several substrates each of which function as a limiting nutrient and support growth at different rates. In a multi-stage process, the more rapidly utilized substrate would be consumed in the first stage with other substrates being utilized in other stages (Ref. 83).

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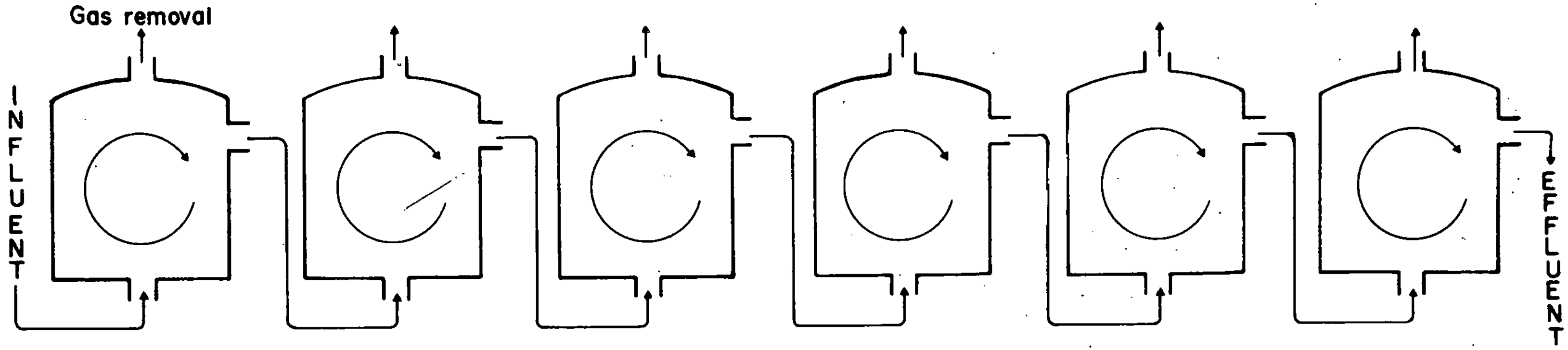
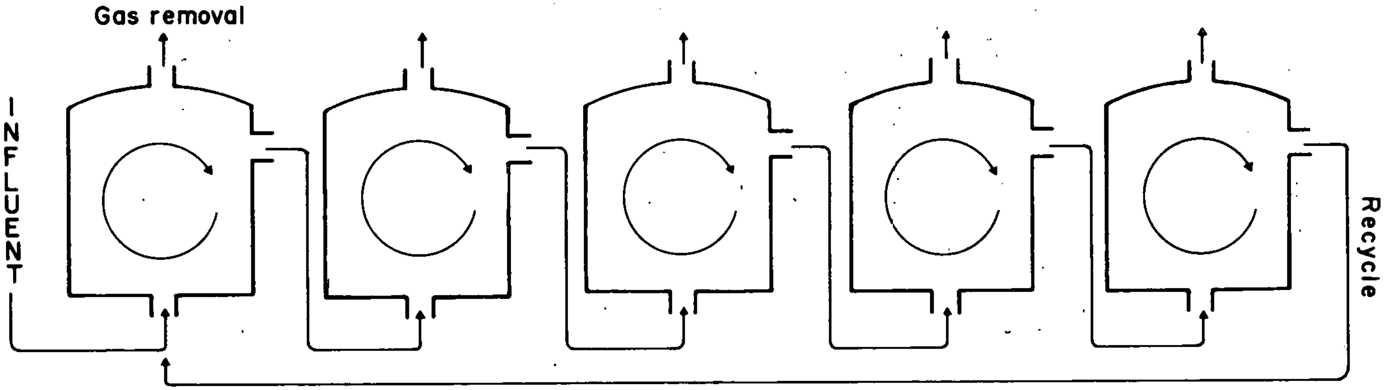
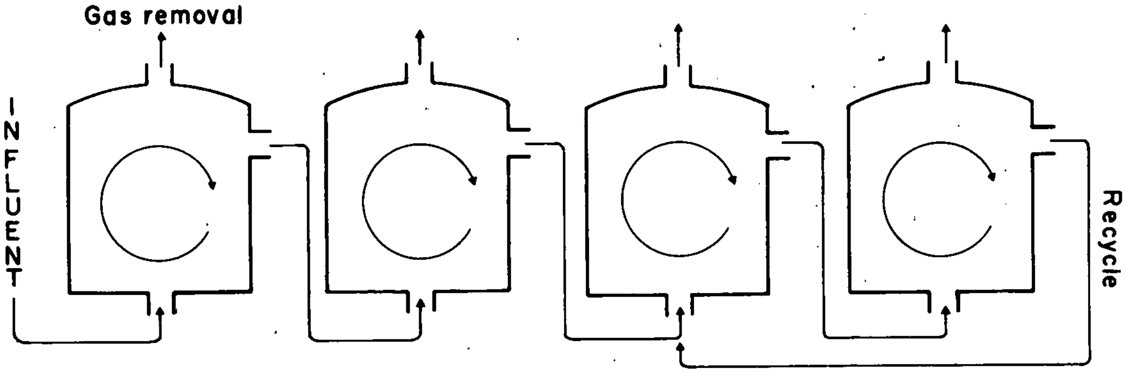
* Processes involving an aerobic stage before anaerobic digestion are sometimes termed "multi-stage". In this report, however, the aerobic stage is considered a pretreatment process.



Series of continuously stirred tanks



Series of continuously stirred tanks with complete recycle



Series of continuously stirred tanks with partial recycle

Figure 19 Flow diagram of three types of multistage processes.

Another application of multi-stage systems can be made to systems involving series reactions in which the product of one reaction is inhibitory to organisms carrying out other reactions in the process. In this instance the various reactions are carried out in separate vessels to segregate toxic compounds from susceptible organisms (Ref. 56).

2.2.6 Biomethanation

Several gasification processes have the potential to produce methane from a synthesis gas mix, containing principally hydrogen, carbon dioxide, and carbon monoxide. Such thermochemical processes for biomass utilization have an efficiency advantage, in that all organic components of biomass, including lignins, may be effectively converted to synthesis gas. On the other hand, high temperature, high pressure catalytic methanation systems are required to convert this synthesis gas mix to pipeline quality methane. However, a simpler and potentially more economical alternative to complex catalytic methanation is to utilize a biological system to carry out the conversion to methane of biomass pyrolysis gases.

It has been clearly established that microorganisms carry out the reactions (both shift conversion and methanation) required to methanate synthesis gas. Both technical feasibility and methods for improvement of the rate of these reactions have been demonstrated at the bench scale (Ref. 202). Results showed methanation rates of up to two orders of magnitude higher than those typical for sewage sludge, solid waste, or animal residue digesters. Operation was at atmospheric pressure as well as operation to 30 atm. In addition, an engineering and economic analysis projected favorable economics for the process relative to chemical catalytic methanation.

A problem with anaerobic digestion of biomass is that biomass components with significant fractions of fuel value, particularly lignin, are not susceptible to digestion and remain unconverted by the process. Further, these refractory components tend to shield the digestible cellulosic fraction so that energy-intensive pretreats may be required for full utilization of the digestible fraction. As a consequence, processes for gasification of biomass have come under consideration.

With gasification, biomass is subject to controlled combustion in the presence of oxygen and, sometimes, steam. Gasification has the advantage that all organic components of the waste are converted to a synthesis gas mix consisting largely of CO, CO_2, H_2. The net energy yield based on feedstock in the product synthesis gas mix increases with decreasing moisture content in the feed. For low moisture and high refractory content feedstocks (such as wood, which is high in lignin) net energy yield with gasification may be expected to exceed that of digestion.

Gasification processes under development and suitable for biomass include the Union Carbide Purox® process, the critical pressure water gasification process of Wright-Maita, and the Fullman-Kellogg hot alkali carbonate process. It is probable that other existing gasification processes developed for coal, such as the Lurgi and Koppers-Totzek, could also be applied to biomass with modification, although such research is yet to be performed.

A factor common to all present or likely future gasification processes is that the product will consist of a synthesis gas mixture of varying percentages of carbon monoxide, carbon dioxide, hydrogen, and methane, and trace gases. Methanation will be required to upgrade this product gas to pipeline quality.

The best-known and most intensively studied technology for the conversion of synthesis gas to pipeline methane involves the use of chemical catalysts. Although these chemical catalytic techniques have been under development for an extended period of time a number of problems remain. Problems include catalyst inactivation by carbon deposition, the potential for inactivation by sulfur, and the requirement for large quantities of steam at some points in the process.

The basic chemistry of methanation can be described using only a few reactions. When the hydrogen to carbon monoxide ratio in synthesis gas is equal to or greater than 3, the conversion of CO and H_2 to methane can be described by the reaction:

$$3H_2 + CO \rightarrow CH_4 + H_2O + 49.3 \text{ k cal/mol} \qquad (1)$$

Two other reactions also act to produce methane from carbon oxides:

$$2H_2 + 2CO \rightarrow CH_4 + CO_2 + 59.1 \text{ k cal/mol} \qquad (2)$$

$$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O + 39.4 \text{ k cal/mol} \qquad (3)$$

However, hydrogenation of carbon dioxide, Equation 3, does not occur in the presence of carbon monoxide. It may be pointed out that Equation 2 can be considered to be a combination of Equation 1 and the so-called shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2 + 9.9 \text{ k cal/mol} \qquad (4)$$

Although the shift reaction does not produce methane, it plays an important role in the catalytic conversion process by altering the H_2/CO ratio to increase utilization of CO.

A further reaction is important since it may lead to a deposition of carbon on the catalyst with eventual fouling of the catalyst - this reaction is:

$$2CO \rightarrow C + CO_2 + 41.2 \text{ k cal/mol} \qquad (5)$$

From a purely technical viewpoint, high selectivity to methane is not difficult to obtain. Rather, the problems are prevention of catalyst inactivation by sulfur compounds or carbon deposition and also those arising from the highly exothermic nature of methanation. With nickel catalysts, for example, it is usual to limit sulfur in the gas to less than 1 ppm by rigorous purification. Carbon deposition on the catalyst is avoided by operation with a sufficiently high H_2/CO mole ratio.

Over-all, while a substantial amount of work has been done, fundamental problems exist and are inherent in the entire catalytic conversion process. An alternative processing concept is that of carrying out the entire conversion of coal gasifier product gases to methane by anaerobic fermentation.

Biomethanation involves anaerobic fermentation of CO, CO_2, and H_2 to methane. Anaerobic fermentation is most often considered as the splitting of a substrate into two or more fragments, part oxidized and part reduced - relative to the composition of the original substrate compound. With this above definition of fermentation,CO may be viewed as intermediate between carbon dioxide (totally oxidized carbon) and methane (reduced carbon). With this insight, Fischer, Lieske and Winzer (Ref. 203) demonstrated experimentally that microorganisms (derived from an anaerobic sewage sludge digester) were able to bring about an anaerobic conversion of CO into CO_2 and CH_4. The work of Fischer et al, with (a) CO and (b) CO and H_2 pointed at the conclusion that the primary reaction in which CO takes part is always:

$$CO + H_2O \rightarrow CO_2 + H_2 \tag{6}$$

In the presence of a sufficient amount of hydrogen, these products are converted according to the reaction of Equation 7.

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \tag{7}$$

Thus, the fermentation (or biomethanation) of synthesis gas, consisting of CO, CO_2, and H_2, will produce methane as described by Equations 6 and 7.

It was experimentally shown (Ref. 202) that the rate of methane production from carbon dioxide and hydrogen is exceedingly rapid. CO_2 and H_2 were the only carbon and energy source fed to the fermentor. Under mesophilic conditions (37°C), the methanation rates reached values of 96 VVD. Under thermophilic conditions (60°C), volumetric rates of methane production approached 240 VVD. These values, as one would expect, compare exceptionally well with methane production rates of 5 VVD observed from cellulose digestion.

2.3 Pretreatment Processes

2.3.1 Physical Pretreatment

The mode of physical pretreatment required to prepare a substrate for anaerobic digestion varies with composition and physical consistency of the feedstock. In general terms physical pretreatment includes: 1) fractionation to separate out inorganic non-digestible materials leaving a high percentage of organic matter, 2) particle size reduction of the organic material to provide maximum surface area for biological decomposition by microorganisms, and 3) alterations in moisture content to allow easy flow of the feedstock through the digestion system. Progress in the technology of these processes has been due primarily to advancement in municipal waste treatment methods. The same processes are applicable to the preparation of biomass for anaerobic digestion.

Fractionation

Municipal solid wastes consist of a conglomeration of food wastes, vegetative matter, cloth and synthetics, paper products, plastics, leather, rubber, glass, metal and dirt, of which about 75 - 80% (2/3 or more is cellulose) is organic and 20 - 25% is inorganic. Similarly biomass feedstocks may consist of inorganic dirt and nondigestible material which is undesirable in the digester. As illustrated in Figure 20, fractionation of municipal wastes to remove light and heavy inert materials and ferrous metals is achieved by a step-wise pretreatment system involving primary shredding, magnetic separation, screening and air classification. Secondary shredding follows these steps to achieve further size reduction and increased biodegradability. Pretreatment using these five steps may involve a number of alternative systems by altering the sequence of the steps or deleting steps or repeating steps, (Refs. 204 - 207).

The function of the primary shredding step is to reduce the raw solid waste to a reasonably uniform particle size to ease process handling. The primary shredder is sized and powered to permit reduction to a nominal particle size of 10 - 15 cm (Ref. 206).

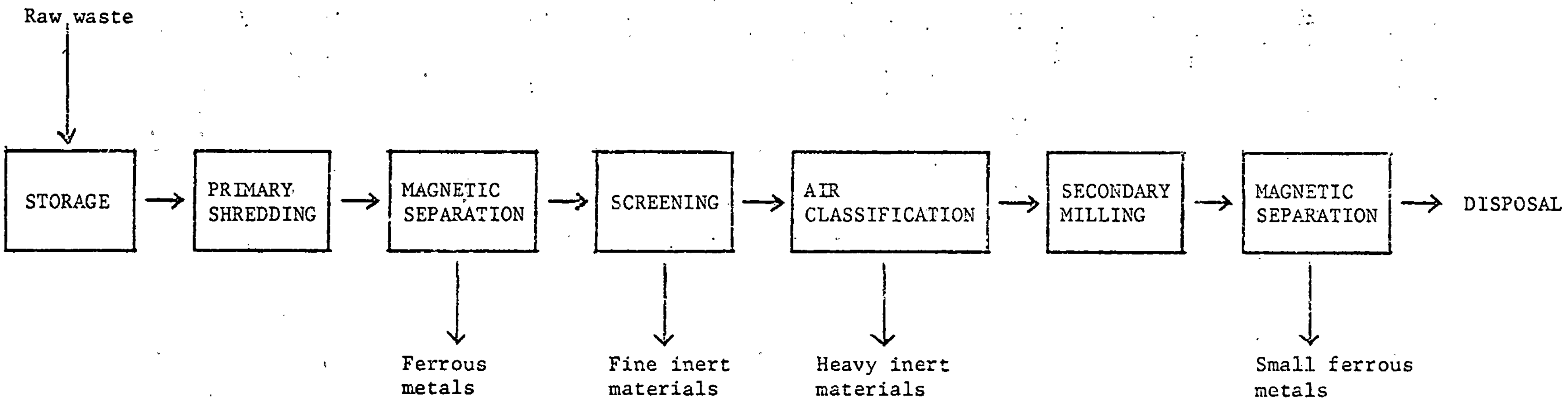


Figure 20 Fractionation of municipal wastes to remove light and heavy inert materials and ferrous metals.

After the primary shredding operation, separation of magnetic ferrous materials may be accomplished by utilizing an electromagnetic separator which will remove about 95% of the magnetic metals. The purpose of this separation is to reclaim the magnetics for resource recovery and to prevent these materials from interfering with the mechanics of the operation.

The refuse then passes through a trommel screen to remove fine inorganic ash, shattered glass, ceramics and dirt. The degree of removal of the inert materials is partially a function of the particle size of the inert materials relative to the size of the screening openings. The quantity of material removed will depend upon the particle size as well as the mode of operation. The amount of material removed by the screen may be as low as 10% of the total raw waste if the air classifier precedes the screen and as great as 30% of the total raw waste when secondary milling precedes the screening operation without prior air classification. A vibrating-type screen is not considered suitable for removal of small inert materials, due to the heterogeneous nature of milled solid waste and due to the frequent occurrence of blinding on a horizontal screen surface. Since the quantity of material passing through the screen varies considerably, depending on the desired particle size, a trommel screen with provision to vary the screen opening size is required. The screen openings in one set of plates should be about 1 cm in diameter and in the other set, about 2 cm in diameter (Ref. 205).

An air classification step is employed to remove the heavier denser fraction of waste from lighter organic materials more suitable for anaerobic digestion. The heavy fraction consists of the remaining particles of metal, glass and ceramics, and the more dense organic particles of leather, rubber, plastic, agglomerates of wet paper and waste food particles. The quantity of material in the dense fraction depends on the mode of operation as well as relative particle size. When the waste has been coarsely ground without prior screening the heavy fraction may be as great as 27% of the total raw waste, but if the material has already been screened, the heavy fraction is only about 7.5% of the total raw waste.

Size Reduction

After fractionation municipal solid wastes are further treated by secondary milling to a size that is desirable for the anaerobic digestion process. The particle size of the end product of the pretreatment process is usually about 15 - 20 cm for landfilled solid waste, but less than 6 cm for deposition into tank digesters.

In landfill operations, milling, shredding or pulverizing refuse provides for significant volume reduction so that more refuse can be deposited per cell and hence, less cover material is needed. The refuse can be compacted tighter during daily operations and there is less settling upon completion of the landfill. Since the surface area of the refuse increases, there are more areas of contact for microorganisms and decomposition of organic matter proceeds at an accelerated rate (Ref. 208). The composition of leachates from landfills containing shredded waste indicates anaerobiosis and methane production occurs initially at a much higher rate than in unshredded waste landfills (Ref. 22). While excavating 25 year old unshredded municipal refuse at the Palos Verdes Landfill in California researchers found that the rate of decomposition was not uniform, resulting in large quantities of unreacted organic materials (Ref. 23). Since shredding provides a homogenous mixture of waste with increased surface area this patchiness in decomposition should not occur in pretreated landfills.

Shredding of refuse before deposition also promotes movement of moisture throughout the landfill. Moisture levels in landfills are generally low, ranging from 15 - 40%. Over a long period of time moisture accumulates in pockets in the landfill leaving some refuse very wet and some very dry (Ref. 23). It has been demonstrated that without a minimum of 25% water by weight the anaerobic process is virtually non-existent. As moisture content increases, gas production increases to a maximum production at 50 - 80% moisture content (Refs. 25, 69, 209-211). However, excessive wetness can be detrimental to methanogenesis (Refs. 25, 26, 209, 212). A more uniform distribution of moisture throughout the landfill should increase the production of gas.

Characteristics of the leachate, on the other hand, remain about the same regardless of whether it comes from a fill of milled refuse or refuse which has not been milled (Ref. 213).

Researchers at the Institution of Gas Technology suggest a tertiary milling stage in a patented Biogas Process (Refs. 214, 215). After secondary shredding, the waste is conveyed pneumatically to storage by a cyclone separator, and as the stored material is required, it is fed to high-speed tertiary shredders for particle-size reduction to minus 20 - 30 mesh. The final step of the Biogas Process before digestion is mixing the refuse with municipal sewage and removal of any remaining inorganics. The finely ground material is conveyed to an air-solids cyclone separator and fed to a blending tank where it is mixed with raw sewage. Particles of grit settle out and are removed from the tank, leaving a highly organic mixture. The prepared feedstock is fed to a holding tank where it is retained periodically until required for continuous feeding to the digestion system.

In two patents, Schlentz (Refs. 216, 217) identifies several operational problems with combining raw sewage and organic refuse pretreated by shredding. The principal problem is due to the non-digestible grit particles that remain in the refuse. When added to a conventional sewage sludge digestion system this material settles out quickly, clogging pipes and valves. Schlentz indicates that even when the sewage-refuse mixture is held in grit chambers, 25% of the refuse solids pass out of the settling tank in the form of colloidal and dissolved constituents of the effluent. The second problem is the volume of the raw sewage-ground refuse mixture. During the grinding stage, water is added to refuse to achieve maximum operational efficiency. As a result, there is an excess of liquid when ground refuse and raw sewage are combined. These problems are solved by an improved method of combining the organic refuse fraction with supernatant as it enters the refuse digester. The refuse is fractionated by conventional means leaving organic material and a small quantity of non-digestible grit for fermentation. Before the refuse is fed to the digester, it is combined with supernatant containing seeding organisms from a separate sewage sludge digester, and the mixture is passed through a grinder. Gas from the refuse digester is circulated to the sewage sludge digester for mixing, and gas from both digesters is collected under a dome cover. There is no circulation of solids through the system since grit from the refuse digester and settled solids from the sewage sludge digester are removed from below.

Although vegetative residues often have already undergone some size reduction during processing operations, further milling is advantageous to increase degradation rates. Investigation of the effects of physical modification of cellulosic materials on their digestibility by ruminant-inhibiting microorganisms have indicated that either fine grinding or ball milling is the most effective treatment to date for increasing digestibility of wood and other cellulosic material (Ref. 218). Ball milling not only results in a reduction in particle size but also a reduction in crystallinity, a reduction in mean degree of polymerization, and a marked increase in the fraction of material that is water soluble. The increase in size reduction increases the available surface of both amorphous and crystalline cellulose, thus increasing accessibility to treatment by large enzyme molecules (Refs. 219-225). Data reported by Mandels et al. (Ref. 226) on the digestibility of milled cellulose in relation to time milled indicate that a linear relationship exists between increase in digestibility and increase in milling time. Millett et al. (Ref. 227) notes, however, that the degree of effectiveness of ball milling varies considerably with different materials.

Alterations in Moisture Content

Depending on the original moisture content of the feedstock and the digestion process to be incorporated, dilution or thickening may be necessary. Alterations in moisture content of woody and herbaceous vegetative residues and manure are made employing methods similar to those used for sewage treatment. The raw feedstock can be diluted by the addition of fresh water, recycled water from the digestion system sewage sludge or other highly liquid materials.

Dilution is practiced to lower energy requirements during milling operations, to ease the flow of material through the digester apparatus, to increase the efficiency of mixing during digestion and to increase the digestibility of the feedstock. For digestion of continuously stirred tank reactors the solids content should be less than 6% (Ref. 66).

Several digestion processes, including the contact process and attached film processes, are designed for treatment of wastes high in moisture content by separation of liquid effluent and solids recycle (Section 2.2.3). Under these conditions the only physical pretreatment required is size reduction of large solids. However, for treatment by conventional and high rate digestion systems which do not employ solids separation, solids thickening is necessary for water removal and subsequent volume reduction, for solids concentration and to eliminate excess supernatant problems. In addition, the thickening stage provides blending and grit removal.

In conventional sewage treatment, organic solids separation is achieved in several stages by clarifiers and thickeners. Both of these stages are applicable to pretreatment of other high moisture materials. The theory and design of operation of these stages for wastewater treatment are reviewed in detail by Black, Crow and Eidsness (Ref. 228).

Clarification accomplishes the removal of particulate matter, chemical floc and precipitates from suspension through gravity settling. Sedimentation basins are designed for slow uniform water movement with a minimum of short circuiting. Partitioning baffles guide the flow vertically to collecting troughs that extend across and around the periphery of the clarifier. Scrapers drawn by endless chains slowly move the settled solids to a sludge hopper at the inlet end. The clarified slurry is then thickened by any of several methods, including gravity thickening, centrifugation and air flotation.

Compared to the clarifier, a gravity thickening tank is deeper to accommodate a greater volume of material and has a heavier raking mechanism.

Waste material enters from behind an inlet wall in the center of the tank and is directed downward. Pickets or palings attached to collector arms stir through the slurry providing cavities for the release of trapped water. The supernatant overflows a peripheral weir, while overflow of thickened sludge is drawn from a bottom sump in the tank. Gravity thickeners are sized on the basis of solids loading per square foot of tank and on the rate of consolidation of individual waste materials (Ref. 3).

Two types of centrifuges are used for thickening: the basket type and the solid bowl scroll type. The basket type consists of a spinning cylinder creating high centrifugal forces to push solids against a drum wall. Feed slurry enters in the center at the bottom and clarified liquid discharges over a lip ring at the top. When cake depth in the bowl reaches a predetermined thickness, the centrifugation process is interrupted to remove collected sediment.

The solid bowl scroll type centrifuge consists of a rotating solid bowl in the shape of a cylinder with a cone section on one end and an interior rotating screw conveyor. Feed slurry, entering from the center, is held against the bowl wall by centrifugal force. Settled solids are moved by the conveyor to one end of the bowl while liquid effluent is discharged at the other end. The conical bowl shape at the solids discharge end enables the conveyor to move the solids out of the liquid for drainage before being discharged. The major advantage of this centrifuge over the basket type is operational flexibility. By controlling bowl volume, bowl speed and conveyor speed a wide range of moisture contents can be obtained in the solids discharge.

Thickening by dissolved air flotation is employed whenever particles tend to float or if the waste material has a long subsidence period and resists compaction for thickening by gravity. The waste material is fed to a pressurized mixing chamber where air is introduced through a revolving impeller or porous media. Particles float with the air bubbles to the surface and the clarified effluent is discharged under a baffle and over an adjustable weir which controls the depth of penetration of skimming blades. Thickened material is removed by a skimming mechanism and grit is removed from below along with settled solids (Refs. 228, 229).

2.3.2 Temperature Pretreatment

By subjecting organic materials to excessively high and low temperatures, chemical bonds are broken down to yield a product which is more easily susceptible to hydrolytic activity during anaerobic digestion. Methods used for temperature treatment include boiling, steaming and freezing. Many industrial wastes have already been thermally treated during normal processing (Refs. 150, 230) and require no further heat treatment. However those organic materials which have not been subjected to temperature changes may benefit from temperature pretreatment. Research on temperature pretreatment has focused on lignocellulosic residues (wood plants, some crops, animal manure) and nitrogenous residues (sewage, many crops, animal manure, herbaceous plants, aquatic plants) as two distinct groups, but as revealed in the following discussion temperature effects on biodegradability of all the residues are very similar.

High Temperature

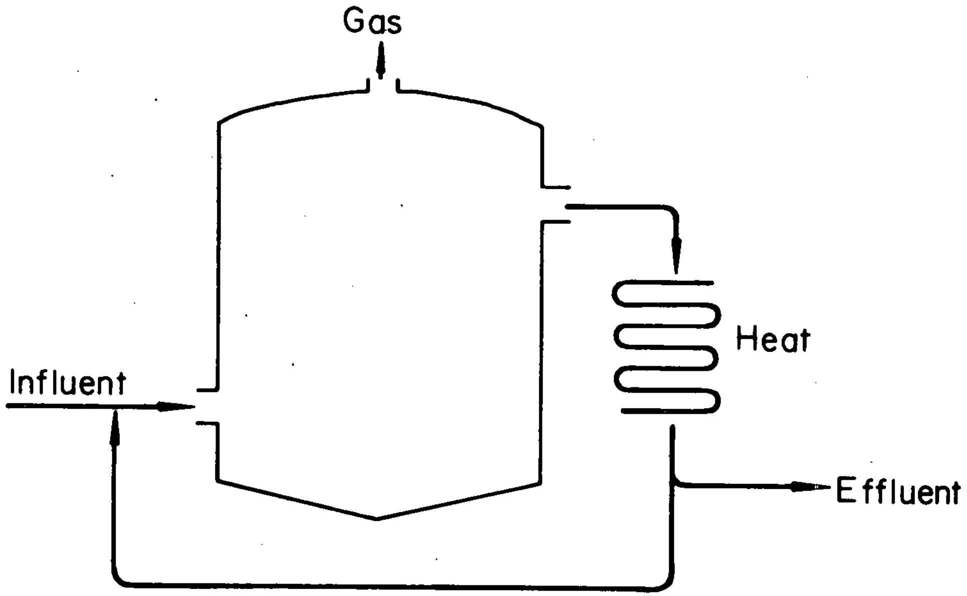
In an early patent (Ref. 231) Buswell and Boruff recommended boiling cellulosic materials for about one hour to obtain high rates of anaerobic decomposition. Later studies by Sharkov and his colleagues (Refs. 232, 233) indicated that long heating of cellulose for 3 hours at 200°C in nonpolar liquid, such as kerosene, or in dry air or nitrogen resulted in a product having a greatly enhanced rate of acid hydrolysis. Millett and Goedken (Ref. 234) reconfirmed the optimum temperature of 200°C, but indicated that further extension of the heating period did not increase the rate of decomposition. After 32 hours of heating at 200°C the maximum increase in rate of dilute acid hydrolysis was about 35%.

An extensive series of studies at Stanford University (Refs. 235-239) have been undertaken to determine optional conditions for heat treatment of lignocellulosic waste materials. Municipal refuse which consists mainly of paper and other wood products was subjected to temperatures from 25° to 250°C, pH from 1 to 13 and for times up to 3 hours. Refuse which had previously been anaerobically digested to remove the readily biodegradable material, was adjusted for pH and heated and the resulting slurry was assayed by measurement of gas production over a three-day period. The results of these short-term bioassays indicated that the optimum temperature for improved biodegradability increased with increasing pH from 130°C at pH 1 to 200°C at pH 13, when using a one-hour contact time.

Treatment at higher temperatures resulted in a decrease in biodegradability, presumably through the polymerization of heat treatment products and formation of humic materials. Semi-continuous digestion studies using 1500 ml mechanically-mixed digesters operated at 35°C and a 15-day retention time were carried out on 1) classified unheated refuse prepared in a slurry with sewage, 2) heat-treated raw refuse, and 3) heat-treated digested refuse. Heat pretreatment was operated at 200°C and under alkaline conditions. A semi-continuous digestion system in which half of the digester effluent was heat-treated and recycled to the digester influent (Figure 21A) resulted in a 49% increase in methane production over the control. It was projected that a heat treatment stage between the first and second stages of a two-phase digestion system (Figure 21B) would possibly result in an additional 73% volatile solids destruction, a 73% increase in methane production and a 36% reduction in volatile solids for ultimate disposal.

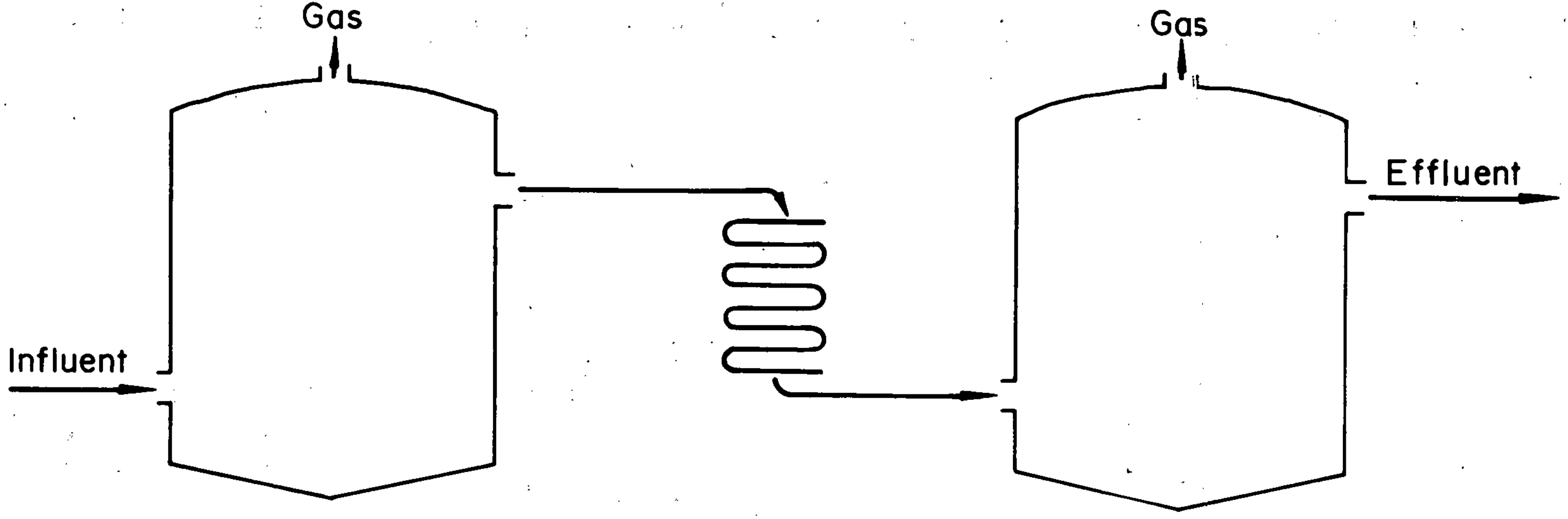
Heat treatment has also been applied by steaming lignocellulosic residues including straw (Refs. 240 - 246) and wood (Refs. 247 - 250). Hardwoods treated by steaming are more easily digested by farm animals than treated softwoods (Ref. 277). Bender and co-workers (Resf. 249, 250) have demonstrated that aspen chips steamed for 2 hours at 100 - 115 psi (160 - 170°C) are readily accepted by sheep at up to 60% of the total ration and provide normal weight gains and carcass yields. Overall processing costs for 3.6 x 10^6 kg/yr of steam treated aspen chips were estimated at about $0.055/kg.

A



HEAT TREATMENT WITH RECYCLE

B



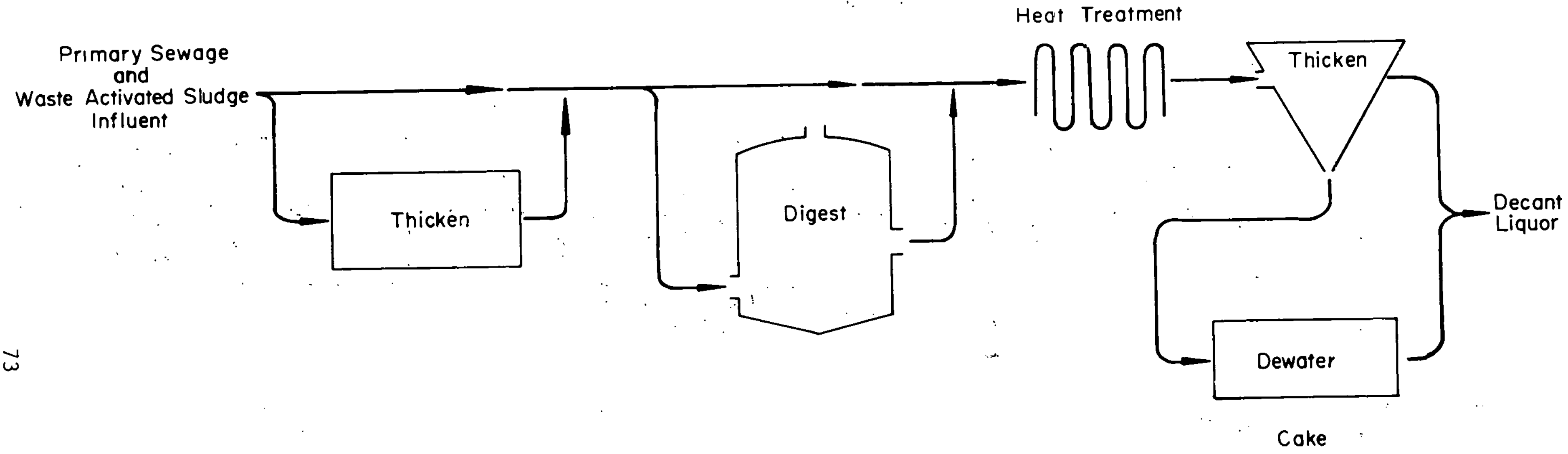
TWO-PHASE DIGESTION WITH A HEAT TREATMENT BETWEEN THE TWO DIGESTERS

Figure 21 A flow diagram illustrating heat treatment used in a digestion system with recycle (A) and a two-stage digestion system (B).

The results of research at Stanford University on the biodegradability of heat-treated waste activated sewage (Refs. 251, 252) were similar to those for thermal treatment of lignocellulosic materials. Significant increases in biodegradability were evident at treatment temperatures of 175° - 200°C. Unlike lignocellulosic materials, however, sewage did not require the addition of chemicals for increased biodegradability. Heat treatment appears to increase degradability by solubilizing nitrogenous components of the cell and also by weakening insoluble organics for subsequent attack by hydrolyzing enzymes.

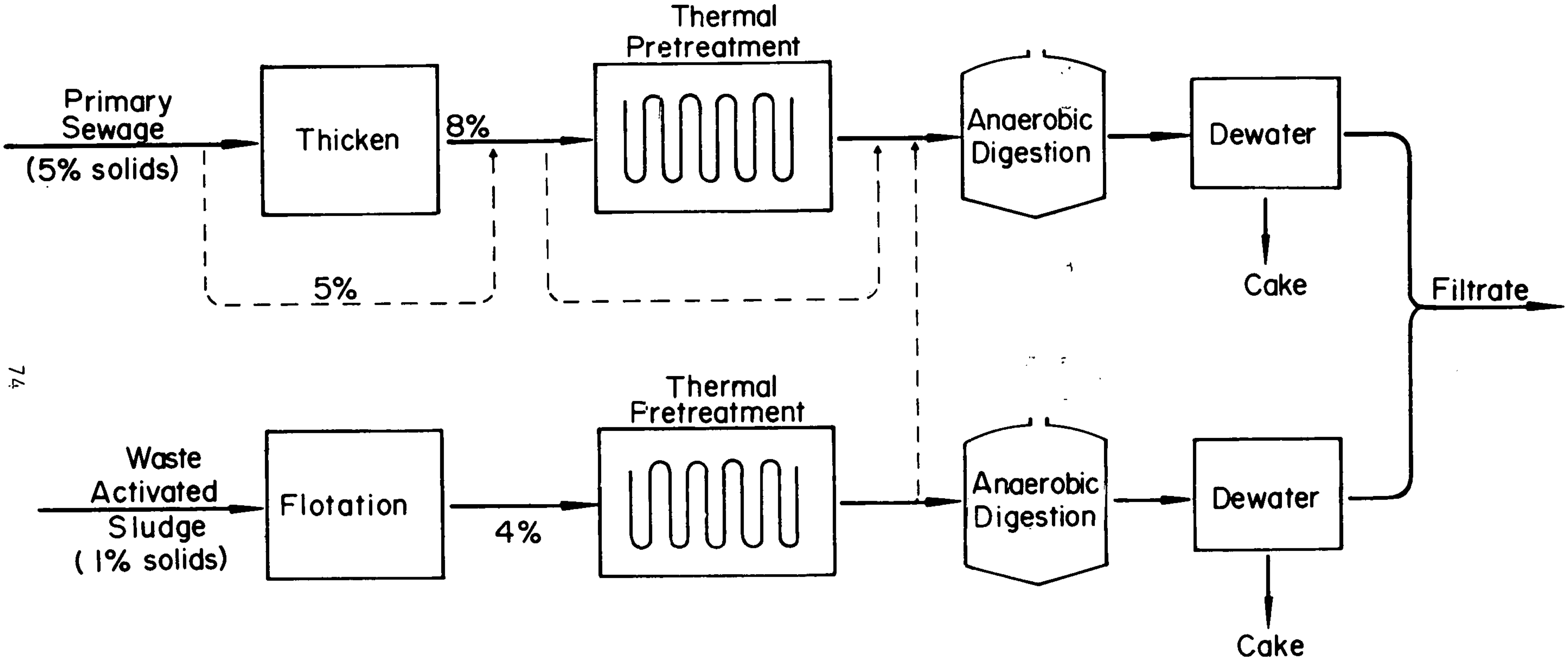
Thermal treatment as used conventionally in waste sewage treatment is illustrated in Figure 22. Heat treatment normally functions as a conditioning process applied to raw or digested sludge prior to dewatering. Haug *et al* (Ref. 253) described such a system in which activated sludge was dissolved, air thickened to about 3% solids and thermally conditioned for 30 minutes at 177°C so that 60 - 70% of the suspended solids were solubilized. Dewatering produced a 30 - 40% cake and the decant liquor was treated in an anaerobic filter. The major advantages of the conventional system are improved dewaterability and sterilization of pathogenic sewage. Improvement in dewaterability as a result of thermal treatment is thought to be caused by a breakdown of the gel structure and reduction of bound water associated with organic and biological solids (Ref. 253).

The disadvantages of the system as outlined by Haug (Ref. 253) are: 1) odors in the decant liquor and sludge cake produced during heat treatment, 2) the decant liquor must be treated before recycling to the secondary treatment plant, 3) large consumption of heat energy and 4) corrosion and organic fouling of heat exchanger tubes. Haug (Ref. 253) proposed a system which would relieve some of these problems by employing thermal pretreatment of primary and/or secondary sewage sludge before the anaerobic digestion stage. A flow diagram of the proposed system is shown in Figure 23. Haug made a comparison of ten systems (Figures 24A-D) employing thermal pretreatment prior to digestion, thermal treatment after digestion and digestion without any thermal treatment. Pretreatment prior to digestion was compared for mesophilic and thermophilic digestion. Assessment of the alternative systems was based upon five variables



HEAT TREATMENT AS CONVENTIONALLY PRACTISED FOR SLUDGE CONDITIONING

Figure 22 Flow diagram illustrating heat treatment as a component of conventional waste sewage treatment.



HAUG'S SYSTEM

Figure 23 A flow diagram of the system proposed by Haug employing thermal pretreatment of primary and/or secondary sewage sludge before the anaerobic digestion stage.

for optimization in the analysis including odor control, sludge dewaterability, liquid sidestream treatment requirements, production of a pathogen-free end product and net energy production. None of the systems optimized all variables, but thermal pretreatment before digestion had special merit.

Systems 1, 2 and 3 in Figure 24A employing digestion without thermal treatment produced considerable energy, controlled odors and liquid sidestreams from dewatering could be recycled directly to the treatment plant; however, a considerable amount of solids remained after digestion, dewaterability was affected by digestion and some pathogens were not controlled. Conventional sewage treatment systems 4, 5, and 6 (Figure 24B) employing thermal treatment after digestion had the advantages of good dewaterability and a sterile end product, but as mentioned previously, required a large expenditure of energy, produced a liquid fraction needing further treatment and had an odor problem.

Systems 7, 8, 9 and 10 employing thermal pretreatment had the advantages of good dewaterability and residual heat remaining in the thermally treated sludge. The end product of systems 7, 8 and 9 (Figure 24C) would not be sterile but would be free of pathogens. Since sludge and liquor would not be exposed to the atmosphere from time of entry into the thermal treatment unit to removal from the digester, there would not be an odor problem. After digestion, remaining sludge volume should be reduced because of increased biodegradability and solubilization of organic acids. Any residual solids that did remain could be dewatered and the liquid sidestream should not exert a significant oxygen demand on secondary treatment facilities if directly recycled. Net energy production of these systems would be slightly less than for system 1, 2, and 3. If a pathogen-free end product was not required, or if another pathogen-control process such as composting was employed, system 10 (Figure 24D) - blending thermally treated secondary sludge and untreated primary sludge for mesophilic digestion - would be a good alternative since net energy production would be optimized, odors would be controlled and dewatering properties would be efficient.

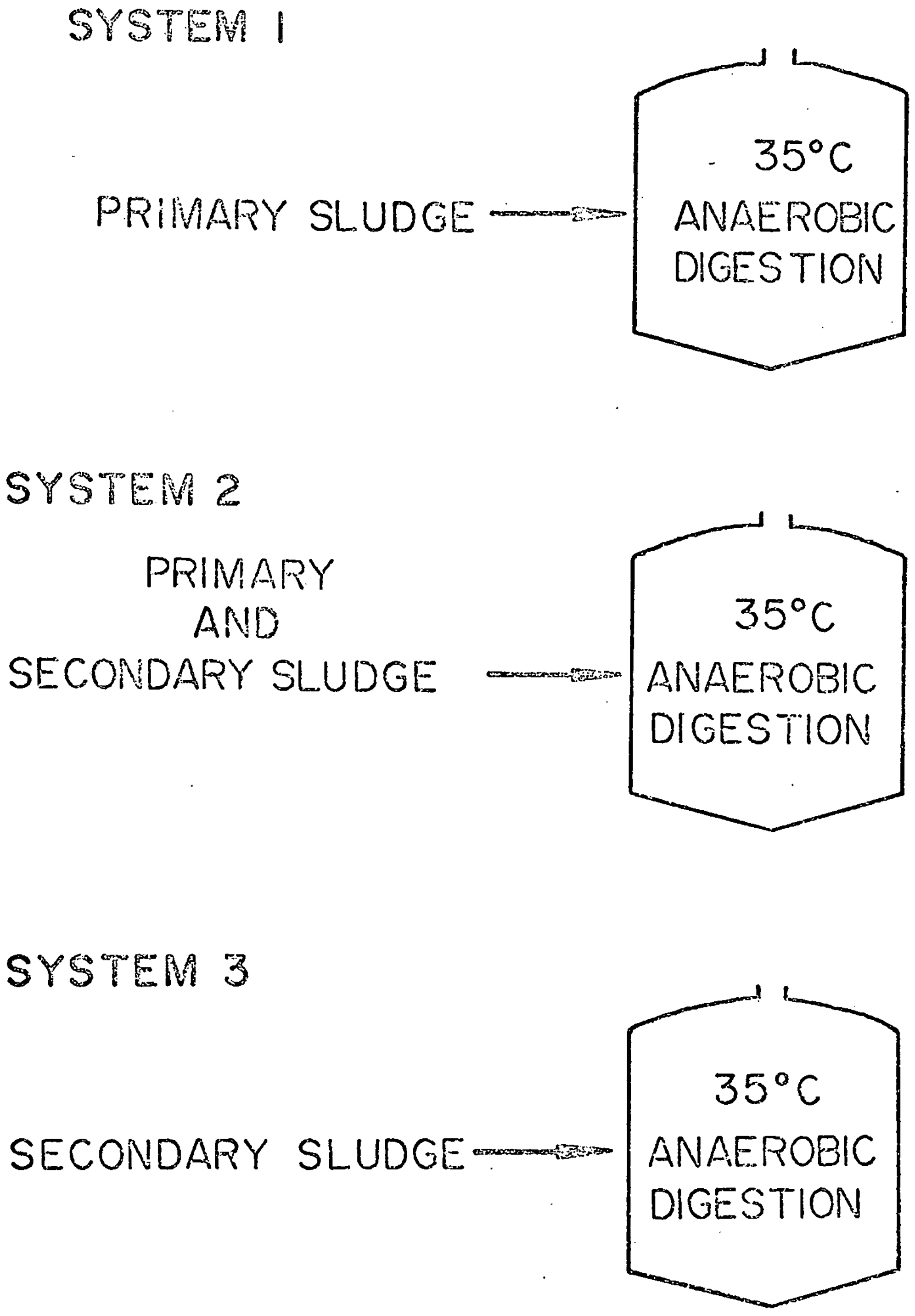


Figure 24A Three systems employing digestion of sludge without thermal treatment.

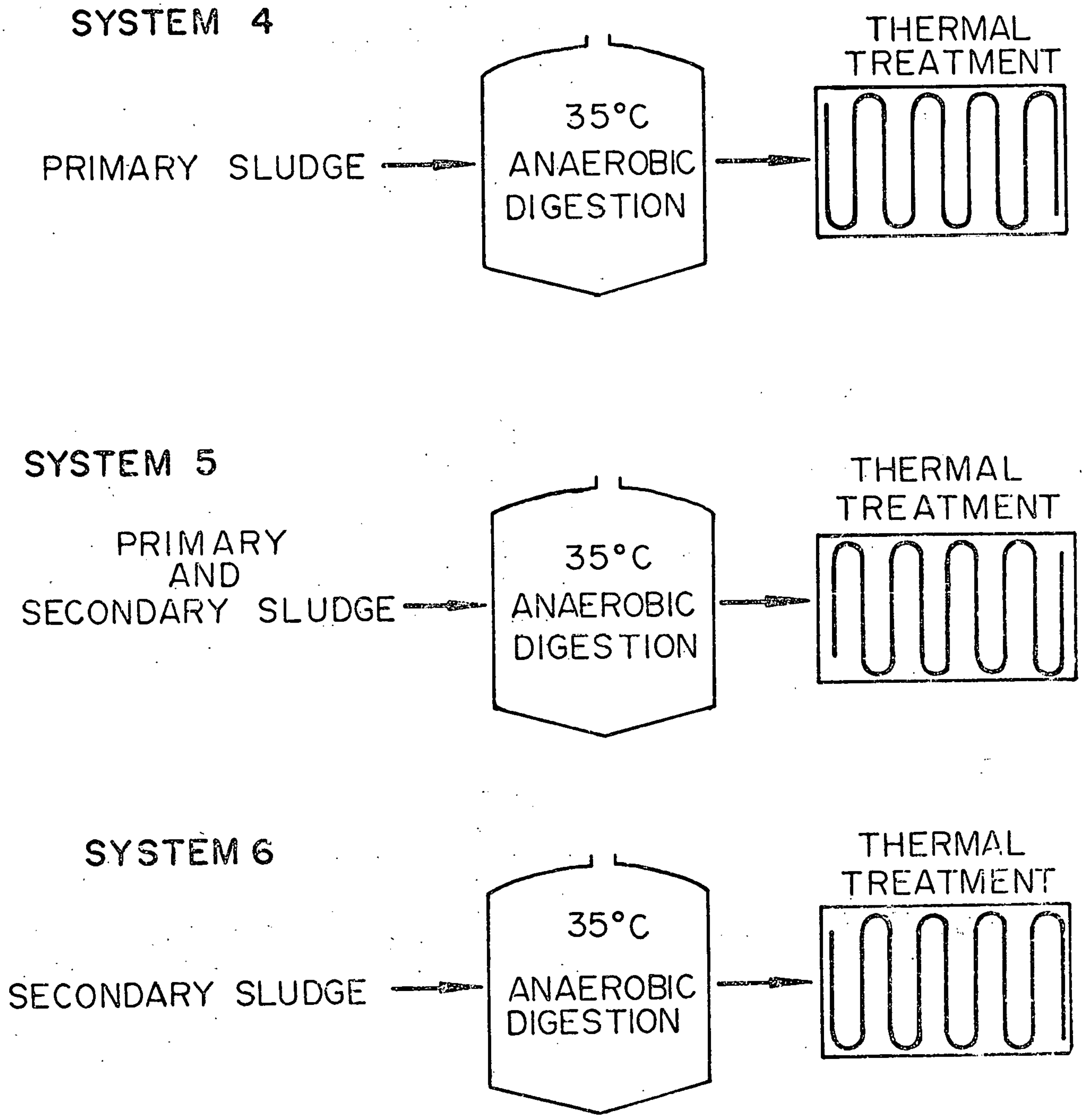


Figure 24B Three systems employing thermal treatment after digestion.

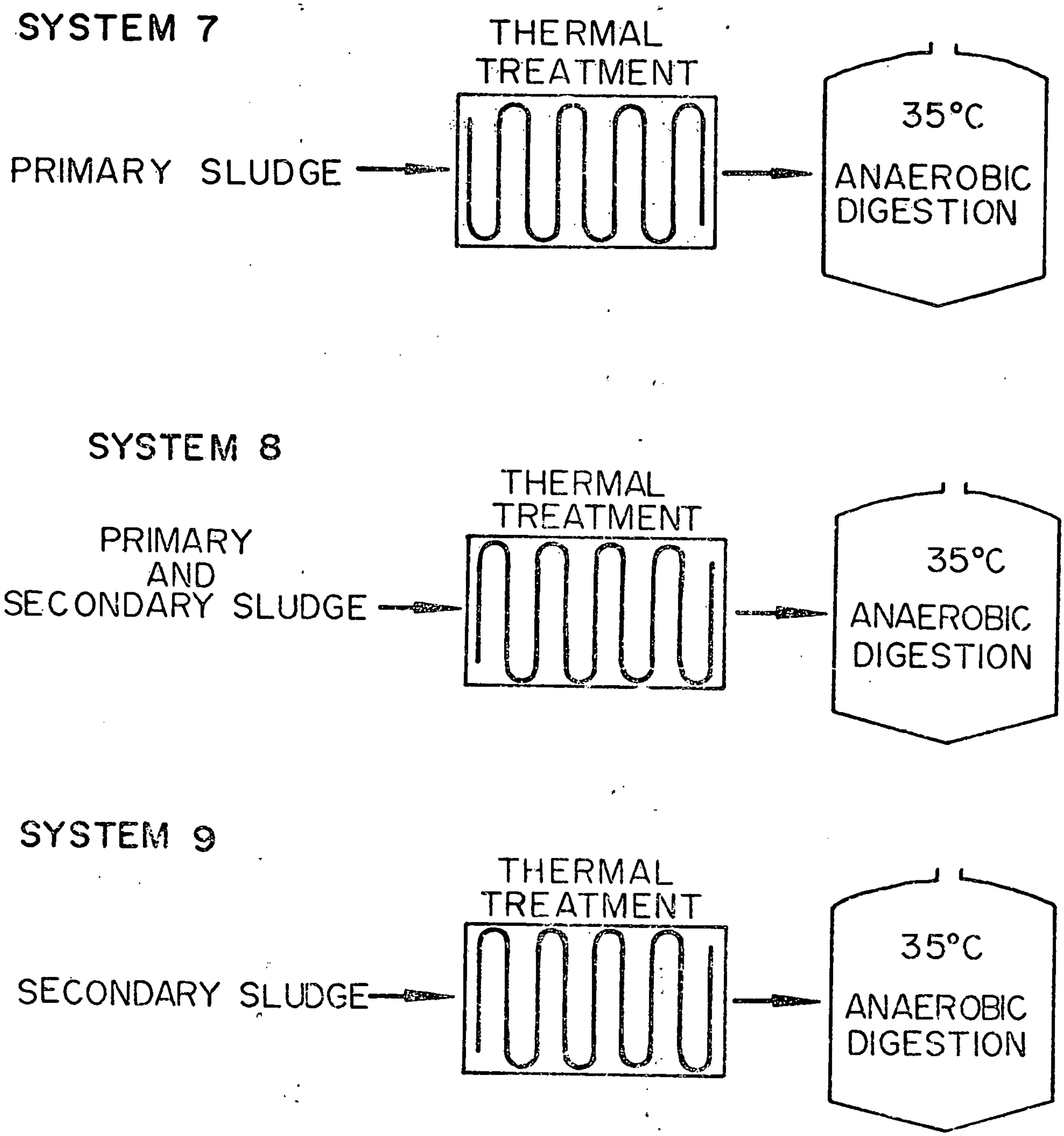


Figure 24C Three systems employing thermal treatment before digestion.

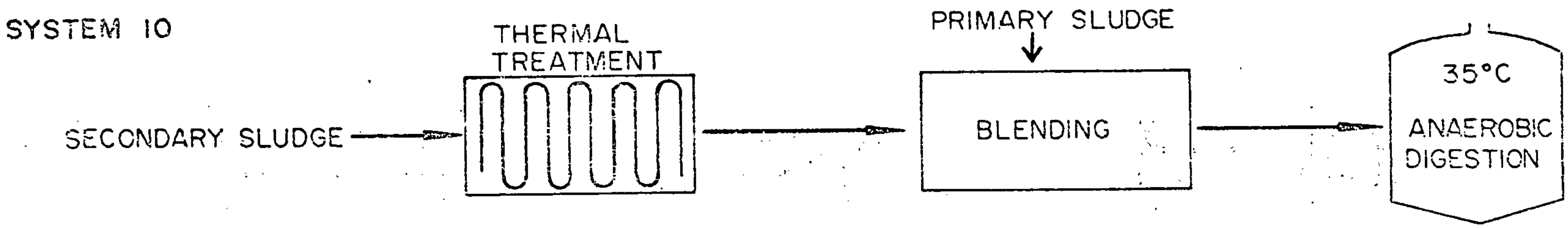


Figure 24D A system in which thermally treated secondary sludge is blended with untreated primary sludge for mesophilic digestion.

Low Temperature

Freezing organic materials in water suspension has also been reported to reduce both strength and degree of polymerization and to increase reactivity as measured by dye adsorption (Ref. 254). The process involving repeated cycles of freezing and thawing is highly energy intensive (Ref. 227).

2.3.3 Pressure Treatment

Pressure pretreatment of organic materials is usually reported in the literature as a function of thermal pretreatment and hence, it is difficult to determine the magnitude of pressure-induced increases in biodegradability. The few reports of materials subjected to pressure treatment without thermal treatment have suggested that pressure plays an important role in the degradation process, however. By compression of a cotton hydrocellulose for 30 min at room temperature, Sharkov and Levanova (Ref. 255) found that the quantity of material dissolved during ethanolysis doubled as the pressure was increased from 0 - 8000 kg/cm^2. Repeated compression of spruce sulfite pulp sheets between calender rolls increased solubility during ethanolysis from 12 - 54%.

Pretreatment of organic materials by milling has been discussed in Section 2.3.1 as a physical pretreatment for size reduction, but it is important to note that increased biodegradability may also be attributed to pressures exerted on cell structure during milling. Odintsov and Beinart (Ref. 256) reported substantial improvement in cellulose saccharification by pressure milling with low ratios of 75% sulfuric acid.

2.3.4 Acid Pretreatment

The acid pretreatment process originated with the research of Porteous (Refs. 257, 258) in an attempt to identify an economically feasible method of municipal refuse disposal. Porteous conducted an economic analysis of the acid hydrolysis of the cellulosic material in municipal refuse to sugar with subsequent fermentation to ethanol, utilizing data developed by Saeman

(Ref. 259) on the hydrolysis of cellulose in wood chips. The analysis indicated that from an economic point of view, it is necessary to carry out the hydrolysis of cellulose in dilute mineral acid at high temperatures. The process incorporating these conditions has become known as the Porteous-Saeman Process.

Based on Saeman's data for wood, Porteous originally predicted a maximum sugar yield of 55% at 230°C with 0.4% acid. In a latter kinetics study, Fagan et al (Ref. 260) predicted a maximum yield of 52% at 230°C with 1.0% acid. The yield at 0.5% acid would be 39%. Development of this latter model was based on experimental results of acid hydrolysis of paper or refuse which had been Wiley ball-milled to 2 mm particles and treated with sulfuric acid. Sugar and residual potential sugar (cellulose) content were recorded for each acid concentration (0.2%, 0.5% and 1.0% by weight sulfuric acid) at temperatures up to 260°C. A schematic diagram of such a Porteous-Saeman Process is shown in Figure 25.

Dilute acid treatment at elevated temperatures functions to hydrolyze the cellulosic fraction of lignocellulosic material, leaving the lignin portion unaffected. Glucose is released rendering the carbohydrate more accessible to enzymatic attack. Glucose is then chemically degraded in a series of dehydration-condensation reactions yielding probably insoluble humic substances (Ref. 239). This process is often used in combination with enzyme hydrolysis.

2.3.5 Alkaline Pretreatment

Alkaline treatment is a common practice in the pulp and paper industry for purification of cellulose materials. Typically the process is operated either at relatively low temperatures (room temperature or below) with the use of strong alkaline liquors; or at higher temperaures (80 - 120°C) using relatively weak alkaline liquors (0.5 - 2% NaOH solutions) (Refs. 261 - 265). The former operation is referred to in the industry as the cold alkaline purification process and the latter as the hot alkaline refining process.

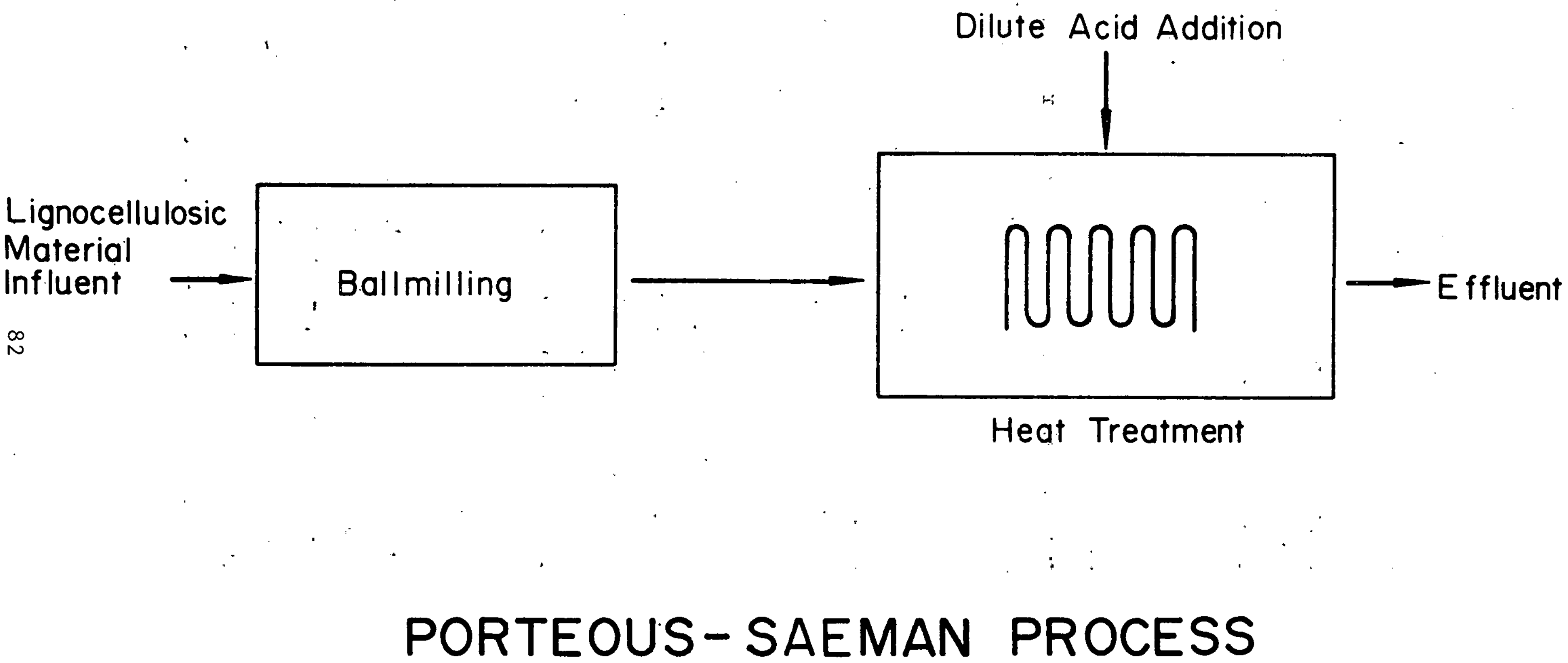


Figure 25 A flow diagram illustrating the process utilized by Porteous and Saeman for pretreatment of lignocellulosic materials.

Complex chemical and physical changes occur in treatment of lignocellulose with alkali. Maksimow (Ref. 265) postulated that in the cold alkaline refining process an intermicellar swelling occurs as caustic soda penetrates inside the cells and dissolves the deeply embedded hemicelluloses. In the hot process, the alkali acts from the surface of the micelles and is able to penetrate deeper only by destroying the cellulose.

Macroscopic physical changes which occur in the alkali-treated wood are well documented (Refs. 225, 266). Treatment of cellulose with sodium hydroxide solutions of mercerizing solutions (above 20%) causes extensive swelling and separation of structural elements. As a result, mercerized cellulose can undergo acid hydrolysis up to 40% faster than cellulose without pretreatment (Ref. 267).

Chemical changes of sodium hydroxide-treated wood have been studied extensively, particularly by Tarkow and co-workers (Ref. 266). Upon treatment of woods with mild alkali, the first and most rapid reactions occurring are the scission or saponification of acetic acid from the acetyl esters of xylan (polypentose fraction of wood) and the hydrolysis of methoxyl groups as methanol from the methyl esters of xylan and the phenolic moieties of lignins. These phenomena are accomplished by the appearance of free acetate ion and methanol in solution, which can be assayed chemically, providing a convenient means for measuring the extent of reaction. In addition, other ester bonds which crosslink cellulose, hemicellulose and lignin, are hydrolyzed. The scission of the esteric crosslinks has been demonstrated by increased calcium ion exchange ability of the alkaline treated wood (Refs. 268, 269).

Basically alkali-pretreatment causes a swelling which increases the size of pore spaces and separation of lignin from the carbohydrates through direct solubilization of lignin to hydroxylated aromatics allowing greater access by enzymes and therefore greater digestibility. Evidence for increased accessibility to high molecular weight materials is given by Feist et al. (Ref. 270) who showed increasing permeability of wood to a series of well-characterized polyethylene glycols of increasing molecular weight. Although the physiochemical changes in plants other than wood have not been investigated, it is probable that the mechanism is similar, namely an opening up of structure and increased accessibility to enzymatic attack.

Alkali treatment of lignocellulosic materials has been used extensively to increase the biodegradation of straw (see reviews by Refs. 271 - 282), hardwoods and softwoods (Refs. 218, 225, 227, 266, 283-285), peat (Ref. 251), municipal refuse (Refs. 235, 236, 239) and animal manure. In general, the material is soaked in 0.1 - 0.15% sodium hydroxide for 1 - 24 hours. This process is based on the original alkaline treatment process patented by Beckman in 1919 (Ref. 286) and used extensively in Europe during both world wars. Another technique which is commonly used was developed in the early 1960's by Wilson and Pigden known as a "dry" process in which the material was steeped in a minimum volume of 20% alkali and then left in situ. This technique overcomes the problem of a loss of about 20% of the organic dry matter during washing operations and subsequent loss of nutrients.

Bioconversion rates have been reported from 15% (Ref. 283) to 85% (Ref. 287) greater for alkaline treated materials than untreated materials digested in rumen fluid at 30°C or less. The degree of efficiency depends on the composition of the material being digested, but there seems to be a digestion ceiling for alkaline-treated cellulose at about 70 - 90%. For example, Bellamy (Ref. 288) found no increase in digestibility when the time of pretreatment of feedlot waste fiber with 0.2% NaOH was increased from 4 hours to 20 hours, although NaOH was present in excess. Wilson and Pigden (Ref. 283) found no increase in digestibility for either wheat straw, or poplar wood when the concentration of NaOH was increased above 7 grams per 100 grams of material, although digestibility increases were linear with alkali addition below this level. As demonstrated in paper pulp manufacture (Ref. 289) significant destruction of the hemicellulose fraction occurs when treatments are severe, and may result in the formation of toxic byproducts (Ref. 290). Research at Stanford University (Refs. 235 - 239) on the biodegradability of chemical-heat-treated municipal refuse showed that peak biodegradabilities, measured by gas production, decreased in order from pH 1 through pH 11, but rose to the highest maximum gas production of all (151 ml STP gas/g COD) at pH 13 at 200°C. These results were obtained from experiments on pretreatment of shredded municipal refuse at temperatures ranging from 25°C to 250°C and from pH 1 to pH 13. More than 90% solubilization of organics was achieved with pH 13 pretreatment at 250°C. McCarty et al. (Ref. 104) demonstrated that heat pretreatment of newspaper under alkaline conditions increased bioconversion to methane from 25% up to 47%. They predicted that the increased convertibility would reduce the quantity of solids requiring subsequent disposal by about 30%.

Although sodium hydroxide is the most commonly used base for alkaline pretreatment processes, other chemicals have been used to control pH, to disrupt the lignocellulosic complex, and to solubilize nitrogeneous materials. There is an extensive number of alkalis that could be employed for pretreatment depending on the chemical composition of the feedstock material and the type of reaction required. A brief discussion of the more prevalent alkalis is presented below.

One of the most important parameters which limits methanogenesis during anaerobic digestion is pH. Based on studies of anaerobic stabilization of sewage, optimum pH values range from 6.4 to 7.4 (Refs. 127, 131, 291 - 294). Outside of that range methanogenesis ceases. Dague (Ref. 295) reported that lime, sodium bicarbonate, potassium hydroxide and ammonia may be used to neutralize the feedstock without being toxic to the digestion process. Addition of lime (calcium hydroxide) is commonly practiced as a pretreatment and digestion treatment to control pH in sewage sludge treatment. The value of using calcium hydroxide was realized in the early 1930's. A patent disclosure by Fischer pointed out increases in efficiency of the digestion process when lime was used rather than sodium nitrate. Another patent relating to disposal of sewage, industrial, and trade wastes (Ref. 296) involved the addition of organic acids of the type of calcium acetate (also ammonium carbonate and sodium carbonate) or lime, the latter of which reacts with acetate acid to form calcium acetate. The use of lime pretreatment has recently been applied to anaerobic digestion of a mixture of seaweeds containing primarily the kelp Laminaria saccharina (Ref. 297). Dried, finely shredded seaweed was pretreated with saturated lime at 37°C for 5 days at pH 11.2 and then neutralized with carbon dioxide until the pH was below 8. Digestion of the pretreated seaweed seeded with sewage sludge in anaerobic continuous stirred tank digesters at 37°C resulted in a higher methane yield than digestion of seaweed that was not treated with lime. The primary advantage of calcium hydroxide pretreatment is the economical cost of the chemical. In addition, since calcium forms a relatively insoluble divalent cation, formation of toxic substances is not a problem, as is the case of high cation concentrations.

In landfills pH may be influenced by industrial waste discharges, alkalinity, clear water infiltration and relative rates of organic acid production and methane generation (Ref. 22). Unlike tank digestion, additional alkalis can not be easily added after the deposition of feedstock into a landfill. Sufficient quantities of lime could not be added to the feedstock during pretreatment to neutralize organic acids formed in intermediate stages of the digestion process without raising pH to toxic levels.

Alternatively pretreatment with calcium carbonate acts as a buffering system throughout decomposition within a "controlled landfill". A feature of anaerobic digestion is that carbon dioxide production by microorganisms is invariably present during both acid and methane formation stages. When an excessive amount of calcium carbonate is introduced into the digestion system, beyond that which can dissolve in the aqueous phase at any organic acid level, carbon dioxide is given off. Organic acid anion formation is balanced by calcium cations going into solution and pH drop is small enough for methane production to proceed. In the reverse situation, when acid anions are consumed and converted to methane, carbon dioxide generated as a consequence of the digestion process, dictates precipitation of calcium carbonate/bicarbonate by solubility product relations. Calcium cations are removed from the aqueous phase, the pH rise is small and digestion continues.

Laboratory studies by Augenstein et al. (Ref. 298) have demonstrated the feasibility of controlling pH by pretreating shredded municipal refuse with calcium carbonate. A total of 27 ℓ of sewage sludge (pH = 5.25; 4% total solids concentration) were mixed with 1200 g $CaCO_3$ for pH adjustment to 6.3 and held overnight; 8 ℓ of digester effluent from a laboratory digester operating on municipal solid waste were then mixed thoroughly with the neutralized sludge to provide an inoculum of microorganisms, and that mixture was mixed with 18.1 kg of shredded municipal refuse using a rotary blade mixer. Digestion of the material in an unstirred digester at 37°C resulted in a conversion of fuel gas up to 0.128 m^3/kg dry sludge/waste solids. This value is to be compared with a methane yield of 0.095 m^3/kg dry sludge/waste solids obtained from the same solid waste source in a continuous stirred tank with a 25-day retention time.

A second experiment was conducted to further buffer pH during digestion using alkaline pretreatment. The shredded waste (18.1 kg) was mixed with 60 ℓ of a 2.5 percent $Ca(OH)_2$ slurry in water, stored for 5 days at 37°C and neutralized to pH 7.8 by sparging 1.1 m^3 of CO_2 through the slurry. After pressing, this mixture was innoculated with digester effluent (8 ℓ), sewage sludge (27 ℓ) and calcium carbonate (670 g) as described previously with a resulting 44.9 percent solids content. A methane yield of 0.13 m^3/kg dry solid waste was reported from unstirred digestion in comparison to a yield of 0.095 m^3/kg solid waste in a stirred reactor. The alkaline pretreatment appears to have merit for controlled landfill operations (Ref. 298).

The concept of a "controlled landfill" implies pretreatment of high solids by shredding, addition of calcium carbonate to buffer the digestion reaction, addition of inoculum to seed the culture and possible addition of nutrients. Other aspects of the controlled landfill are discussed in Section 2.2.1.

Improved digestibility has been reported for chlorine dioxide pretreated wheat straw (Ref. 299) and wood (Refs. 300, 301). Millett <u>et al</u>.(Ref. 227) questions the economic feasibility of this process, however, estimating $200/ton for chemical cost alone.

The ammonium bisulfite process used in conventional pulping results in a sulfite pulp with low lignin content (15%), but with high nutrient content. This delignification process may have application in anaerobic fermentation pretreatment of lignin-containing materials. In acetic acid solution, sodium chlorite solubilizes lignin specifically in the production of "holocellulose", the total carbohydrate portion of lignocellulose (Ref. 302) which has been shown to be highly digestible (Ref. 303).

To increase methane production from sewage sludge, a patent disclosure by Rudolfs (Ref. 304) suggested the addition of sodium aluminate to the feedstock before introduction to the digester. He claimed that small amounts (5.0 ppm $Na_2Al_2O_4$ /2000 ppm sewage) of the chemical either in the form of meta-aluminate ($Na_2Al_2O_4$) or tri-sodium aluminate (Na_3AlO_3) greatly increased the caloric content and percentage of methane in the gas produced. The solids resulting from digestion were more valuable as fertilizer due to a relatively higher nitrogen content.

Pretreatment with aqueous or gaseous ammonia results in extensive swelling of wood and cellulose (Refs. 227, 266). To a degree there is a transformation of the cellulosic structure to cellulose III (Refs. 305, 306). Unfortunately, ammonia pretreatment often results in digester failure, and depending on the species of the feedstock, it may not increase digestibility significantly. The primary advantage of ammonia addition is the increase in nutrient content of the culture. Increases in digestibility attributed to ammonia pretreatment range from 2% for Sitka spruce and red oak to 50% for aspen (Ref. 218). *In vivo* digestion in ruminants has been reported with increases of 6% (Ref. 243) and 28% (Ref. 307) for rice straw.

2.3.6 Gaseous Sulfur Dioxide Pretreatment

One of the most recent investigations of digestibility of wood products at the U.S. States Forest Products Laboratory involves pretreatment with gaseous sulfur dioxide. Rather than disrupting the lignocellulosic complex by conversion and/or removal of either lignin or cellulose, this technique functions to disrupt the complex without altering either component significantly.

As reported by Millett *et al*.(Ref. 227) shredded wood residue diluted with water was subjected to an initial SO_2 pressure at room temperature of 30 psi for 2 hrs (hardwoods) or 3 hrs (softwoods) at 120°C. After blowdown and evacuation the treated residues were neutralized with sodium hydroxide and air dried. The results of the pretreatment process indicated that cellulose digestion was minimal and all lignin was retained. Millett et al. (Ref. 227) suggested, however, that the original lignin was extensively depolymerized and converted to soluble products. Digestibility was increased substantially by the pretreatment process - as much as 60 - 65% for the hardwoods.

Dunlap et al.(Ref. 308) points out the potential advantages of gaseous sulfur dioxide over sodium hydroxide treatment: 1) the gas could be removed easily after treatment and reused, 2) the treated material would be in solid form and solubilized materials could be removed by minimum liquid washing, 3) penetration of the cellulose structure should be rapid and complete, and 4) sulfur dioxide is less expensive than sodium hydroxide.

2.3.7 Aerobic Digestion

Aerobic digestion is the aeration of waste primary sludge, waste biological sludge, or a combination of the two in an open tank. Usually the process is used to stabilize excess activated sludges or the excess sludges from small plants which do not have separate primary clarification. The process also has application in the stabilization of industrial wastewaters, i.e.; citrus fruit processing (Ref. 137) and brewery wastes (Ref. 138). The process involves the direct oxidation of any biodegradable matter by the biologically active mass of organisms and oxidation of microbial cellular material. Major objectives of aerobic digestion include odor reduction, reduction of biodegradable solids and improved sludge dewaterability.

Air is the usual oxidizing agent, but pure oxygen may be used in aerobic digestion to stabilize thicker sludges in which high oxygen uptake rates cannot be satisfied with air aeration. The Purifax process oxidizes sludge with heavy doses of chlorine (2,000 mg/ℓ). Purifaxed sludges may require chemical conditioning prior to dewatering on vacuum filters, since the sludge after treatment has a low pH (about pH 2). The supernatant and filtrate from the process contains high concentrations of chloramines.

In conventional sewage wastewater treatment (Figure 26) the aeration stage is employed between the primary and secondary clarifiers for the purpose of removing suspended solids and to oxidize nitrogen compounds. The supernatant liquor from the primary settling tank is aerated and again settled for separation of clear supernatant and waste activated sludge. Both the primary sludge and waste activated sludge are then thickened before entering the anaerobic digester.

A commonly used variation of the conventional process is the contact-stabilization process (Ref. 309) which takes advantage of the change in physiological properties of microorganisms. In this process the waste stream flows into a contact basin (retention time of 25 - 90 min) in which the organics are absorbed by the sludge floc. The activated sludge then is separated from the mixed liquor in a sedimentation tank and passed into a stabilization basin in which the absorbed organics are metabolized by the organisms. Aeration occurs in the stabilization basin for a retention time of 1.5 - 5 hours depending on the strength of the waste.

A method for the removal of nitrogen from wastes under aerobic conditions has been proposed by Wuhrmann (Ref. 310). In arid regions where the reuse of water is practiced, there is a gradual buildup of nitrogen compounds of which as much as 90% of the effluent is in the form of NO_3^-, NO_2^- and NH_4. Wuhrmann's denitrification process (Figure 27) involves the oxidation of nitrogen compounds to NO_2^- and NO_3^- in an aerobic stage, followed by the reduction of oxides into gaseous N_2 and N_2O in an anaerobic stage.

A variation of the conventional process was set forth in a patent by Koruzo and Mulvaney (Ref. 311). As illustrated in Figure 28, the raw sewage passes through a grit removal unit, a grinder and a flow recorder before entrance into a mixing tank, where the raw sewage is mixed with aerated digested sludge. The mixture is separated in a settling tank with removal of the liquid effluent, and the sludge undergoes anaerobic digestion. Most of the digested sludge is stored for disposal, but a measured quantity of sludge is recycled through an aeration unit. The aeration unit is maintained at a temperature of at least 70°F (21°C), but preferably at 85 - 90°F (29 - 32°C).

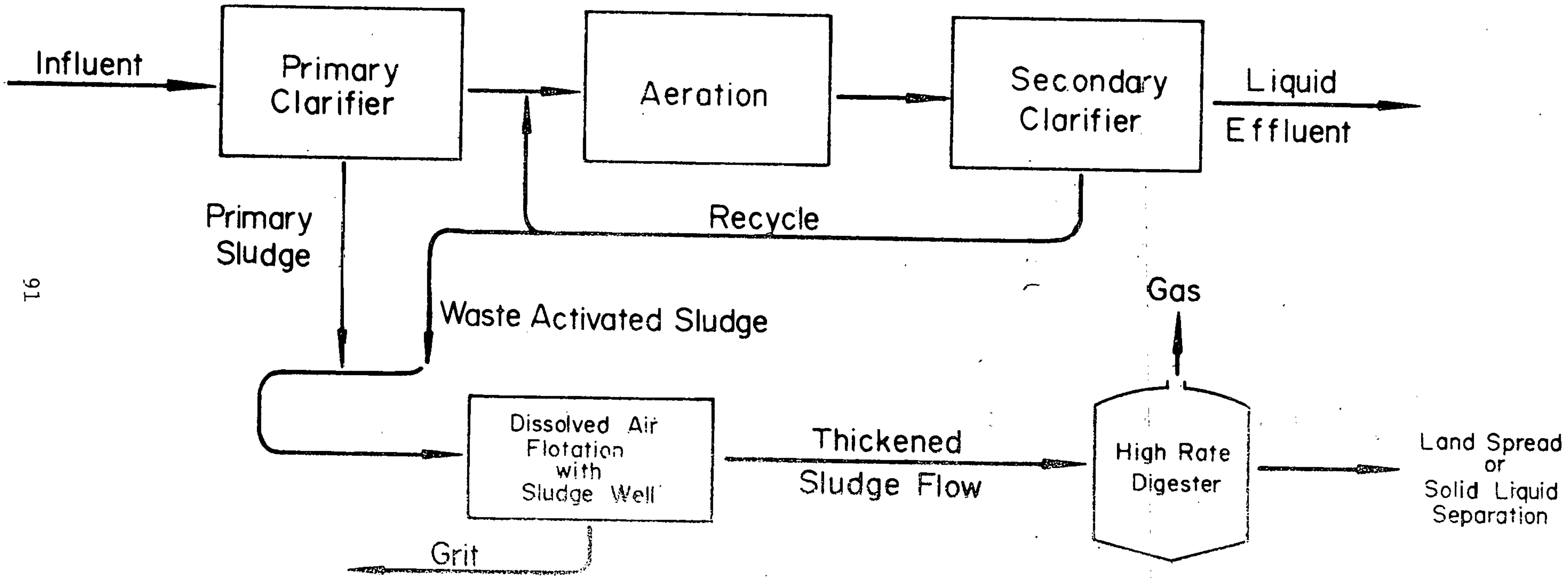


Figure 26 A flow diagram of a sewage treatment system in which the aeration stage is employed between the primary and secondary clarifiers.

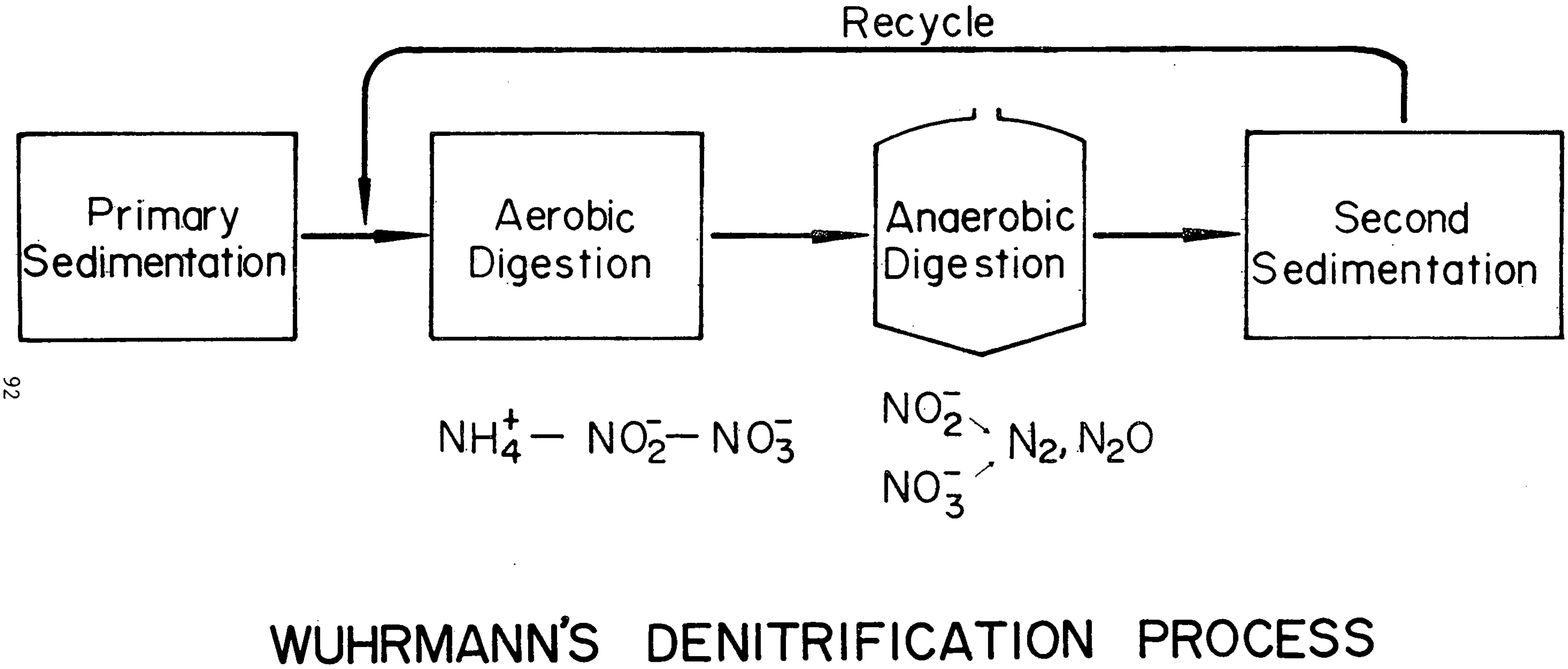


Figure 27 A flow diagram of Wuhrmann's denitrification process involving the oxidation of nitrogen compounds.

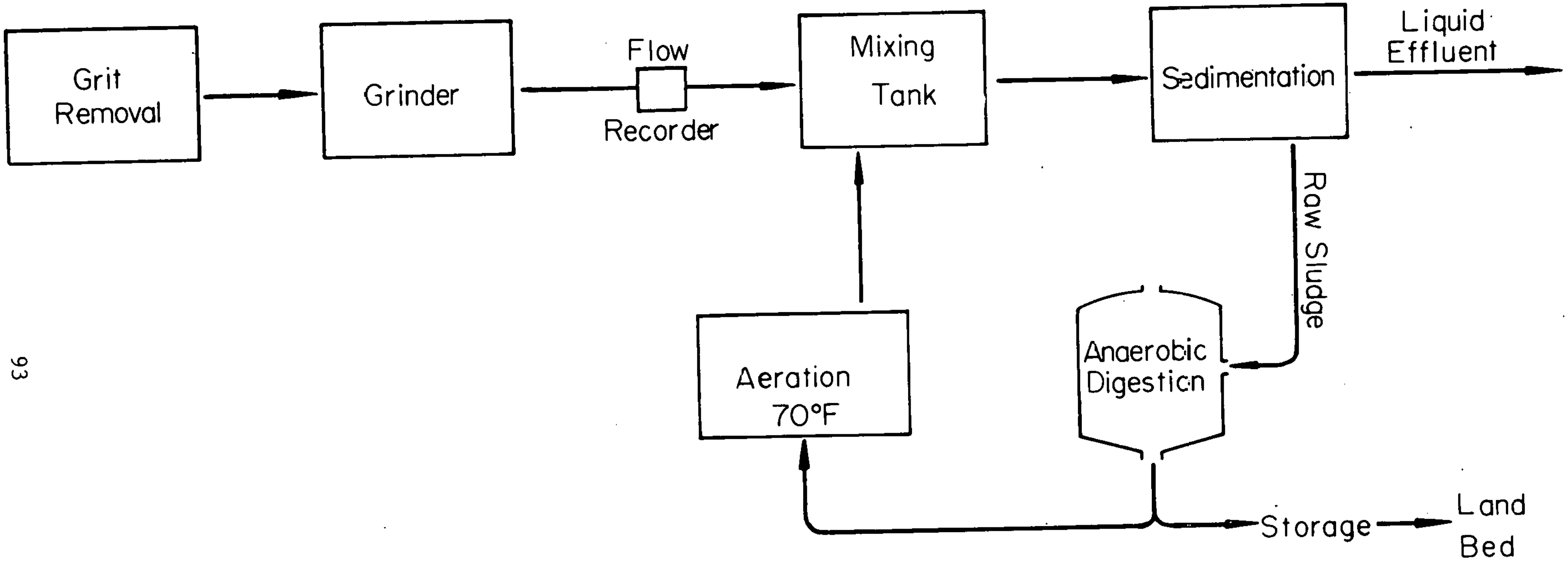


Figure 28 A flow diagram of the sewage treatment system patented by Koruzo and Mulvaney (modified from Ref. 311).

After complete aeration, hydrogen-ion concentrations are maintained in a neutral zone, nitrites usually remain at zero and nitrates up to 300 ppm. The high nitrate content is important so that bacterial oxygenation sets enzymatic action in operation as soon as the inoculum makes contact with the raw sewage. The nitrates also coordinate the carbonaceous cycle with the nitrogenous cycle by buffer action.

A similar system was patented by Hays (Ref. 318) and is illustrated in Figure 29. In this system raw sewage in the primary settling tank is seeded with aerated digested supernatant. The mixture is passed through a series of aeration tanks to coagulate and precipitate suspended and colloidal solids and settling tanks to separate liquid and solid fractions. Sludge from all the aeration and settling tanks is further thickened in a reconcentrating unit before entry into the digestion tank. Liquid removed during thickening is recycled to the primary settling tank or preaeration chamber. Digested sludge from the digester is removed for disposal and the anaerobic supernatant is reduced to an aerobic condition by aeration and alkaline treatment. This liquid is then recycled to seed the primary settling tank with aerobic microorganisms.

2.3.8 Irradiation

Irradiation of straw and various woods by gamma rays or by high velocity electrons substantially increases digestibility of these cellulosic materials by ruminants (Refs. 218, 312-314). Irradiation of cellulose provides increased digestibilities in material with lower degree of polymerization, lower crystallinity, and higher moisture adsorption capacity (Ref. 308). Experimentation of wheat straw and various hardwoods (Refs. 313, 314) indicated that a level of about 5×10^7 rad was needed to increase digestibility significantly. The degree of increased digestibility is highly species specific, however (Ref. 227).

In batch hydrolysis experimentation Saeman *et al.* (Ref. 315) reported that by electron irradiation pretreatment, a substantial increase in rate of hydrolysis and maximum sugar yield were attained at 5×10^8 rad when the cellulosic material became water soluble and lost nearly 45% of its original carbohydrate

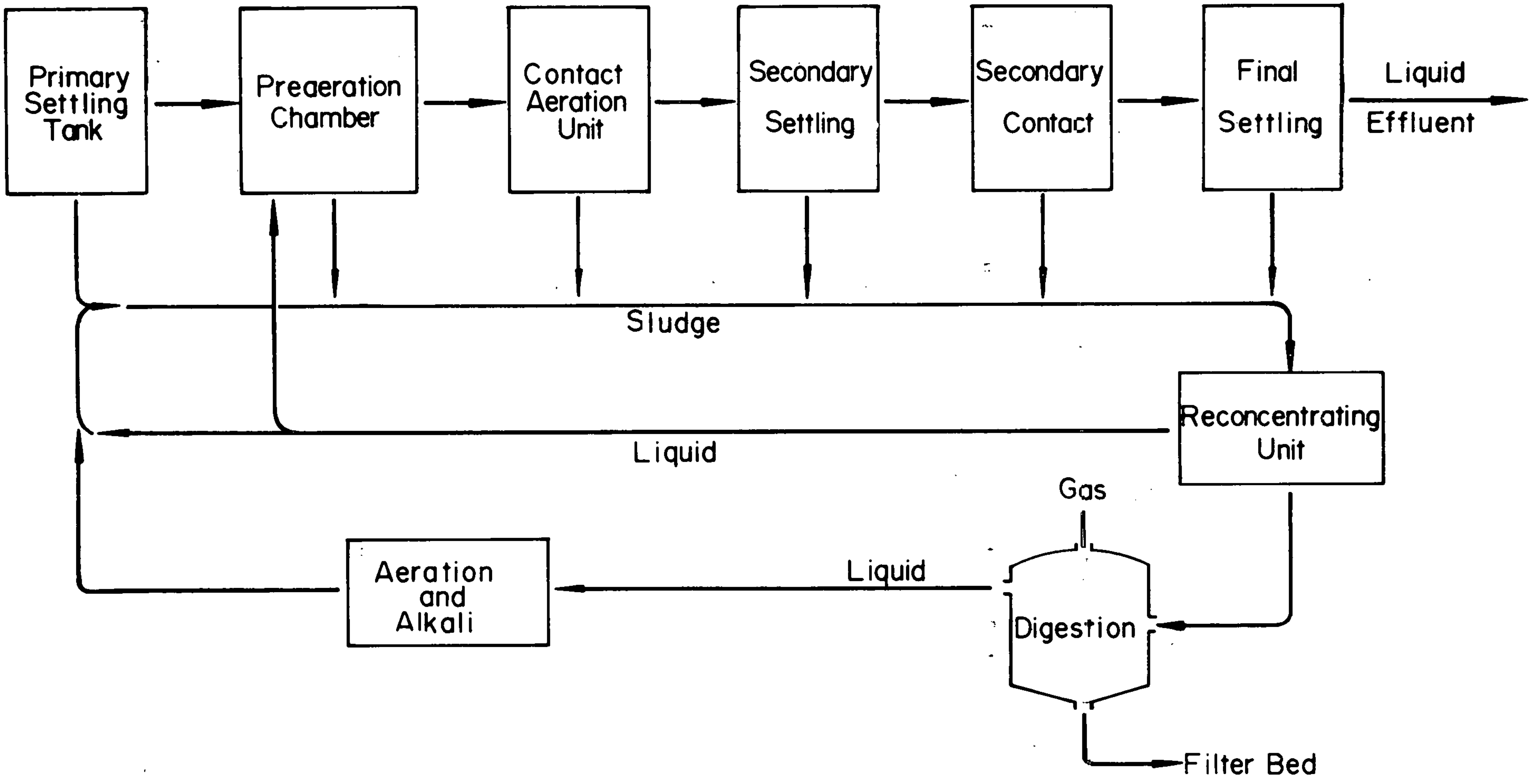


Figure 29 The treatment process patented by Hays (modified from Ref. 318) utilizing aerobic pretreatment.

content. Peak digestibility was reached at about 10^8 rad, at which dosage sugar yield was nearly three times that of the control and the rate of hydrolysis had increased seventeenfold. There was an indication of extensive bond breakage, but only about 14% carbohydrate destruction.

Photodegradation, another radiative pretreatment has been shown to increase biodegradability of cellulosic materials fourfold after a 24 hour irradiation period (Ref. 316). The patented process (Ref. 317) involves exposure of polysaccharides to high intensity ultraviolet light (3650 Å) in the presence of a photosensitizing agent, sodium nitrite.

These irradiation techniques are highly effective for increasing biodegradation. Unfortunately, however, the cost of irradiation equipment is probably out of the practical realm of commerical utilization. Millett *et al*. (Ref. 227) have estimated the cost of gamma and electron irradiation at well over $0.11/kg at a dosage level of 10^8 rad.

2.4 Solids Handling

Experimental research on the uses of anaerobically digested biomass effluent is limited. Based on experience with solids handling of digested municipal and industrial wastes, however, there are several possible methods of disposal and utilization.

In the past one of the major methods for final disposal of sludge has been in lagoons or landfills similar to those described in Section 2.2. When mixed with other landfill wastes, the digested biomass may provide a bacterial inoculum to increase decomposition rates within the landfill. As mentioned in Section 2.2.1, the controlled landfill, once inoculated with methanogenic bacteria and buffered with calcium carbonate, will produce high levels of methane gas.

An alternative disposal method is ocean dumping. Due to increasing regulations to prevent oceanic and shoreline pollution, however, this method is becoming less acceptable.

Recycling of the nutrients in digested biomass to plants or animals is a viable alternative to disposal. The fertilizing properties of fermented manure is well documented (Refs. 320-329). The nitrogen content of cowdung is increased by anaerobic fermentation. Rates of nitrification in soil are often low and crop yields are considerable. Information on the nutritive value of digested plant biomass (Ref. 330) suggests that the digester effluent could be a good fertilizer for crops.

To feed dried digested biomass to animals is another possibility. The majority of information on this topic is limited to manure refeeding (Refs. 331-335). Significant weight gains have been attained with feedings of plant biomass and manure (Ref. 336).

2.5 Gas Treatment

Fuel gas obtained from the anaerobic digestion of biomass usually needs to be treated before being transmitted in existing pipeline systems. The purification scheme involves the removal of carbon dioxide, hydrogen sulfide, and moisture from the digester gas to meet pipeline specifications. An economic analysis was performed comparing the various methods of gas purification (Ref. 337).

A preliminary survey of the available gas purification technologies indicated that viable methods for the treatment of digester gas include physical absorption, chemical absorption, adsorption and membrane separation processes. Most commercially available processes are proprietary in nature and cost estimates had to be obtained from the process licensors. Cost estimates were obtained for the Selexol process, a physical absorption system using the dimethyl ether of polyethylene glycol as the solvent, and the chemical absorption Benfield, Catacarb, and Amine-Guard processes which utilize potassium carbonate, a potassium salt solution, and monoethanolamine, respectively, as the solvents. Cost for the Fluor Solvent process, a physical absorption process utilizing propylene carbonate, was approximated from costs presented in the literature by use of updating and scaling factors. Cost data were also obtained for the molecular sieve adsorption Marsco process and a membrane separation process. In addition, two non-proprietary processes, namely physical absorption water scrubbing and a chemical absorption aqueous phosphate buffer system, were designed by computer simulation.

The results of the analysis of these systems are presented in Figures 30 and 31. It should also be noted that the cost analyses in this study were based on the assumption of no recovery of the CO_2 by-product. It was found that credits for the recovery of the CO_2 could be significant. Since chemical absorption processes can produce higher quality CO_2 by-product stream, they would benefit more from any credit applied to CO_2 recovery.

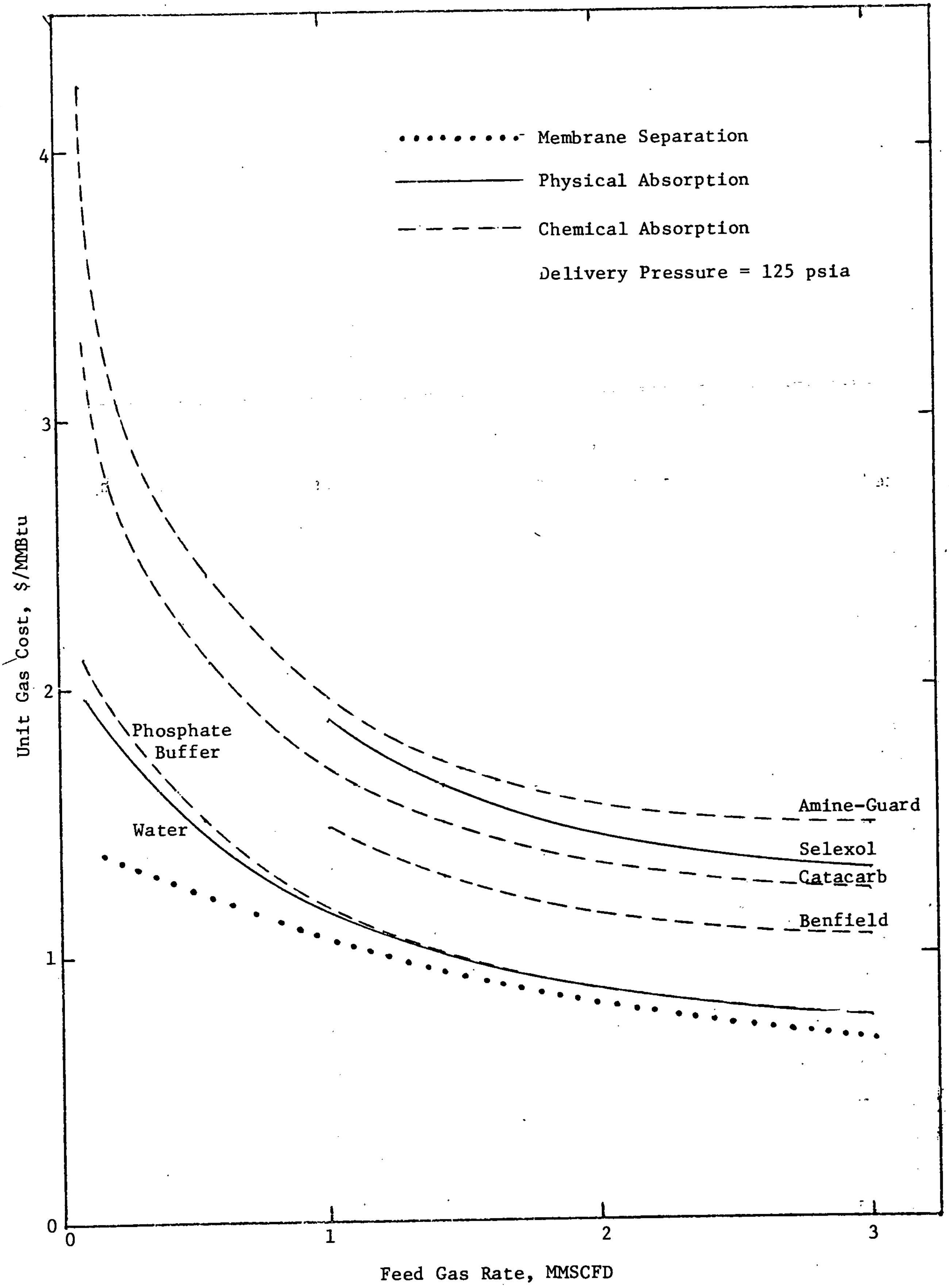


Figure 30. Variation of Unit Gas Cost with Gas Feed Rate for Delivery Pressure of 125 psia

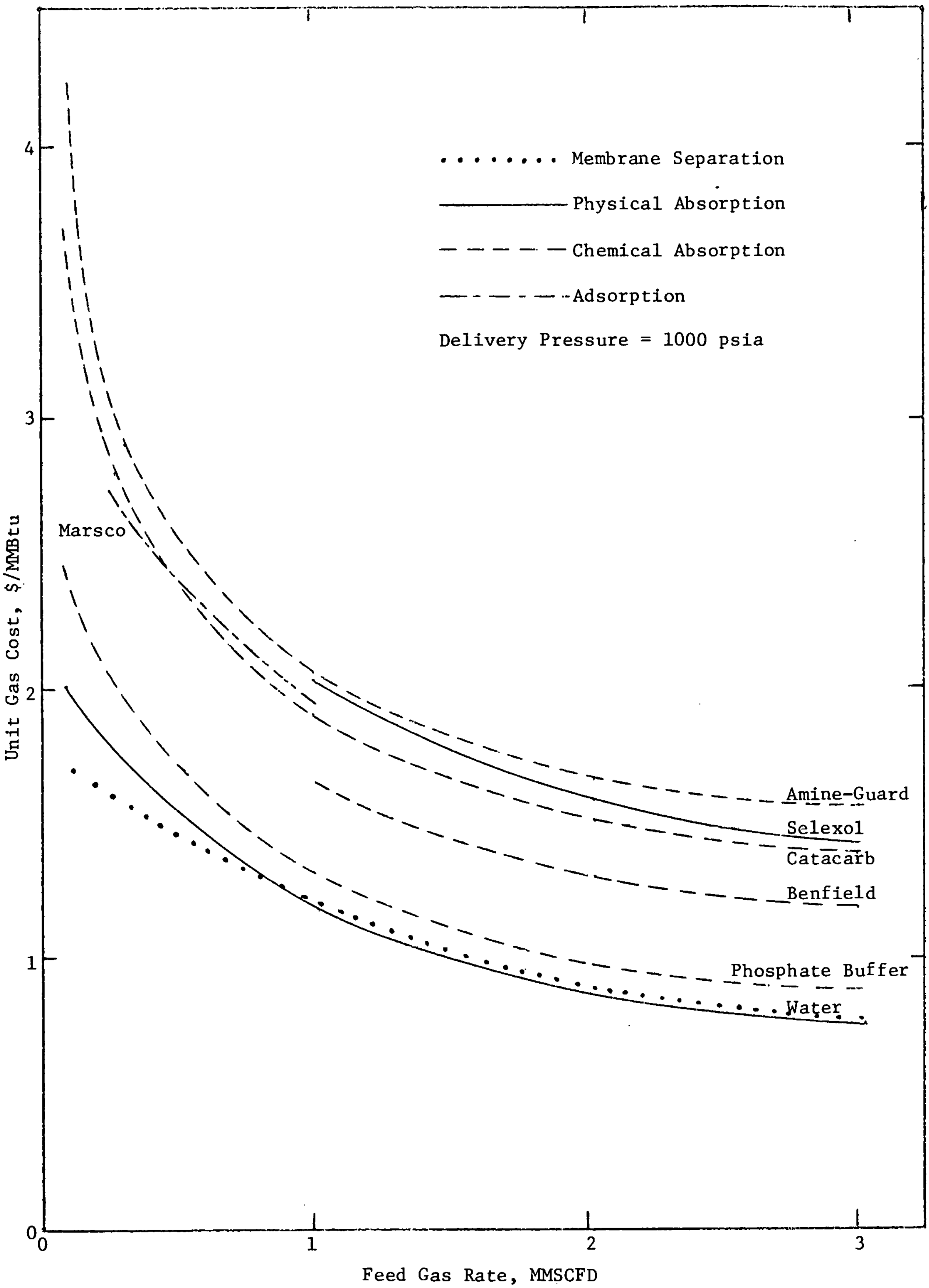


Figure 31. Variation of Unit Gas Cost with Gas Feed Rate for Delivery Pressure of 1000 psia

Section 3

BASIS OF COST ESTIMATE

To ensure an objective and equitable economic evaluation of the various processes, a consistent and uniform set of cost estimating criteria must be applied. The general methods and procedures used for cost analysis in this project are similar to and consistent with those applied in previous economic studies by Dynatech R/D Company on commercial production of fuel gas from anaerobic digestion of various biomass feedstocks.

3.1 Cost Escalation Index

Cost estimates used in this study were based on price levels at the middle of 1978. Most cost data obtained from the literature, empirical cost correlations, and various other sources are often based on price levels at some time in the past. These data must be updated to present costs before they can be used to provide a reliable analysis. This can be done by the use of a cost index. Current costs can be determined by multiplying the original cost by the ratio of the present index value to the index value of the time when the original cost was obtained.

The index used in this study is the Chemical Engineering (CE) Plant Cost Index (Ref. 338). The composite CE Plant Cost Index is based on nationally averaged costs for equipment, machinery, supports, labor, buildings, engineering, and supervision. It is commonly accepted and used in the chemical industry. Based on a value of 100 for 1957-59, the composite CE plant cost index for the base time frame of August 1978 is 220.

3.2 Equipment Cost by Scaling

Digestion systems for the handling of various feed stream sizes were studied in this project. It was necessary to estimate the cost based on data for similar equipment with a different capacity. A commonly used scaling relationship for the estimation of equipment cost is the power factor rule (Ref. 339):

$$\text{Cost of equip. a} = \text{cost of equip. b} \left(\frac{\text{capacity of equip. a}}{\text{capacity of equip. b}}\right)^x \qquad (1)$$

where the scaling factor x is a constant. The value of the scaling factor can vary from less than 0.2 to greater than 1.0 depending on the type of equipment being scaled. In the absence of other information, it is common practice to assume a rule-of-thumb value of 0.6 for the scaling factor. In this study, the power factor rule was used when necessary for estimating equipment costs based on scaling factors considered to be most suitable for the particular equipment being considered.

3.3 Gas Cost Calculation

The calculation of unit gas cost was based on the public utility method developed in 1961 by the American Gas Association (Ref. 340) and modified in 1971 by the Panhandle Eastern Pipeline Company. The procedure was described in a report from Esso Research and Engineering Co. to the Federal Power Commission (Ref. 341). It consisted of general bases for calculating total capital requirement, operating cost, and average unit gas cost.

3.3.1 Capital Cost

The procedure for calculating the total capital cost is outlined in Table 3.1. Capital requirements include all installed onsite plant sections, supporting facilities, contractor's overhead and profits, engineering and design, project contingency, interest during construction, start-up costs, and working capital. No land acquisition cost was included in the calculation of capital requirement. The different components of the total capital cost are discussed below.

Table 3.1

BASIS FOR CALCULATING TOTAL CAPITAL REQUIREMENT

Capital Investment		
All Onsite Plant Sections Installed		XXX
Supporting Facilities	5% of All Pl. Sec.	XXX
Total Capital Investment		XXXX
Contractor's Overhead & Profits	10% of Tot. Cap. Inv.	XXX
Engineering and Design	5% of Tot. Cap. Inv.	XXX
Subtotal Plant Investment		XXXXX
Project Contingency	15% of Sub. Pl. Inv.	XXXX
Total Plant Investment		XXXXX
Interest During Construction	9% x Tot. Pl. Inv. x 1 yr.	XXX
Start-up	20% of Gr. Oper. Cost	XXXX
Working Capital	2% of Tot. Pl. Inv.	XXX
Total Capital Requirement		XXXXXX

Onsite Plant Sections

Costs for all onsite plant sections include process equipment delivered, piping, instrumentation, and installation. Capital costs for pumping and materials handling were found to be about 5% of the total equipment cost for a typical stirred tank anaerobic digestion process (Ref. 342). This fraction has been assumed for this analysis. The cost for piping, electrical, and instrumentation was estimated to be about 4% of the installed equipment cost.

Supporting Facilities

Supporting facilities include equipment for the generation and distribution of power, waste disposal, storage, fire-protection, landscaping, fencing, painting, maintenance and office equipment, outdoor and indoor lighting, communication equipment, and other miscellaneous service items. Most of these items represent incremental addition to supporting facilities required for the operation of the overall fuel gas production plant. It was estimated that supporting facilities require about 5% of the cost of all installed plant sections.

Contractor's Overhead and Profits

Contractor's fees were assumed to amount to 10% of total capital investment. This includes overhead such as field and home office setups during construction, supervision, construction coordination and engineering, insurance, taxes, and other indirect expenses incurred by the contractor as well as the necessary profit for assuming the risks and responsibilities involved.

Engineering and Design

This component of the capital requirement includes the costs for construction design and engineering, drafting, purchasing, accounting, construction and cost engineering, and general overhead involved in the preparation of construction plans and specifications. An allowance of 5 percent of the total direct costs of the process plant was used as fees for engineering and design services (Ref. 339).

Project Contingency

Contingency funds are usually included in the estimate of total capital requirement. In addition to counterbalancing possible errors in estimation, contingency funds are necessary to compensate for unforeseen expenses such as additional pollution control equipment due to change in regulatory rulings, small design changes, unexpected delays, sudden price changes, and others. Contingency usually does not include regular cost escalation due to inflation. In this study, an allowance of 15 percent of subtotal plant investment (see Table 3.1) was included for contingencies.

Interest during Construction

For publicly financed ventures, interest for the borrowed capital must be paid during the construction period. In this study, a 9 percent interest rate on incurred debt was used. An average construction period of 1.875 years is usually assumed by the American Gas Association for construction of coal gasification plants. However, construction and installation of digestion systems which are relatively simple in comparison with coal gasification equipment are not expected to require as much time. Interest allowance for a one year construction period was used in this study.

Start-Up Costs

Modifications and adjustments are often necessary after plant construction has been completed before the plant can operate at maximum design conditions. These changes involve expenditures for materials, equipment, and labor. Funds must be budgeted for these expenses as well as to cover the loss of income while the plant is shut down or is operating at only partial capacity. These start-up costs were assumed to be 20 percent of the estimated annual gross operating cost of the venture in this study.

Working Capital

Funds are needed to cover capital tied up in raw materials and supplies inventories, accounts receivable, available cash for monthly payment of operating expenses, accounts payable, and taxes payable. The necessary working capital was estimated to be 2 percent of total plant investment.

3.3.2 Operating Costs

The procedure for determining the annual net operating costs is outlined in Table 3.2. Operating costs are expenses incurred directly by the operation of the plant. These expenses are usually calculated on an annual basis and include production materials, purchased utilities, labor, administration and overhead, supplies, local taxes and insurance, and credit for any valuable by-product. A service factor of 90% was used in calculating all operating expenses with the exception of process operating labor, for which a 100% service factor was used. Each of the operating cost items is briefly discussed below.

Production Materials

Production materials include raw materials and replacement for expendable supplies necessary for the normal operation of the facility. The primary raw material for the digestion system is the raw feed which was assumed to be available at no cost. Other production materials include chemicals for pretreatment.

Purchased Utilities

Utilities requirements for digestion processes include electric power and steam. Process water and cooling could also be required. The cost of these utilities usually varies widely depending on the amount of consumption, plant location, and source. Unit costs for purchased utilities assumed in this study are described below.

Table 3.2

BASIS FOR CALCULATING NET OPERATING COSTS

Production Materials		XXX
Purchased Utilities		
Electric Power (3.0¢/KWH)		XXX
Steam (\$4/10^6 Btu)		XXX
Labor		
Process Operating	2 men/shift, \$6/man-hr	XXX
Maintenance	1.5% of Tot. Pl. Inv.	XXX
Supervision	15% of Op. and Maint. Labor	XXX
Administration & Overhead	60% of Total Labor	XXX
Supplies		
Operating	30% of Process Op. Labor	XXX
Maintenance	1.5% of Tot. Pl. Inv.	XXX
Local Taxes and Insurance	2.7% of Tot. Pl. Inv.	XXX
Total Gross Operating Cost		XXXXX
Credit for By-Product		(XXXX)
Total Net Operating Cost		XXXXX

Electric power to drive pumps, fans, and compressors was assumed to be available at 3.0¢/kwh. Power requirements for miscellaneous items such as lighting were considered to be negligible.

Heat could be necessary for some digestion systems and its cost was taken to be $\$4/10^6$ Btu. In addition to fuel costs, this price includes fixed charges such as local taxes, insurance, and depreciation for the additional increment of steam generating facilities required.

Labor

Manpower required for the normal operation of the digestion facilities include process operating, maintenance, and supervisory labor. It was estimated that all treatment processes handling 4.5×10^4 kg/d (50 tons/day) of biomass feed require two men per shift, but only for the day shift. Labor cost was assumed to be $6 per man hour and a 100% service factor was used in calculating annual process operating labor cost. The effect of lower and higher labor requirements is investigated in the sensitivity analysis.

Maintenance labor requirement is related to the scale and complexity of the operation. An allowance of 1.5 percent of total plant investment for maintenance labor was assumed.

The amount of supervisory labor needed is directly related to the total amount of operating and maintenance labor. Cost of supervisory labor was estimated to be 15 percent of operating and maintenance labor for digestion systems.

Administration and Overhead

This item of costs involves indirect operating expenses that are required for routine plant operation. These expenses include executive and clerical support as well as general overhead expenditures such as medical services, general engineering, safety services, employee benefits, control laboratories, janitorial services, shops, communications, receiving facilities, etc. Administrative and overhead costs were estimated to amount to 60 percent of total labor requirements.

Supplies

Replenishment of expendable supplies are necessary to maintain normal operation of the plant. Operating and maintenance supplies include miscellaneous items such as charts, lubricants, janitorial supplies, test chemicals, etc. The cost of operating supplies was assumed to be 30 percent of process operating labor while maintenance supplies were estimated to be 1.5 percent of total plant investment.

Local Taxes and Insurance

Local taxes and insurance are charges with magnitudes which vary with the type and location of operation. They were estimated to amount to 2.7 percent of total plant investment for gas treatment systems.

Credit for By-Product

Credits for recovery of the digester effluent, and possibly the acid gas stream, could be significant and might alter the economics of the entire operation. However, no credit for by-products was assumed in this study so that the processing costs for each system could be more easily compared. It should be noted, however, that certain processes are more suited for by-product recovery and this could significantly affect the over-all economics.

3.3.3 Unit Gas Cost

The procedure for calculating the unit gas cost based on the utility financing method used by the American Gas Association is outlined in Table 3.3. The average unit gas cost based on this method is given by:

Table 3.3

GAS COST CALCULATION PROCEDURE
UTILITY FINANCING METHOD*

Basis:

- 20-year project life
- 5%/year straight line depreciation on Total Capital Requirement excluding Working Capital

Essential Input Parameters:

- Debt/equity ratio used to split Total Capital Requirement
- Percent interest on debt
- Percent return on equity
- Federal income tax rate

Derived Parameters:

- Rate Base = Total Capital Requirement less Accrued Depreciation (includes ½ depreciation for given year)
- Percent Return on Rate Base = Fraction Debt × Percent Interest + Fraction Equity × Percent Return on Equity

Calculated Cash Flows in Given Year:

- Return on Rate Base = Rate Base × (Percent Return on Rate Base ÷ 100)
- Return on Equity = (Fraction Equity × Rate Base) × (Percent Return on Equity ÷ 100)
- Federal Income Tax = Return on Equity × (Percent Tax Rate ÷ [100 - Percent Tax Rate])
- Depreciation = 0.05 × (Total Capital Requirement - Working Capital)
- Total Gas Revenue Requirement in Given Year = Return on Rate Base + Federal Income Tax + Depreciation** + Total Net Operating Cost

Gas Costs:

- In given year: Total Gas Revenue Requirement ÷ Annual Gas Production
- 20-year average: Total Gas Revenue Requirement Over Project Life ÷ (20 × Annual Gas Production)

Notes:

* AGA Method as modified by Panhandle-Eastern Pipeline Company and used by Synthetic Gas-Coal Task Force

**Depreciation is split according to the debt/equity ratio and used to pay back debt and equity in annual installments. (Working capital is used to offset unpaid debt and equity at the end of the project life.)

$$UGC = \frac{N + 0.05(C - W) + 0.5[p' + \frac{t}{1 - t}(1 - d)r]\ (C + W)}{G_y}$$

where N = total net operating cost, \$/year
C = total capital requirement, \$
W = working capital, \$
p' = fractional return on rate base [$p' = di + (1 - d)r$]
t = fractional federal income tax rate
d = fraction debt
r = fractional return on equity
i = fractional interest on debt
G_y = annual gas production, MMBTU/year
UGC = unit gas cost, \$/MMBTU

The first term on the right side corresponds to the net operating cost, the second term is due to 5 percent per year straight-line depreciation, and the third term accounts for the return on rate base as well as federal income tax. The sum of these terms gives the total average annual revenue requirement. The unit gas cost is obtained by dividing the average annual revenue requirement by the annual gas production. No escalation of operating cost during the life of the project was assumed.

The following bases were used in this study:

Debt/equity ratio = 75%/25%
Percent interest on debt = 9%
Percent return on equity = 15%
Federal income tax rate = 48%

3.4 Credits/Penalties

In most energy conversion processes, the economics could be significantly affected by credits and/or penalties associated with the feed and various effluent streams. The values associated with these credit/penalty streams could vary from today's estimate when demand or supply is altered by establishment of the energy conversion system (Ref. 343). This will obviously affect the final unit gas cost, but not the actual conversion cost. In order to eliminate the uncertainty of these factors, the economics presented in this report do not include any credits/penalties to determine the unit gas cost for conversion of biomass to methane. Specifically, the cost of biomass is taken to be $0. as is the value for digester effluent or any other byproduct. If the reader is interested in determining the effect of such credits/penalties, it would be a simple matter to calculate such costs.

Section 4

SELECTION OF SYSTEM PARAMETERS

The digestion process was analyzed by combining material balances and energy balances, as given in Appendices C and D in order to size the equipment. Costs for the process were then determined from the relationships in Appendix A and the economic routine given in Section 3 was then used to obtain the unit gas cost. A computer program was developed to analyze the many possible combinations. A listing of the computer program is presented in Appendix E, with a listing of all the variables utilized.

The various systems discussed in Section 2 can be combined in many ways to provide numerous processes for conversion of biomass to methane, and these could be utilized for conversion of different feedstocks. However, the purpose of this analysis is to indicate the economic impacts of the various processing steps and to determine the differences associated with the various digester concepts. Therefore, only selected examples were used to indicate these effects. The following indicates which systems and variables were selected and the reasons for the selection.

4.1 Feedstock

The feedstocks considered in this analysis are environmental feedlot manure, dirt feedlot manure, rice straw, and bagasse. The compositions assumed for these are presented in Table 4.1. These were chosen to include low and high moisture content, ash content, and biodegradable content. Both the crop and animal residues included are ones with a large potential availability.

4.2 System Size

The economics of the conversion process will usually improve with increasing size. The system sizes were determined by the feedstock rate, which ranged from 4.54×10^4 to 9.08×10^6 kg/day (50 to 1000 tons/day) on a dry basis. The lower value is equivalent to the solids output from a 10,000 head environmental feedlot.

Table 4.1

COMPOSITION OF FEEDSTOCK

Feedstock	Moisture Content (%)	Volatile Solids Content (% of Solids)	Biodegradable Volatile Solids Content (% of Solids)	Reference
Environmental Feedlot Manure	87	80	44	342
Dirt Feedlot	30	50	22	342
Rice Straw	10	82.5	40	344
Bagasse	50	83	64	345

The sizes of several pieces of equipment are also dependent on residence (or retention) time in that equipment. The residence time for digestion was determined from the fractional conversion of biodegradable solids, ranging from 50 to 95%. The residence time for chemical pretreatment was 2 hours, and for storage and premixing it was 2 days.

4.3 Digestion Conditions

The rate of anaerobic digestion was assumed to follow first order kinetics as presented by Ashare _et al_ (Ref. 342). Digestion temperatures of 37°C and 60°C were used, i.e., both mesophilic and thermophilic conditions were considered for each reactor concept with the exception of the "land-fill" type batch digester for which only 37°C was used.

4.4 Digester Concepts

The various digester concepts and other process options which were analyzed using the computer program are presented in Table 4.2. The other digester concepts indicated in Table 1.1 are analyzed by extrapolation and/or comparison with the computer analyzed results.

4.5 Input Variables

The values for the process and system variables for the analysis are presented in Table 4.3. These are the values utilized with the computer program. Several values utilized should be specifically discussed. The operating labor requirement of 2 men per shift, 8 hrs is used for only one shift per day. The analysis does not include any credits/penalties for feed stream or effluent streams, so the resultant unit gas costs are for the conversion of biomass to methane.

Table 2

PROCESS OPTIONS ANALYZED

Feedstock	Shredder	Degritter	Chemical Pretreatment	Storage	Digester	Heat Exchanger	Dewatering	Gas Purification
Environmental Feedlot Manure	N	N	N	Y	CSTR	Y	N	N
Dirt Feedlot Manure	N	Y	N	Y	CSTR	Y	N	N
Dirt Feedlot Manure	N	Y	N	Y	Plug	Y	N	N
Dirt Feedlot Manure	N	N	N	N	Batch	N	N	N
Rice Straw	Y	N	N	Y	CSTR	Y	N	N
Rice Straw	Y	N	Y	Y	CSTR	Y	N	N
Rice Straw	Y	N	Y	Y	Plug	Y	N	N
Rice Straw	Y	N	Y	N	Batch	N	N	N
Bagasse	N	N	N	Y	CSTR	Y	N	N
Bagasse	N	N	Y	Y	CSTR	Y	N	N
Bagasse	N	N	Y	Y	Plug	Y	N	N
Bagasse	N	N	Y	N	Batch	N	N	N
Bagasse	N	N	N	N	Batch	N	N	N
Bagasse	N	N	Y	Y	CSTR	Y	Y	N
Bagasse	N	N	Y	Y	CSTR	Y	N	Y
Bagasse	N	N	Y	Y	CSTR (in series)	Y	N	N

N = Process Option Not Utilized

Y = Process Option Utilized

Table 4.3

INPUT PROCESS AND SYSTEM VARIABLES

Variable		Value
A	ft^3 CH_4/lb converted	5.6
AW	Labor cost ($/hr)	6.0
CCHM	Cost of pretreatment chemical ($/ton)	300.
CDWV	Credit/penalty for dewatered output ($/ton)	0.
CEFV	Credit/penalty for digester effluent ($/ton)	0.
CEP	Cost of electricity (¢/kwh)	3.0
CF	Heat capacity of digester slurry (Btu/lb°F)	1.0
CHEM	Amount of pretreatment chemical per solids converted	0.2
CONS	Cost constant for equipment	See Appendix A
CPRMC	Maximum capacity of chemical pretreatment equipment (ft^3)	10000
CPRWK	Work for chemical pretreatment (hp/ft^3)	0.
CRF	Cost of feedstock ($/ton)	0.
CST	Cost of steam ($/MM BTU)	4.0
DGRMC	Degritter maximum capacity (tons/hr)	6000
DGRWK	Degritter work (HP)	2.0
DIGMC	Digester maximum capacity (ft^3)	See Note 1
DIGWK	Digester work (HP/ft^3)	See Note 2
DIR	Interest on debt (fraction)	0.09
DNS	Digester slurry density (lb/ft^3)	See Note 3
DPRC	Rate of depreciation (fraction)	0.05
DWTMC	Dewatering maximum capacity (tons/hr)	128
DWTWK	Dewatering work (HP/ton/hr)	3.15
EA	Activation energy for Arrhenius rate equation (cal/mole)	15180
ECHP	Efficiency of chemical pretreatment (fraction)	0.5
EDG	Efficiency of degritting (fraction)	1.0
EDWT	Efficiency of dewatering (fraction)	0.95
EFFS	Efficiency of steam utilization (fraction)	0.80
EX	Equipment cost scale exponent	See Appendix A
FBVSF	Fraction of bio. vol. solids in feed	See Table 4.1

FDBT	Fraction debt	0.75
FITR	Federal income tax rate (fraction)	0.48
FSDF	Fraction of solids in digester feed	See Table 4.1
FSDWT	Fraction of solids in dewatered stream	0.25
FSF	Fraction of solids in feed	See Table 4.1
FSVF	Fraction of vol. solids in feed	See Table 4.1
HEXMC	Heat exchanger maximum capacity (ft^2)	10000
HRTCP	Retention time for chemical pretreatment (hr)	2.0
HRTST	Retention time for storage (days)	See Note 4
HTC	Heat transfer coefficient in heat exchanger (Btu/hrft^{2}F)	80.
OIND	Cost index for year of equipment cost estimation	See Appendix A
OINDX	Current cost index	220.
PAH	Fraction of labor cost for administration and overhead	0.6
PCOP	Fraction of equipment cost for contractor's overhead and profit	0.1
PED	Fraction of equipment cost for engineering and design	0.05
PIDC	Fraction of total plant investment for interest during construction	0.09
PLTI	Fraction of total plant investment for local taxes and insurance	0.027
PM	Fraction of total plant investment for maintenance labor	0.015
--	Fraction of total plant investment for maintenance supplies	0.015
POPS	Fraction of operating labor cost for operating supplies	0.3
PPC	Fraction of subtotal plant investment for project contingency	0.15
PPMP	Fraction of equipment cost for materials handling equipment	0.05
PSEP	Fraction of equipment cost for electricity and piping	0.04
PSF	Fraction of equipment cost for support facility	0.05
PSUP	Fraction of annual gross operating cost for plant start-up	0.20
PSV	Fraction of labor cost for supervision	0.15
PWC	Fraction of total plant investment for working capital	0.02
RE	Return on equity	0.15
RKO	Rate equation constant (day^{-1})	5.92 x 10^9
RNUM	Number of digesters in series	4
SHRMC	Shredder maximum capacity (tons/hr)	25.

SHRWK	Shredder work (HP/tons/hr)	15.
STF	Stream factor (fraction of year in operation)	0.9
STRMC	Storage maximum capacity (ft^3)	300000
STRWK	Storage work (HP/ft^3)	0.0002
TA	Ambient temperature (°C)	10.0
TGRAD	Temperature difference in heat exchanger (°C)	5.0
TMP	Manpower requirement	2.0
TPF	Temperature of feed streams 9°C)	10.0
WD	Hours per work day	8.0

Note 1 Digester Maximum Capacity (ft^3)

CSTR	1×10^6
Plug Flow	4×10^4
Batch	1×10^7

Note 2 Digester Work (hp/ft^3)

CSTR	0.0002
Plug Flow	0.0
Batch	0.0

Note 3 Digester Slurry Density (lb/ft^3)

CSTR	64
Plug Flow	64
Batch	45

Note 4 Storage Retention Time (days) with:

CSTR	2.0
Plug Flow	1.0
Batch	0.0

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Section 5

SYSTEMS ANALYSES

The various digester conceptual designs were analyzed by using the material and energy balances and cost equations presented in the Appendices and the economic routine presented in Section 3. The computer program presented in Appendix E was utilized for the analysis of CSTR, CSTR in series, plug flow, and batch digester concepts for feedstocks of animal residue, rice straw, and bagasse. The systems analyses are presented in this section. Comparisons are made of the various digester concepts by indicating differences in unit gas costs and the reasons for such differences. In addition, the effects of other process options on unit gas cost are presented. A sensitivity analysis of the effects of changing important system variables is also performed.

5.1 Comparison of Digester Design Concepts

The systems analyses are presented in Figures 32, 33, and 34 as unit gas cost (\$/GJ) vs feed rate (kg/day) for feedlot manure, rice straw, and bagasse respectively. The numerical computer outputs for these analyses are presented in Appendix F. These analyses utilize the assumption that the processes are operating without any interruptions.

For each type of feedstock, the plug flow concept results in the lowest unit gas cost. A comparison with the CSTR system indicates several reasons for the difference in unit gas cost. The retention time for a given fractional conversion is highest for the CSTR design concept. (Relationships of retention time and fractional conversion are shown in Table 5.1.) Thus, the CSTR system will have a larger digester volume than the plug flow system. A comparison of Tables F.8 and F.12 shows a retention time of 4.6 days for plug flow and 6.1 days for CSTR with digester volumes of 600 m^3 (21000 ft^3) for plug flow and 1600 m^3 (56,000 ft^3) for CSTR. It should also be noted that the plug flow system has a higher fractional conversion of biomass even though it has a shorter retention time. This results in a higher gas production.

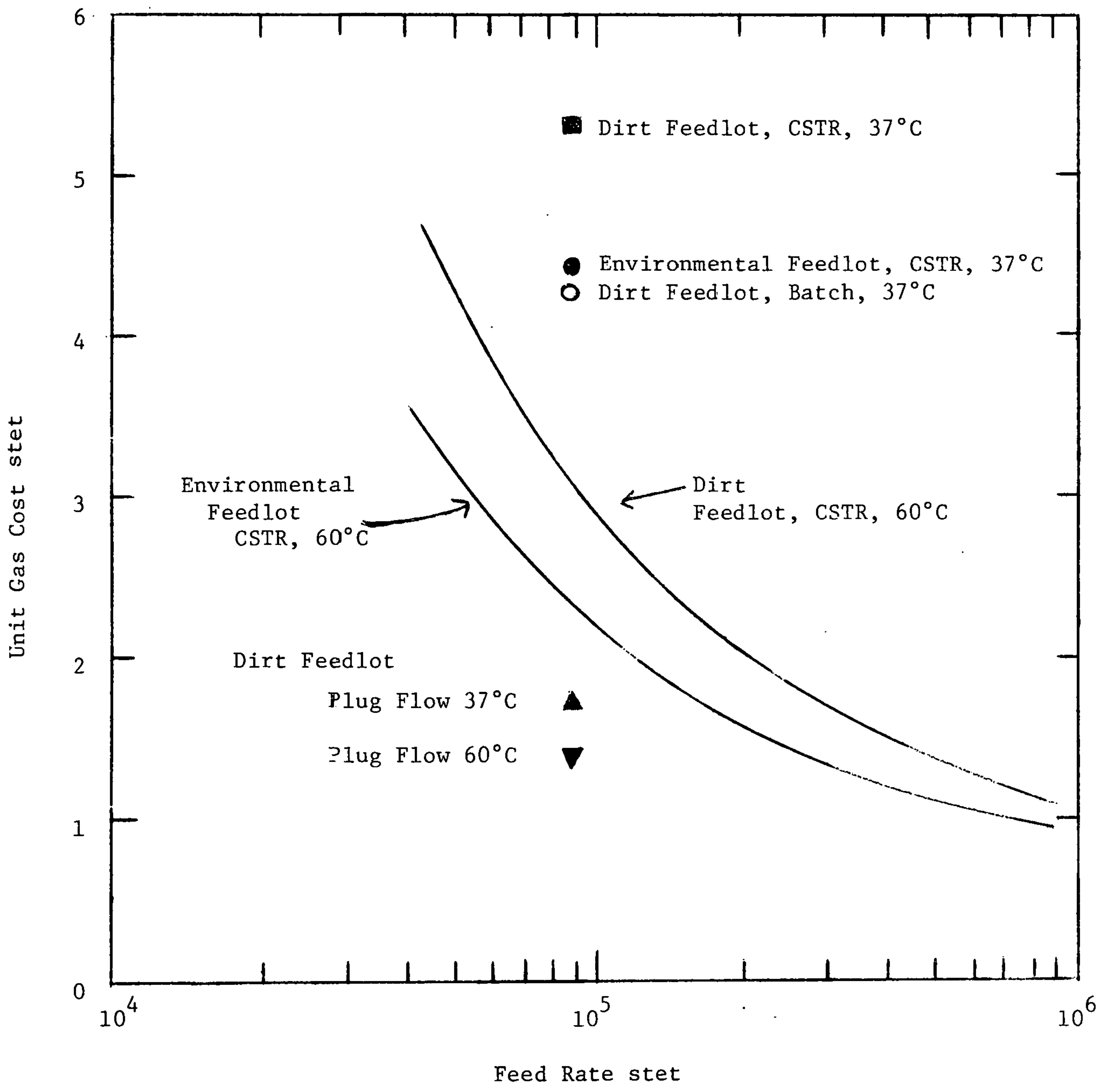


Fig. 2 Unit Gas Cost for Digestion of Feedlot Manure

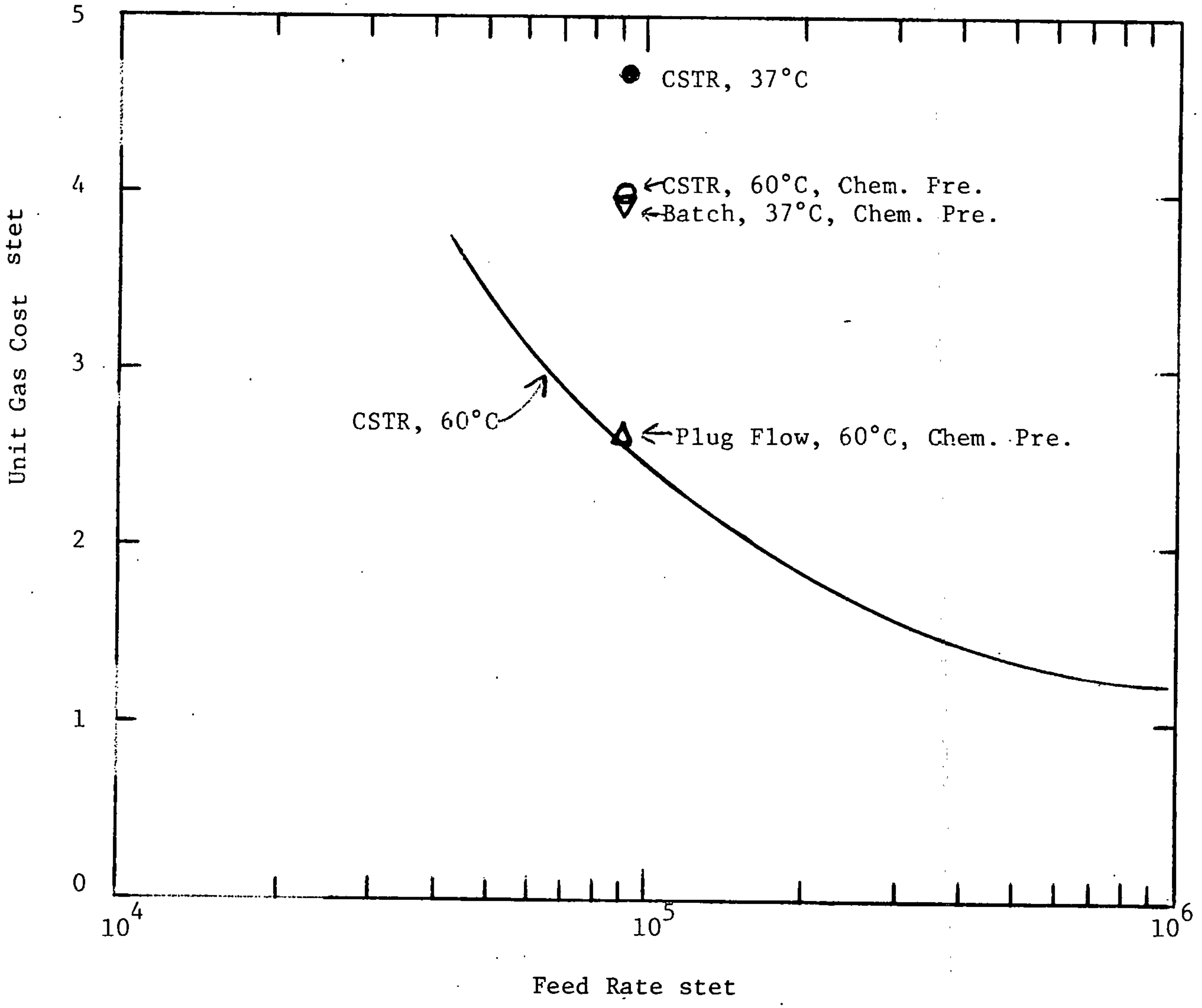

Unit Gas Cost stet
5
4
3
2
1
0
CSTR, 37°C
CSTR, 60°C, Chem. Pre.
Batch, 37°C, Chem. Pre.
CSTR, 60°C
Plug Flow, 60°C, Chem. Pre.
10^4
10^5
10^6
Feed Rate stet

Figure 34

Unit Gas Cost for Digestion of Bagasse

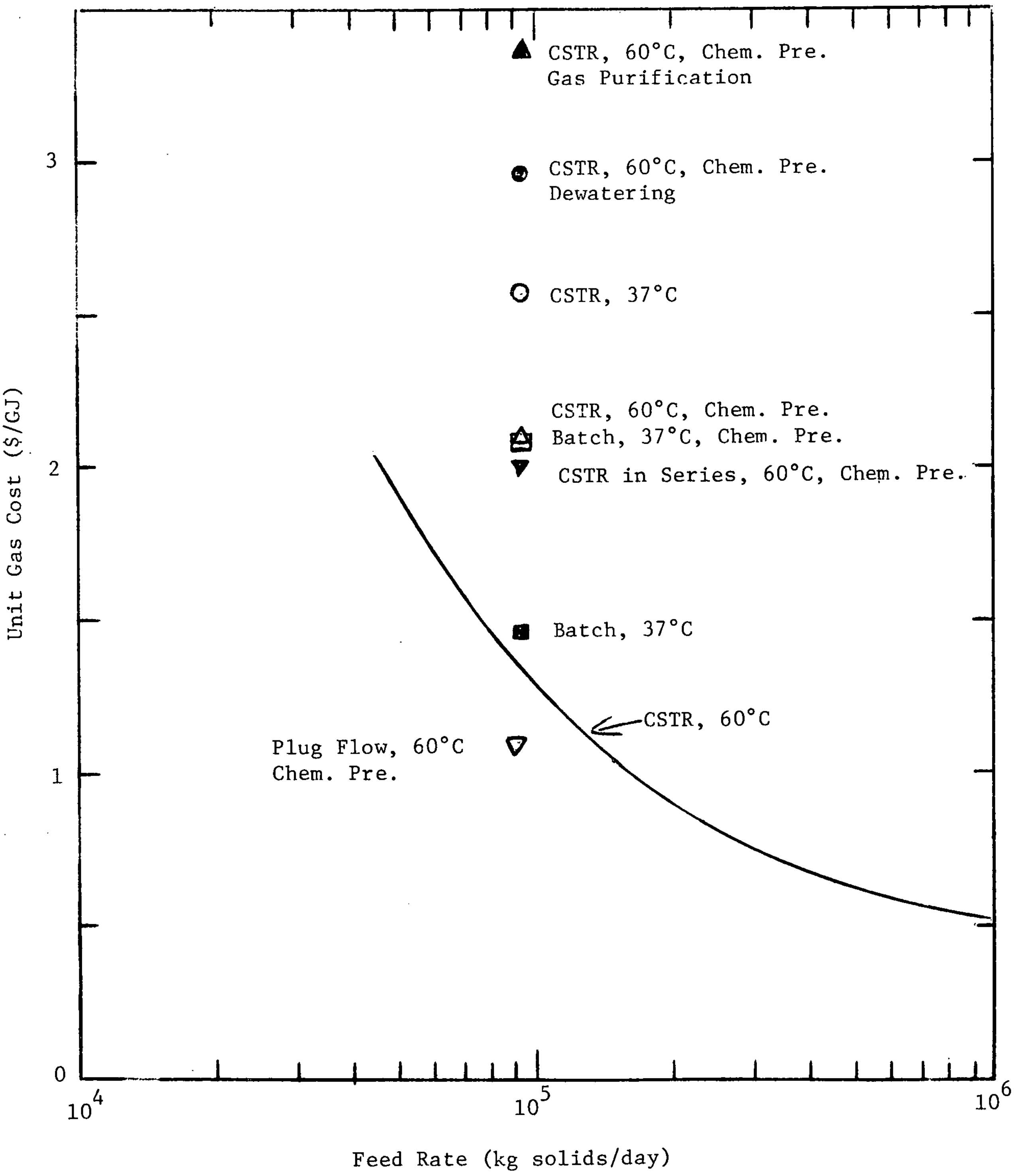


Table 5.1

RELATIONSHIP BETWEEN RETENTION TIME AND FRACTIONAL CONVERSION OF BIOMASS

CSTR	$\tau = \left(\frac{1}{1 - x} - 1\right) / k$
CSTR in Series	$\tau = \left[\left(\frac{1}{1 - x}\right)^{\frac{1}{N}} - 1\right] N / k$
Plug Flow and Batch	$\tau = \left[\ln\left(\frac{1}{1 - x}\right)\right] / k$

τ = retention time

k = first order kinetic rate constant

x = fractional conversion of biomass

N = number of CSTR digesters in series

Another major factor affecting unit gas cost is digester cost. For the 90800 kg/day (100 ton/day) feed rate, the cost for the CSTR digester was approximately $\$170/m^3$ ($\$130/yd^3$) whereas for the plug flow system, the cost was $\$13/m^3$ ($\$10/yd^3$). The CSTR system is a standard high cost reactor while the plug flow system is a low cost "hole-in-the-ground" reactor. The result of these differences in digester volume and cost is seen in Tables F.8 and F.12, where the CSTR digester cost is $274,000 and the plug flow digester cost is only $16,000. The resultant unit gas costs are $3.04/GJ ($3.21/MM Btu) and $1.41/GJ ($.149/MM Btu) for the CSTR and plug flow systems, respectively. The difference in unit gas costs is due primarily to the difference in digester costs. These effects are also noted for the other types of biomass feeds. This emphasizes the economic advantage of the low capital system over the standard tank digester.

For dirt feedlot manure feedstock the unit gas cost for a batch land-fill type digester ($4.26/GJ) is higher than the unit gas cost resulting from a CSTR digester ($3.04/GJ). This cost is due primarily to the increased digester cost for a batch system. Even though the per volume digester cost for a batch system is significantly lower than for the CSTR system, $\$5.22/m^3$ ($\$4/yd^3$) vs $\$170/m^3$ ($\$130/yd^3$), the digester volume required is much greater, 165,000 m^3 (5.8×10^6 ft^3) vs 1600 m^3 (56,000 ft^3), resulting in a greater digester cost for the batch system compared to the CSTR, $865,000 vs $274,000. The greater volume for the batch system is required to contain a one year loading of feedstock (assuming a one year cycle is required for filling digesting, and emptying). This greater digester cost has more of an effect on increasing the unit gas cost than the increased gas production has on decreasing the unit gas cost. It should also be noted that for the batch system, any pretreatment equipment will be utilized only on a part-time basis during the loading step. For both rice straw and bagasse feedstocks, and with utilization of chemical pretreatment, the batch system is about equivalent to the CSTR system on a unit gas cost basis. In these cases, the increased digester cost is balanced by the increased production to result in the same unit gas cost.

The CSTR in series system is slightly more economical than the straight CSTR digester. This concept will approach a plug flow concept with respect to gas production so that the total digester volume will be less than for a CSTR system. However, due to a cost scaling exponent of 0.75, the cost/m^3 will be higher for the CSTR in series concept since this approach uses four smaller digesters to provide the necessary volume.

5.2 Effect of Batch Digester Cost

The unit gas cost for the batch digester concept was shown to be about the same as for the CSTR concept with rice straw and bagasse feedstocks. It was assumed for these cases that the "landfill" type digester cost was \$5.23/m^3 (\$4/yd^3). If it were possible to develop a land-fill digester system for a lower per volume cost, the unit gas cost would be correspondingly reduced. The sensitivity of unit gas cost to batch digester per volume cost is shown in Figure 35. The limit of \$0./m^3 yields a \$2.80/GJ unit gas cost. This limiting case can be considered as the expected unit gas cost resulting when no site excavation or preparation is necessary.

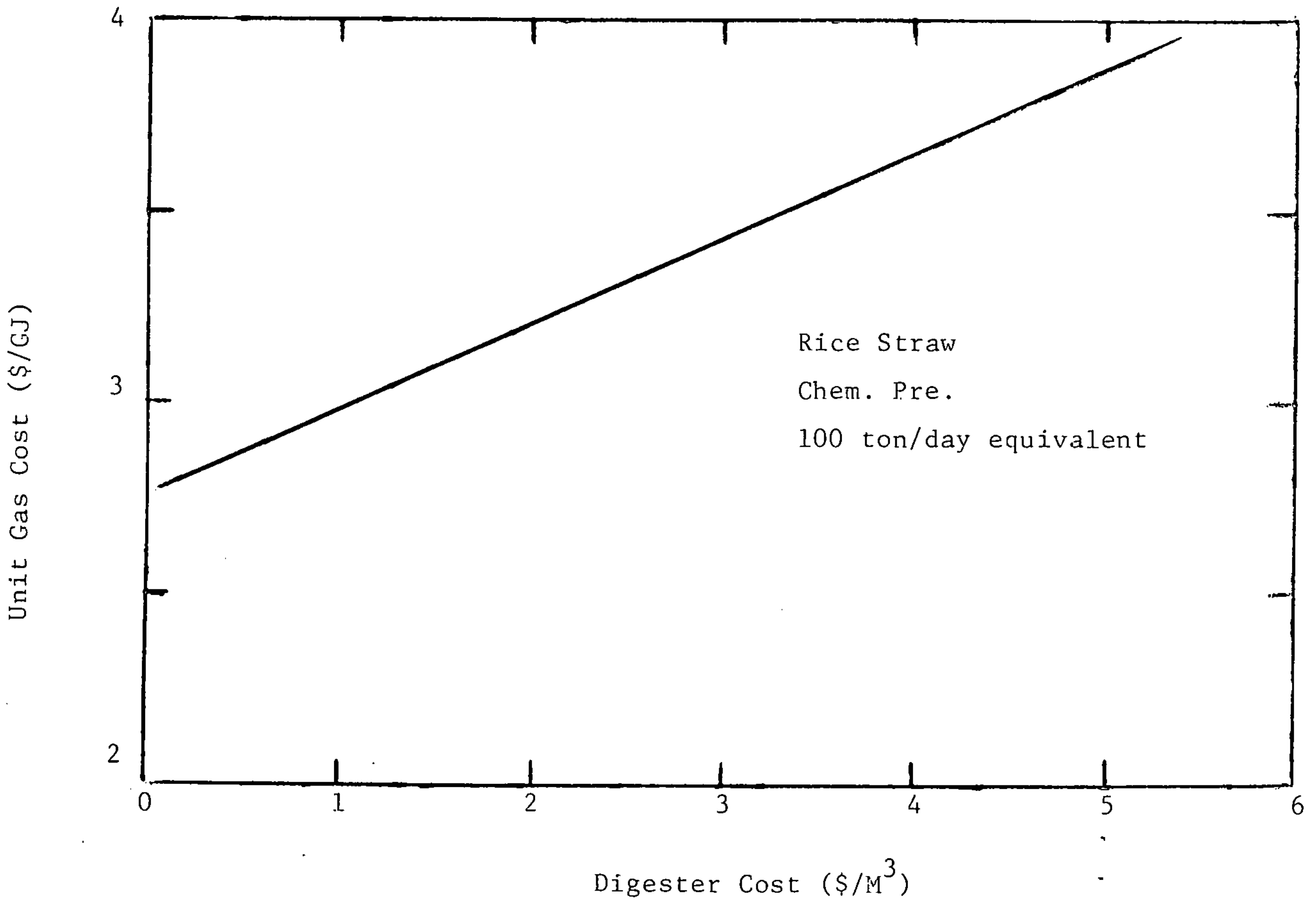
5.3 Effect of Chemical Pretreatment

Chemical pretreatment is utilized to improve the digestibility of the feedstock. The cost of chemical pretreatment must be compared to the additional productivity to determine economic feasibility. As can be seen from Figures 33 and 34, the use of chemical pretreatment results in over a 50% increase in unit gas cost. This increase is due to changes in both operating and capital costs.

For rice straw as a feedstock a comparison of Tables F.16 and F.19 indicates that the contribution to unit gas cost due to capital costs decreases with use of chemical pretreatment. This can be explained by the 90% increase in gas production, so that even though the total capital costs increase, the capital contribution to unit gas cost decreases from about \$1.07/GJ to \$0.86/GJ or about 20%. On the other hand, since such a large fraction of the feedstock is converted to biodegradable solids, a large quantity of pretreatment

Figure 35

Effect of Batch Digester Cost on Unit Gas Cost



chemical will be required. This results in significantly greater contributions of operating costs to the unit gas cost, $3.11/GJ with chemical pretreatment compared to $1.49/GJ without, over a 100% increase. The net result is a 55% increase in unit gas cost.

A comparison of Tables F.24 and F.27 for bagasse indicates an increase in the capital contribution to unit gas cost from $0.56/GJ to $0.63/GJ (about 12% greater). The operating cost increase is from $0.77/GJ to $1.48/GJ, or over 90% increase. The net result is about a 58% increase in unit gas cost.

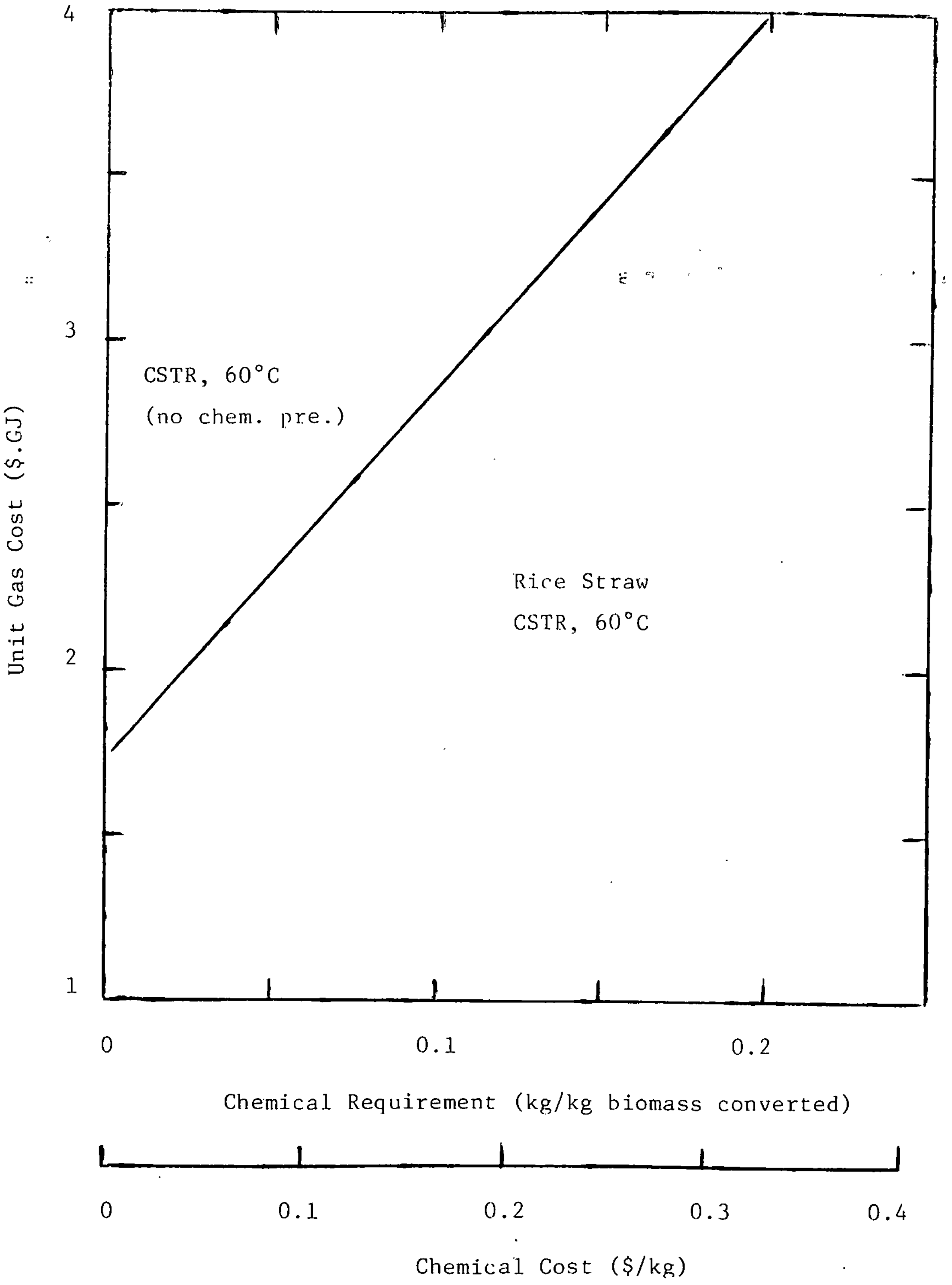
Since the major contribution to increased unit gas cost associated with chemical pretreatment is due to the cost of pretreatment chemical, the sensitivity of unit cost to chemical cost was analyzed. The result of this sensitivity analysis is presented in Figure 36 and Tables F.37, F.38, and F.39 for rice straw. The baseline conditions for use of chemical pretreatment are $0.33/kg ($300/ton) for chemical cost and 0.2 kg chemical/kg biomass converted to biodegradable material. The unit gas cost for the base-line case is $3.98/GJ. As the chemical requirement is decreased, the unit gas cost shows a corresponding decrease, as indicated in Figure 36. The effect of decreased unit chemical cost is similar and is also shown in Figure 36. It should be noted that if either the chemical requirement is decreased below 0.07 kg per kg converted or the cost is decreased below $0.12/kg, the unit gas cost will be lower than the unit gas cost for no chemical pretreatment. This is a consequence of the increased gas production resulting from chemical pretreatment of rice straw which leads to a decrease in the capital contribution to the unit gas cost, as indicated above. These limits correspond to a 65% decrease in total chemical cost for pretreatment.

5.4 Effect of Other System Components

The use of other system components also will influence the unit gas cost, but not as significantly as chemical pretreatment. These other options should be considered only when it is necessary to ensure proper operation of the digestion process. For example, a degritter should be used for high

Figure 36

Effect of Pretreatment Chemical Requirement and Cost on Unit Gas Cost



ash containing feedstocks to eliminate any problem arising from ash build-up in the digester. Also, dewatering should be used only when necessary for handling of the effluent.

The use of a shredder was included for the systems with rice straw. It contributes approximately $0.16/GJ ($0.17/MM Btu) to the unit gas cost for a 90,800 kg/day (100 ton/day) feed rate. This increase in cost is due primarily to the capital cost of the shredder unit ($77,000) as can be seen from Table F.16, which presents the results for a CSTR digestion system for rice straw.

A degritter was included in the treatment of dirt feedlot manure since this feedstock has a high ash content which could lead to digester failure if not removed. The use of a degritter results in approximately no increase in unit gas cost, since the capital cost for the degritter is only about 0.5% of the system capital cost.

The effect of gas scrubbing was evaluated for a CSTR with bagasse feedstock and the results are presented in Table F.32. A comparison with Table F.37 indicates that this processing step will increase the unit gas cost about $1.26/GJ ($1.33/MM Btu). This is similar to the results presented by Ashare et al (Ref. 337). Since the cost of gas scrubbing is so high this option should only be used when necessary for delivery to a pipeline.

The use of a dewatering step to provide digester effluent solids in a more concentrated form and possibly to provide recycle water will result in an increased unit gas cost of about $0.80/GJ ($0.86/MM Btu) as can be seen by a comparison of Tables F.27 and F.31. This option should be used when a credit for the concentrated effluent compensates for the increased cost, or if it is necessary to satisfy environmental regulatory requirements. However, an evaluation of these by-product credits or requirements should be done on a site specific or case-by-case basis.

5.5 Other Digester Design Concepts

The results presented in Section 5.1 were for the simpler single phase digester design concepts. Other more complicated processes are possible as indicated in Table 1.1. These more complicated processes can be considered as either multi-digestion processes in series, such as for multi-stage digestion, or pretreatment followed by digestion, such as for biomethanation.

The economic analysis results for the CSTR in series system are generally applicable to the multi-stage digestion concepts which are designed to yield greater gas production from two (or more) smaller digesters. A resonable result from this technical analogy is that the unit gas cost for the multi-stage digestion process will be comparable to the CSTR in series unit gas cost and slightly lower than the unit gas cost for the conventional stirred tank system.

The biomethanation process can be considered as an oxidation pretreatment followed by a high pressure fermentation. The oxidation pretreatment step produces an intermediate Btu gas, and the cost for this process (based on the Purox process) is approximated to be \$6.12/GJ (\$6.45/MM Btu) for rice straw at 100 ton/day feed rate (Ref. 355), with about 75% of this cost due to capital costs and 25% due to operating costs. An order of magnitude increase in feed rate will decrease the unit gas cost by about 50%. The fermentation step will add to this cost by necessitating further capital expenditures and decreasing the net process conversion efficiency. The cost of the fermentation step will depend on the gas production per volume of reactor per day (VVD) for the process. Preliminary experimental results have indicated a 250 VVD is possible (Ref. 202). For this productivity, a 100 ton/day feed rate would require about a 5,000 ft^3 pressure fermenter. The cost of such a vessel would be approximately \$130,000 (Ref. 348). This would add approximately \$0.10/GJ to the unit gas cost, which does not include ancillary equipment or operating costs associated with the fermentation step. Obviously, higher or lower VVD would result in lower or higher unit gas cost, respectively. Based on these results, a reasonable cost for the biomethanation step would be greater than $7.00/GJ. It should be noted that this cost utilizes a high cost for the pretreatment or pyrolysis step obtained from the literature. The results presented here for biomethanation are based on an analysis performed by others and this must be considered when comparing the biomethanation process with any of the other digester design concepts analyzed in detail and presented as part of this report. It should also be noted that the cost analysis presented by Alich et al. (Ref. 355) is not in agreement with cost data presented by Boegley et al. (Ref. 349).

5.6 Digester Design Productivity

One method of technical comparison of digester processes is the productivity, or VVD, of the system. The productivity of the various processes discussed in Section 4.1 are presented in Table 5.2. These results were obtained from the computer analysis results given in Appendix F. The plug flow system is more productive than either the CSTR or batch process, and this, combined with the low-capital plug flow system, results in the lowest unit gas cost as discussed in Section 5.1.

Table 5.2

PRODUCTIVITY OF DIGESTER DESIGN CONCEPTS

	VVD*			
Feedstock	CSTR	Plug Flow	Batch	CSTR in Series
Dirt Feedlot Manure	3.49	5.53	0.15	-
Rice Straw	4.66	7.39	0.41	-
Bagasse	5.72	9.07	0.49	11.55

*VVD is volume of methane produced per volume of digester per day.

It should be noted that the productivities indicated in Table 5.2 are greater than those reported in the literature. This is due to the digestion conditions utilized in this study, namely high solids concentration and low retention time. These conditions were incorporated since they represent limits (to be attained) which could lead to economically feasible operation. Use of lower solids concentration and longer retention time will give lower VVD's and correspondingly higher unit gas costs.

Section 6

CONCLUSIONS AND RECOMMENDATIONS

The results of the engineering economic analysis performed on the various digester design concepts and processing options have led to the following conclusions and recommendations. These conclusions and recommendations are based on the engineering economic analyses performed as part of this study. For these analyses, it was sssumed that the processes were operating without interruptions. The actual utilization of some of these design concepts to produce methane from biomass should be preceded by preliminary experimental work to determine if the actual performance will be equivalent to the performance determined by these analyses.

6.1 Conclusions

1. The most economical (based on unit gas cost) of the systems analyzed is the plug flow design concept. This is a result of the high productivity (VVD) and low capital "hole-in-the-ground" design used for this concept.

2. The analysis for the batch "land-fill" type digester results in a unit gas cost which is comparable to the conventional stirred tank digester design. However, the cost for the batch system could be lower if the area for the landfill requires little or no preparation prior to loading the biomass.

3. The use of chemical pretreatment will result in a higher unit gas cost, primarily due to the cost of pretreatment chemical.

4. Chemical or thermal pretreatment could be potentially more economical, providing the increased gas production more than compensates for the cost of pretreatment equipment and operation.

5. Other system options such as shredding, degritting, dewatering, and gas purification all result in increased unit gas cost,

and should only be used when necessary for efficient system operation or when necessary to provide purified gas for a pipeline or digester effluent solids as feed or fertilizer.

6. The results of this analysis do not incorporate any credits or penalties for the feedstock or byproducts, but present the costs only for the conversion of biomass to methane. An analysis of credits and penalties should be performed on a case-by-case basis.

7. The unit gas cost for the multi-stage digester concept can be approximated by the CSTR in series system and should be slightly more economical than the conventional stirred tank digester concept.

8. The biomethanation process appears to be significantly less economical than the conventional stirred tank system, but this conclusion is based on costs for the Purox pyrolysis process which are questionable, i.e., two sources of cost data for the Purox process are not in agreement.

6.2 Recommendations

1. Since the low capital plug flow digester concept appears to be the most economical, it is recommended that experimental work be continued for this system for manure and other feedstocks to determine the most efficient operating procedures. A more detailed economic analysis of this concept should be performed using actual performance data, when available, and incorporating equipment costs from vendor quotes rather than cost estimations from the literature.

2. A more detailed economic analysis of chemical and thermal pretreatment should be performed to determine a break-even point for this option. This could then be used to direct the experimental effort in this area to lead to an economically and technically feasible pretreatment process.

3. A more detailed analysis of the biomethanation process is necessary to determine its economic feasibility. This is primarily directed toward the cost analysis for the pyrolysis (or pretreatment) step.

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APPENDIX A

Capital Costs

Capital costs for the equipment utilized for the various system components are estimated from equipment cost data found in the literature.

A.1 Shredder

Shredder costs were estimated as (Ref. 346):

$$\text{cost} = 10{,}600\ (\text{capacity})^{1.0}$$

where the capacity is in tons/hr. This can also be expressed as:

$$\text{cost} = 11.7\ (\text{capacity})^{1.0}$$

where the capacity is in kg/hr. These costs have a 1973 basis.

A.2 Degritter

The cost of a degritter is given by DiGregorio (Ref. 347) on a February, 1968 basis as:

$$\text{cost} = 3800\ Q^{0.35}$$

where Q is the flow rate in MMGD. For a density of 1026 kg/m^3 (64 lb/ft^3), the cost can also be given as:

$$\text{cost} = 57.1\ Q_m^{0.35}$$

where Q_m is the flow in kg/hr.

A.3 Chemical Pretreatment

There are several possible chemical pretreatment processses. One is alkali pretreatment at elevated temperatures, e.g., 200°C. The cost of a pressure vessel for this treatment is given by Guthrie (Ref. 348), on a 1970 basis as:

$$\text{cost} = 129\ V^{0.75}$$

where V is the capacity in cubic feet and the pressure rating is 300 psi. For SI units, the cost is:

$$\text{cost} = 1869\ V^{0.75}$$

where V is in m^3.

Another pretreatment approach is to use a pyrolysis process which is then combined with biomethanation. The cost of pyrolysis is estimated by Boegly _et al_.(Ref. 349) on a 1977 basis as $14 million for a 1000 ton/day facility. Assuming a 0.7 scale factor, this can be estimated by:

$$\text{cost} = 100{,}000\ (\text{capacity})^{0.7}$$

where the capacity is in tons/day. This cost is an approximation for the Purox pyrolysis system. The cost, utilizing SI units, is:

$$\text{cost} = 850\ (\text{capacity})^{0.7}$$

where the capacity is in kg/day.

A.4 Storage

The storage unit for the continuous processes is designed with a holding time of 2 days. The cost of this unit is given by Patterson _et al_. (Ref. 350) as:

$$\text{cost} = 13722\ V^{0.52}$$

where V is the capacity in MCF and the cost basis is 1971. The cost, expressed with SI units is:

$$\text{cost} = 2412\ V^{0.52}$$

where V is in m^3.

A.5 Digester

There are several general types of digesters to consider. One is a high capital standard stirred tank digester for which the cost is (Ref. 350):

$$\text{cost} = 45\ V^{0.75}$$

where V is the volume in ft^3 and the basis is 1971. The cost can also be given as:

$$\text{cost} = 651.8\ V^{0.75}$$

where V is the volume in m^3.

Another digester design is a low capital "hole-in-the-ground" system. The cost of such a system can be estimated from cost of excavation, which is taken as $\$2/yd^3$, plus the cost of liner and cover, taken as $\$1/ft^2$. For large systems (> 1 MM gallons), Kays (Ref. 351) estimates the cost to range from 2 - 5¢ per gallon. This is a range of $\$4/yd^3$ to $\$10/yd^3$. For the analyses presented here, the lower value will be utilized for large land-fill type systems and the higher value will be utilized for small plug-flow hole-in-the-ground digesters.

A third digester design is a pressure vessel for the biomethanation of pyrolysis gases. The cost of pressure vessels has been estimated from Guthrie (Ref. 348) as:

$$\text{cost} = 1300\ V^{0.75}$$

where V is volume in m^3 and the basis is 1970. This relationship is for a pressure rating of 1000 psi.

A.6 Heat Exchanger

The heat exchanger cost is given by Peters and Timmerhaus (Ref. 339), on a 1967 basis as:

$$\text{cost} = 26\ A^{0.75}$$

where A is the area in ft^2. This also can be given as:

$$\text{cost} = 154.5\ A^{0.75}$$

where A is the area in m^2.

A.7 Dewatering

The cost of dewatering is given by (Ref. 350):

$$\text{cost} = 36{,}000\ Q^{0.5}$$

where Q is the flow rate in GPM. For a stream density of 1026 kg/m^3 (64 lb/ft^3), the cost of dewatering is:

$$\text{cost} = 2356\ F^{0.5}$$

where F is the flow rate in kg/hr.

A.8 Gas Purification

The cost for gas purification is estimated from Ashare et al. (Ref. 337) as:

$$\text{cost} = 100\ (G)^{0.6}$$

where G is the total gas flow (on a dry basis) in ft^3/day and the cost is on a 1977 basis. The cost is also given by:

$$\text{cost} = 848.7\ G^{0.6}$$

with G expressed as m^3/day.

APPENDIX B

Power Requirements

B.1 Shredder

The power requirement for shredding is estimated as 0.0165 hp/kg/hr (15 hp/ton/hr) throughput.

B.2 Degritter

The power requirement per degritter is estimated as 2 hp.

B.3 Chemical Pretreatment

The power for chemical pretreatment will depend on whether the process requires mixing. If so the mixing requirement will be estimated at 0.00706 hp/m^3 (0.2 hp/1000 ft^3).

B.4 Storage

Power will be necessary for storage only when mixing will be required. This power requirement is estimated at 0.00706 hp/m^3 (0.2 hp/MCF).

B.5 Digestion

The power requirement for digestion will also be 0.00706 hp/m^3 (0.2 hp/MCF) when mixing is utilized.

B.6 Dewatering

The power requirement for dewatering is estimated as 3.47 x 10^{-3} hp/kg/hr (3.15 hp/ton/hr).

B.7 Gas Purification

The power requirement for gas purification is not explicitly included in the economic analysis as a utility cost. However, it is implicitly included in the operating cost provided for the gas purification system.

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APPENDIX C

Material Balance

Material balances can be made around each piece of equipment for the over-all process. These balances will be for solids (total, volatile, biodegradable volatile, and ash), water, gas, and total.

C.1 Shredder

The input and output streams associated with the shredder are assumed to be identical. The only change is in the size of the particles.

C.2 Degritter

The degritter is used to remove ash. It is assumed that no volatile solids are removed in the degritter. Figure C.1 is a block diagram for the degritter indicating the material balance. In this diagram, TSF and TS1 are the input and effluent total solids streams respectively, and TAF and TA1 are the ash contents of these streams. The amount of ash removed, TAR, is given by:

$$TAR = EDG \times TAF$$

where EDG is the efficiency of ash removal.

C.3 Chemical Pretreatment

The chemical pretreatment material balance is given in Figure C.2. The amount of non-biodegradable volatile solids converted to biodegradable volatile solids is:

$$TCBVS = ECHP\ (TVSF - TBVSF)$$

where ECHP is the conversion efficiency. The amount of chemical added is assumed to be directly proportional to the amount converted.

$$TCHEM = CHEM\ (TVSF - TBVSF) \times ECHP$$

where CHEM is the amount of chemical required per amount converted.

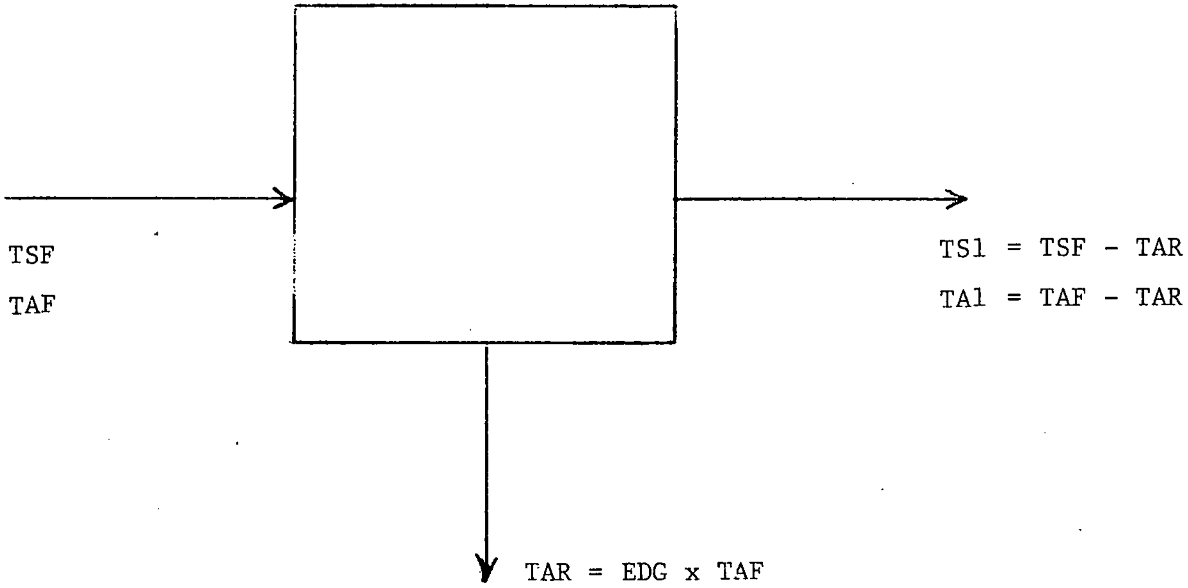

Figure C.1 Degritter Material Balance

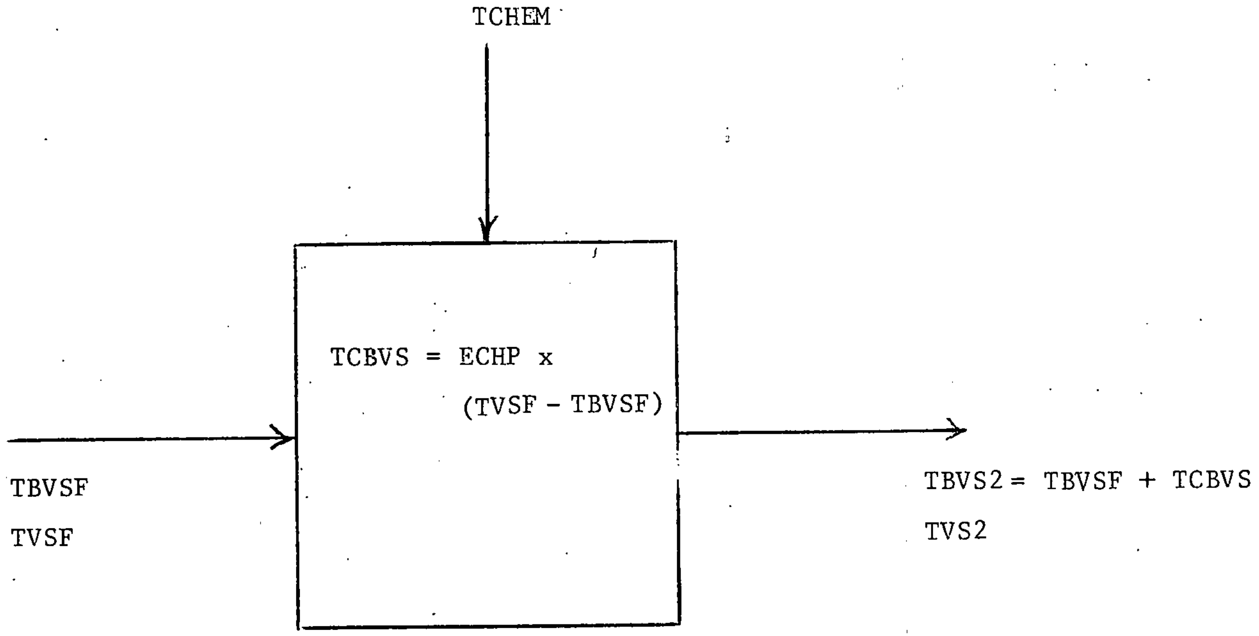

Figure C.2 Chemical Pretreatment Material Balance

C.4 Storage

The storage (and mixing) section is used to prepare the feed material for the digester. This includes mixing (when necessary) and addition of water to obtain the desired concentration. The material balance is indicated in Figure C.3. The solids concentration entering the digester, FSDF, is given by:

$$FSDF = TS3/TF3$$

or

$$FSDF = TS2/(TF2 + W)$$

since only water is added at this step and the total solids remain constant.

C.5 Digester

The material balance for the digester is indicated in Figure C.4. The amount of gas produced is given by:

$$GPD = A(TBVS2)\ (FBVSC)\ (DNS)\ /\ (TF3)\ (HRTDG)$$

where GPC is the gas production (VVD), FBVSC is the fraction of biodegradable volatile solids converted, DNS is the stream density and HRTDG is the retention time. The gas is assumed to consist of 60% CH_4/40% CO_2.

C.6 Dewatering

The dewatering unit splits the digester effluent into two streams, one a high solids content stream and the other a low solids content. The efficiency of solids removal is EDWT and the high solids content is FSDWT (fraction of solids). Hence,

$$TS5 = TS4 \times EDWT$$

and

$$FSDWT = TS5/(TS5 + TW5)$$

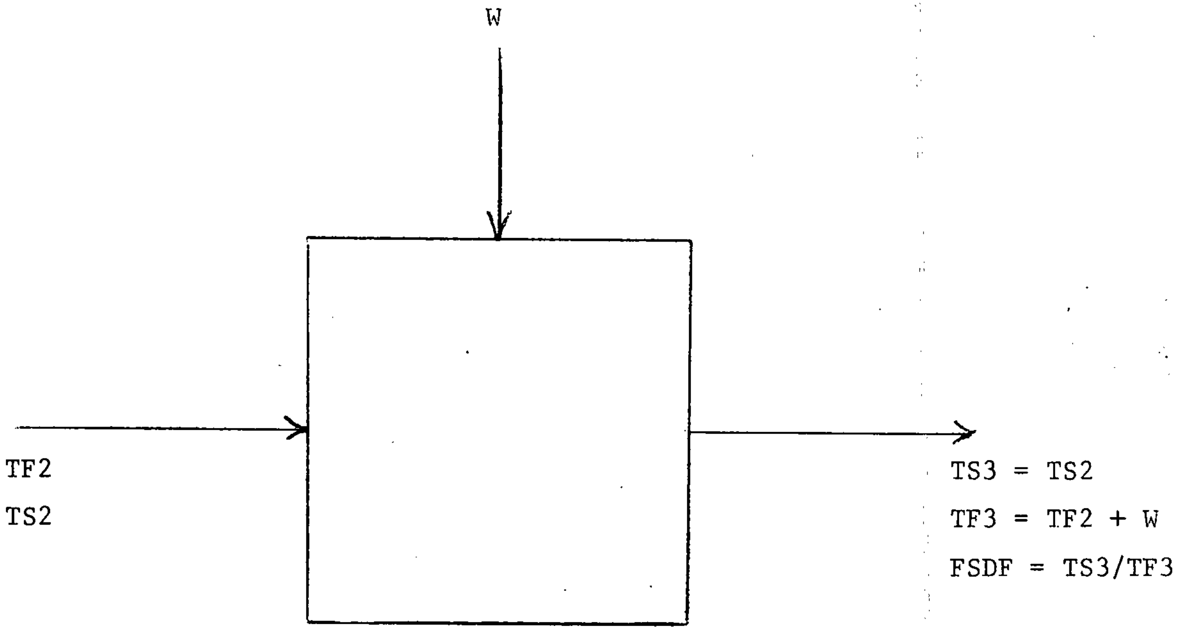

Figure C.3 Material Balance for Storage Section

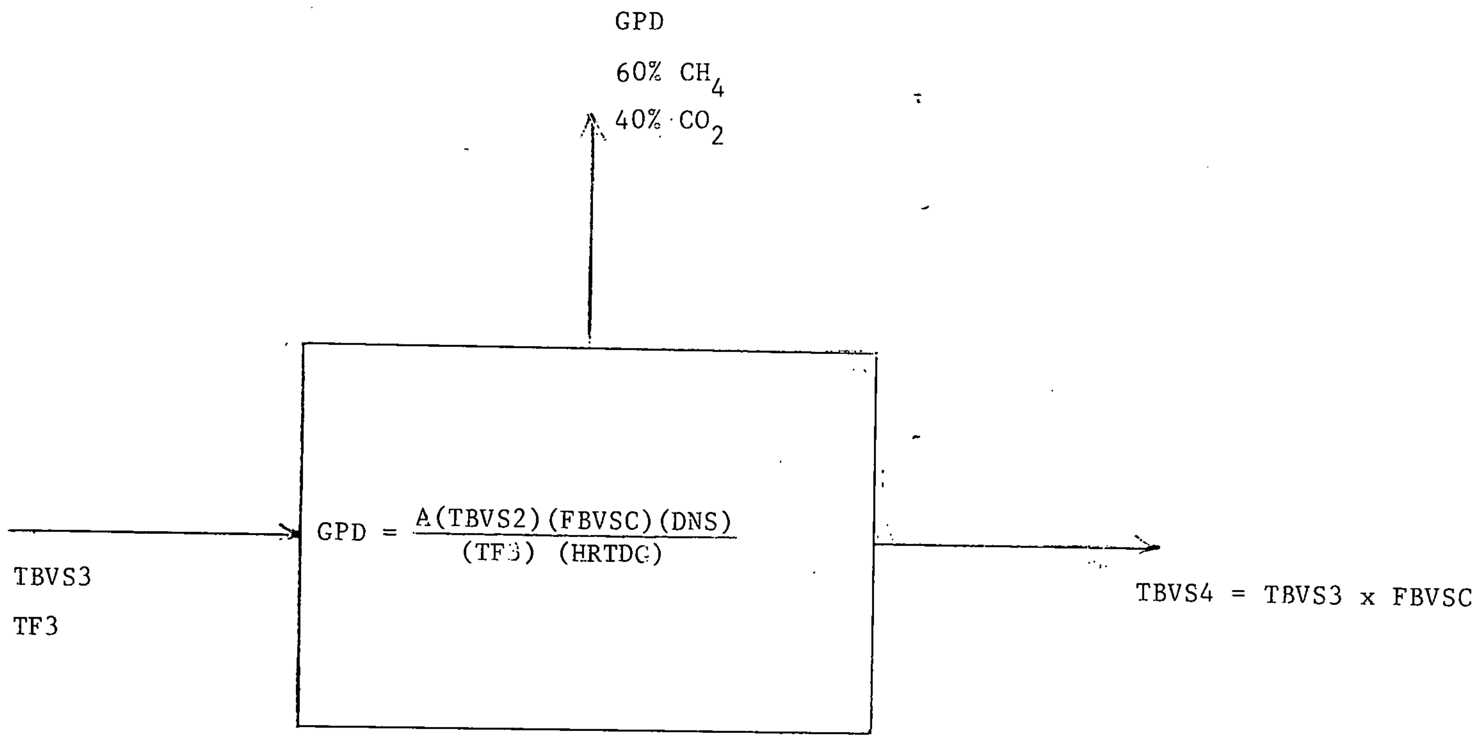

Figure C.4 Digester Material Balance

or

$$TW5 = TS5(1 - FSDWT) / FSDWT$$

The low solids stream is as indicated in Figure C.5.

C.7 Gas Purification

The digester gas consists primarily of CH_4 (∿ 60%) and CO_2 (∿ 40%). If it is necessary to provide a high energy content gas, 3.7×10^7 J/m^3 (1000 Btu/ft^3). a gas scrubbing process should be used. The scrubber will remove most of the CO_2 and a small fraction of the methane. The material balance utilized in this analysis assumes no methane is removed in the gas purification process.

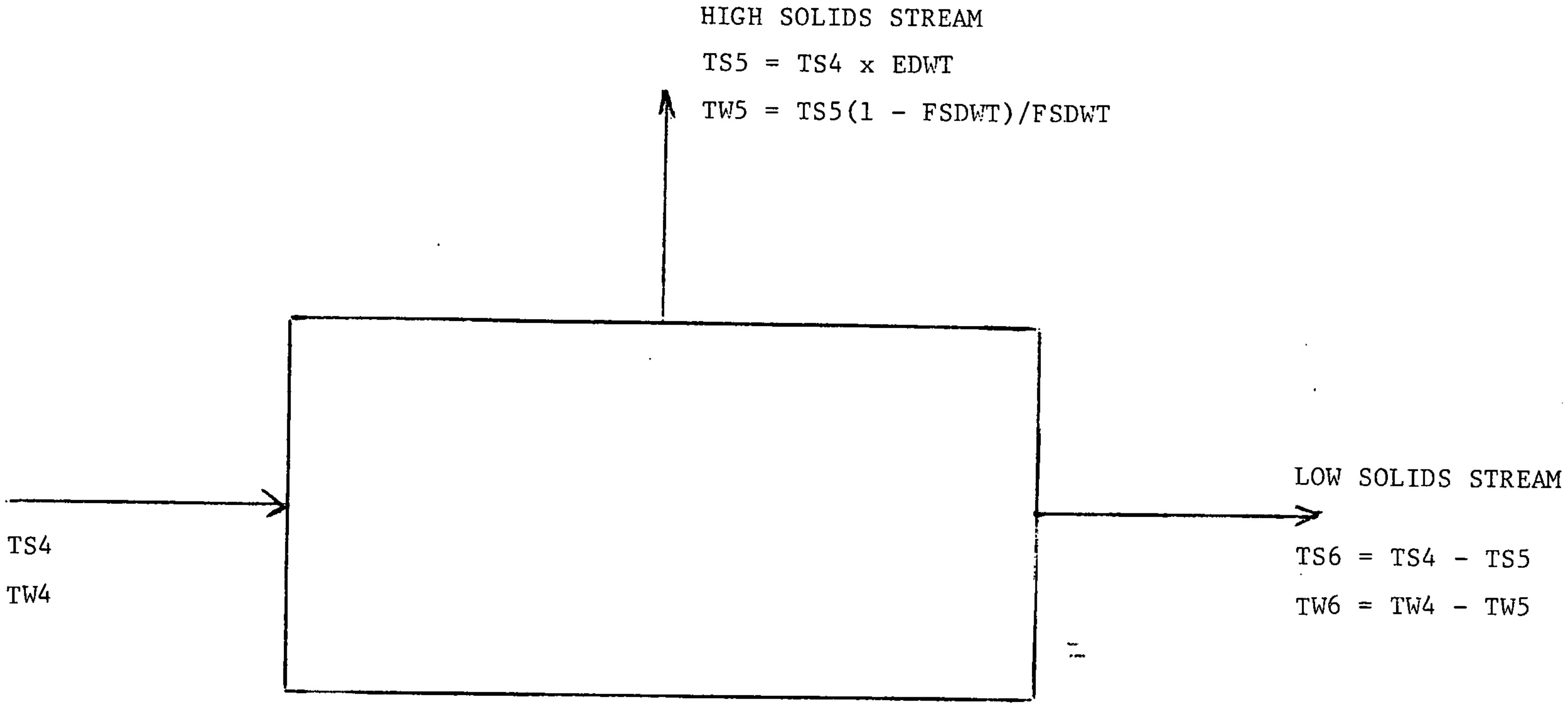

Figure C.5 Dewatering Material Balance

APPENDIX D

Heat Balance

The heat requirements for the digester are given by:

HREQ = HGAS + HEVAP + HEFF + HLOSS - HFEED - HW - HRXN

where

HREQ	is the heat requirement
HGAS	is the sensible heat of the gas stream
HEVAP	is the sensible heat and heat of evaporation of the moisture in the gas stream
HEFF	is the sensible heat of the digester effluent
HLOSS	is the heat loss through the digester walls
HFEED	is the sensible heat of the input stream
HW	is the sensible heat of the water make-up stream
HRXN	is the heat of reaction

For this analysis, the ambient temperature is TA1, the input streams temperature is TPF, and the digester temperature is TC. (For these equations, all temperatures are in °C.)

D.1 Input Stream

The sensible heat of the input stream is:

HFEED = 4184(kg/d) (TPF - TA) J/day

The heat capacity is assumed to be 1 Btu/lb°F.

D.2 Water Make-Up

HW = 4184 (kg/d) (TPF - TA) J/D

D.3 Effluent Stream

The sensible heat of the digester effluent stream is:

$$\mathrm{HEFF} = \frac{kg}{d} \ (\mathrm{TPEX} - \mathrm{TA}) \ 4184 \ \mathrm{J/d}$$

where TPEX is the exit temperature (either from the digester or heat exchanger) and the heat capacity is assumed to be 1 Btu/lb°F.

D.4 Digester Heat Loss

The heat loss from the digester can be expressed as:

$$\mathrm{HLOSS} = U_S A_S \ (T - TS) + U_B A_B \ (T - TS) + U_T A_T \ (T - TS)$$

where the terms on the right hand side represent the loss from the walls, bottom and top respectively. In each case U is the over-all heat transfer coefficient and A is the surface area. For a concrete tank with dirt insulation, approximate over-all heat transfer coefficients are (Ref. 354):

$$U_S = 0.5 \ \mathrm{Btu/hrft^2 {}^\circ F}$$
$$U_B = 0.08$$
$$U_T = 0.37$$

If the digester is assumed to be cylindrical with equal diameter and height, the digester volume is:

$$V = \pi D^{3/4}$$

and the surface areas are:

$$A_T = A_B = 0.923 \ V^{2/3}$$

$$A_S = 3.69 \ V^{2/3}$$

The heat loss is then found to be approximated by:

$$\mathrm{HLOSS} = 1.097 \times 10^6 \ V^{2/3} \ (\mathrm{TC} - \mathrm{TA}) \ \mathrm{J/d}$$

where V is the digester volume (m^3) and the temperatures are in °C.

D.5 Gas Stream

The sensible heat loss with the gas stream is made up of contributions from methane, carbon dioxide, and water. The water is included in HEVAP (D.6).

$$\text{Methane Contribution} = DGP \times \frac{16}{23.7} \times 0.56 \times 0.00397 \times (TC - TA) \times 1000 \times 1055$$

$$\frac{m^3\,CH_4}{day} \quad \frac{gm\ CH_4}{\ell} \quad \frac{cal}{g^\circ C} \quad \frac{Btu}{cal} \quad {}^\circ C \quad \frac{\ell}{m^3} \quad \frac{J}{Btu}$$

[Note that a molar volume of 23.7 was used since the gas industry takes STP at 60°F and the gas production was calculated on that basis. The heat capacity of methane is taken as 0.56 cal/g°C (Ref. 352).

Hence, the methane contribution is:

$$1581\ DGP\ (TC - TA)$$

The CO_2 contribution is found to be (assuming the $CH_4:CO_2$ ratio of 6:4) with $C_{p,\,CO_2} = 0.215$ cal/g°C (Ref. 352):

$$1113\ DGP\ (TC - TA)$$

Thus,

$$HGAS = 2694\ DGP\ (TC - TA)\ J/d$$

D.6 Evaporated Water

The heat loss due to evaporated water is due to the sensible heat loss of the stream and heat of evaporation of water. The amount of water evaporated is dependent on temperature and is calculated assuming the gas stream is saturated with water at digester temperature.

$$HEVAP = H2OEV\ [\Delta H_{VAP} + C_p (TC - TA)]$$

$$HEVAP = H2OEV\ [2.382 \times 10^6 + 1.87 \times 10^3\ (TC - TA)]\ J/day$$

where H2OEV is the amount of water evaporated (kg/day).

The amount of water leaving the digester by evaporation is:

$$H2OEV = 1.266\ (DGP)\ (XX)\ /\ (1 - XX)\ \text{kg/d}$$

where XX is the mole fraction of water (assuming saturation) given by:

$$XX = 1.27 \times 10^{6} \exp\left[-5220/(TC + 273.16)\right]$$

D.7 Heat of Reaction

For the digestion of biomass to methane the reaction is exothermic. For:

$$\text{cellulose} + H_2O = 3CH_4 + 3CO_2$$

the heat of reaction is:

$$\Delta H_R = 3(\Delta H_{F,\ CH_4} + \Delta H_{F,\ CO_2}) - \Delta H_{F,\ H_2O} - \Delta H_{F,\ \text{cellulose}}$$

where the ΔH_F's are the heats of formation. These are (Ref. 352):

$$\Delta H_{F,\ CH_4} = 20.3\ \text{kcal/mol}$$

$$\Delta H_{F,\ CO_2} = 94.4$$

$$\Delta H_{F,\ H_2O} = 68.4$$

The heat of formation of cellulose can be obtained from the heat of combustion, which is 671 kcal/mole (Ref. 353):

$$\text{cellulose} + 6O_2 = 6CO_2 + 5H_2O$$

$$\Delta H_{comb} = 6\Delta H_{F,\ CO_2} + 5\Delta H_{F,\ H_2O} - \Delta H_{F,\ \text{cellulose}}$$

from which $\Delta H_{F,\ \text{cellulose}}$ is 237.4 kcal/mol. The heat of reaction is 38.3 kcal per mole reacted or approximately 13 kcal per mole of methane produced. The heat produced in the reaction in the digester is:

$$HRXN = 2.296 \times 10^{6}\ DGP\ \text{J/d}$$

APPENDIX E

Computer Program
and
List of Symbols

```
      PROGRAM B10GAS4(INPUT,OUTPUT,TAPE5=INPUT,TAPE6=OUTPUT)

      DIMENSION ACP(10),ACM(10),TNMBR(10),WK(10),CONS(10),EX(10),OIND(10
     1),CCOS(10),TCOS(10)
      READ(5,96)K
      READ(5,96)KK
      READ(5,95)(CONS(I),I=1,KK)
      READ(5,95)(EX(I),I=1,KK)
      READ(5,95)(OIND(I),I=1,KK)
      READ(5,95)CINDX,PFMP,PSEP,PSF,PCOP,PED,FPC,PIDC,PWC,STF,EFFS,AW,
     1PM,PSV,PAH,POPS,PLTI,PSUP,DIR,RE,FDBT,FITR,DPPC,CEP,CST
      READ(5,95)CF,TPF,TGRAD,HTC,EA,TA
      READ(5,95)SHRMC,SHRWK,EDG,DGPMC,DGRWK,CHEM,ECHP,RNUM,CPRMC,
     1CPRWK,STRMC,STRWK,HEXMC,DWTMC,DWTWK,EDWT,FSDWT
   95 FORMAT(F20.0)
   96 FORMAT(I3)
      DO 99 JJ=1,K
      READ(5,95)CPF,CCHM,TMP,WD,CDWV,CEFV,RKO,CONS(5),EX(5),OIND(5),DNS,
     1HRTCP,FSF,FVSF,FBVSF,A,FSDF,HRTST,DIGMC,DIGWK
      READ(5,96)INSHR,INDGR,INCPR,INSTR,INHEX,INDIG,INDWT,INGPU

C     FEED MAKEUP

      DO 199 IJK=1,4
      IF(IJK-2)198,197,196
  198 FD=50.
      GO TO 191
  197  FD=100.
      GO TO 191
  196 IF(IJK-4)195,194,193
  195 FD=500.
      GO TO 191
  194 FD=1000.
      GO TO 191
  193 CONTINUE
  191 CONTINUE
      TNDO=365.*STF
      TSF=FD*908.
      IF(INDIG-1)180,180,181
  181 TSF=FD*908.*365.*STF/90.
      TNDO=90.
  180 CONTINUE
      TFF=TSF/FSF
      TVSF=TSF*FVSF
      TBVSF=TSF*FBVSF
      TAF=TSF-TVSF
      TWF=TFF-TSF

C     SHREDDING

      IF(INSHR)11,11,12
   12 NSHR=TFF/(SHRMC*2.1792E4)
      TNMBR(1)=1.+NSHR
      ACP(1)=TFF/(TNMBR(1)*24.)
      WK(1)=TNMBR(1)*SHRWK*ACP(1)/908.
      GO TO 15
   11 ACP(1)=0.
```

```
      WK(1)=0.
      TNMBR(1)=0.

C     DEGRITTER

   15 IF(INDGR)21,21,22
   22 TAR=EDG*TAF
      TF1=TFF-TAR
      TS1=TSF-TAR
      TA1=TAF-TAR
      NDGR=TFF/(DGRMC*2.179E4)
      TNMBR(2)=1.+NDGR
      ACP(2)=TFF/(TNMBR(2)*24.)
      WK(2)=TNMBR(2)*DGRWK
      GO TO 25
   21 TF1=TFF
      TS1=TSF
      TA1=TAF
      ACP(2)=0.
      WK(2)=0.
      TNMBR(2)=0.

C     CHEMICAL PRETREATMENT

   25 IF(INCPR)31,31,32
   32 TCHEM=CHEM*(TVSF-TBVSF)*ECHP
      TCBVS=ECHP*(TVSF-TBVSF)
      TBVS2=TBVSF+TCBVS
      TF2=TF1+TCHEM
      TS2=TS1+TCHEM
      TA2=TA1+TCHEM
      TVS2=TVSF
      NCPR=TF1*HRTCP/(CPRMC*DNS*0.454)/24.
      TNMBR(3)=1.+NCPR
      ACP(3)=TF1*HRTCP/(TNMBR(3)*DNS*16.031)/24.
      WK(3)=TNMBR(3)*CPRWK*ACP(3)*35.31
      GO TO 35
   31 TF2=TF1
      TS2=TS1
      TA2=TA1
      TNMBR(3)=0.
      ACP(3)=0.
      WK(3)=0.
      TCHEM=0.
      TBVS2=TBVSF
      TVS2=TVSF

C     STORAGE

   35 IF(INSTR)41,42,42
   42 TF3=TS2/FSDF
      W=TF3-TF2
      IF(W)48,48,49
   48 TF3=TF2
      W=0.
      FSDF=TS2/TF2
   49 NSTR=TF3*HRTST/(DNS*STRMC*0.454)
```

```
      TNMBR(4)=1.+NSTR
      ACP(4)=TF3*HRTST/(TNMBR(4)*DNS*16.031)
      IF(INSTR)45,45,46
   46 WK(4)=TNMBR(4)*STRWK*ACP(4)*35.31
      GO TO 43
   45 WK(4)=0.
      GO TO 43
   41 TF3=TF2
      FSDF=TS2/TF2
      WK(4)=0.
      ACP(4)=0.
      TNMBR(4)=0.
      W=0.
   43 CONTINUE

C     DIGESTION

C     DIGESTER HEAT EXCHANGER

      TC=37.
      DO 90 JTMP=1,2
      IF(INDIG-1)39,39,38
   38 IF(JTMP-1)37,37,90
   37 CONTINUE
   39 CONTINUE
      IF(INHEX)51,51,52
   52 TAHEX=TF3*CF*(TC-TPF-TGRAD)/(HTC*TGRAD*117.24)
      NHEX=TAHEX/HEXMC*10.76
      TNMBR(6)=1.+NHEX
      ACP(6)=TAHEX/TNMBR(6)
      TPEX=TPF+TGRAD
      GO TO 55
   51 TPEX=TC
      ACP(6)=0.
   55 WK(6)=0.

      FBVSC=0.35
      DO 98 IJ=1,4
      FBVSC=FBVSC+0.15
      IF(INDIG-1)53,53,54
   54 FBVSC=0.95
      IF(IJ-1)53,53,98
   53 CONTINUE
      CBI=TBVS2*DNS/TF3
      CBE=CBI*(1-FBVSC)

C     RATE CONSTANT

      RK=RK0*EXP(-EA/(1.987*(TC+273.16)))

C     DIGESTER
      IF(INDIG)61,63,62

C     CSTR IN PARALLEL

   61 HRTDG=(CBI/CBE-1.)/RK
      TVD=TF3*HRTDG/(DNS*0.454)
```

```
      NDIG=TVD/DIGMC
      TNMBR(5)=1.+NDIG
      ACP(5)=TVD/(TNMBR(5)*35.31)
      GO TO 70

C     PLUG FLOW

   62 HRTDG=ALOG(CBI/CBE)/RK
      IF(INDIG-1)65,65,66
   65 TVD=TF3*HRTDG/(DNS*0.454)
      GO TO 67

C     BATCH

   66 TVD=TSF*90./FSDF/DNS/0.454
   67 NDIG=TVD/DIGMC
      TNMBR(5)=1.+NDIG
      ACP(5)=TVD/(TNMBR(5)*35.31)
      GO TO 70

C     CSTR IN SERIES

   63 HRTDG=((CBI/CBE)**(1./RNUM)-1.)*RNUM/RK
      TVD=TF3*HRTDG/(DNS*0.454)
      TNMBR(5)=RNUM
      ACP(5)=TVD/(TNMBR(5)*35.31)
   70 WK(5)=TNMBR(5)*DIGWK*ACP(5)*35.31

C     GAS PRODUCTION

      IF(INDIG-1)401,401,402
  401 GPD=A*(CBI-CBE)/HRTDG
      DGP=GPD*TVD/35.31
      GO TO 405
  402 DGP=TRVS2*FBVSC*A/16.03
      GPD=DGP/TVD*35.31
  405 CONTINUE

C     DIGESTER HEAT REQUIREMENT

      HFEED=TF3*CF*(TPF-TA)*4184.
      HGAS=DGP*(TC-TA)*2694.
      XX=1.27E6*EXP(-5220./(TC+273.16))
      H2OEV=TVD*GPD*XX/(1-XX)*0.0359
      HEVAP=H2OEV*(2.382E6+1.87E3*(TC-TA))
      TF4=TF3-(H2OEV+(CBI-CBE)/DNS*TF3)
      HEFF=TF4*CF*(TPFX-TA)*4184.
      HRXN=DGP*2.296E6
      HLOSS=1.0972E6*(ACP(5)**(2./3.))*TNMBR(5)*(TC-TA)
      HW=W*(TPF-TA)*4184.
      IF(INDIG-1)86,86,87
   86 HREQ=HGAS+HEVAP-HRXN+HEFF+HLOSS-HFEED-HW
      GO TO 85
   87 HREQ=HGAS+HEVAP-HRXN
   85 CONTINUE
      IF(HREQ)77,77,78
   77 HREQ=0.
```

```
   78 TBVS4=C9E/DNS*TF4
      TVS4=TVS2-(TBVS2-TBVS4)
      TW4=TWF+W-H2OEV
      TS4=TVS4+TA2

C     DEWATERING

      IF(INDWT)71,71,72
   72 NDWT=TF4/(DWTMC*2.1792E4)
      TNMBR(7)=1.+NDWT
      ACP(7)=TF4/(TNMBR(7)*24.)
      WK(7)=TNMBR(7)*DWTWK*ACP(7)/908.
      TS5=FDWT*TS4
      TW5=TS5*(1.-FSDWT)/FSDWT
      TF5=TS5+TW5
      GO TO 75
   71 TNMBR(7)=0.
      ACP(7)=0.
      WK(7)=0.
      TF5=0.
      TS5=0.
      TW5=0.
   75 CONTINUE
      TF6=TF4-TF5
      TW6=TW4-TW5
      TS6=TS4-TS5

C     GAS PURIFICATION

      WK(8)=0.
      IF(INGPU)81,81,82
   82 ACP(8)=DGF/0.6
      TNMBR(8)=1.
      GO TO 88
   81 ACP(8)=0.
      TNMBR(8)=0.
   88 CONTINUE
      TFFM=TSF/0.454
      WRITE(6,1001)TSF,TFFM,FSF,FVSF,FBVSF
 1001 FORMAT(1H1/// 24X,*SYSTEM PERFORMANCE*// 1X,*1. FEED MAKEUP*//9X,*
     2TOTAL SOLIDS*,14X,E11.4,* KG/D (*,E10.4,* LB/D)*/9X,*SOLIDS FRACTI
     3ON IN FEED*,14X,F6.3/9X,*VOLATILE SOLIDS FRACTION IN SOLIDS*,3X,F6
     4.3/9X,*BIODEGR. VOL. SOL. FRACT. IN SOLIDS*,2X,F6.3// 1X,*2. PRETR
     5EATMENT*)
      IF(INSHF)800,800,801
  800 WRITE(6,1002)
 1002 FORMAT(/5X,*SHREDDING-NONE USED*)
      GO TO 805
  801 ACM(1)=ACP(1)/908.
      WRITE(6,1003)TNMBR(1),ACP(1),ACM(1),WK(1)
 1003 FORMAT(/5X,*SHREDDING*/9X,*NUMBER OF UNITS*,20X,F5.0/9X,*CAPACITY
     2PER UNIT*,4X,E11.4,* KG/HR (*,E10.4,* TONS/HR)*/9X,*POWER REQUIREM
     3ENT*,20X,E11.4,* HP*)
  805 IF(INDGR)806,806,807
  806 WRITE(6,1004)
 1004 FORMAT(/5X,*DEGRITTER-NONE USED*)
      GO TO 808
```

```
  807 ACM(2)=ACP(2)/908.
      WRITE(6,1005)EDG,TNMBR(2),ACP(2),ACM(2),WK(2)
 1005 FORMAT(/5X,*DEGRITTER*/9X, *EFFICIENCY OF ASH REMOVAL*,13X,F5.3/9X
     2,*NUMBER OF UNITS*,20X,F5.0/9X,*CAPACITY PER UNIT*,4X,E11.4,* KG/H
     3R (*,E10.4,* TONS/HR)*/9X,*POWER REQUIREMENT*,20X,E11.4,* HP*)
  808 IF(INCPR)810,810,811
  810 WRITE(6,1006)
 1006 FORMAT(/5X,*CHEMICAL PRETREATMENT-NONE USED*)
      GO TO 815
  811 ACM(3)=ACP(3)*35.31
      TCHMM=TCHEM/908.
      WRITE(6,1007)CHEM,ECHP, TNMBR(3),ACP(3),ACM(3),TCHEM,TCHMM,WK(3)
 1007 FORMAT(/5X,*CHEMICAL PRETREATMENT*/9X,*RATIO OF CHEMICAL TO FEED*,
     213X,F5.3/9X,*CONVERSION EFFICIENCY*,16X,F6.3/9X,*NUMBER OF UNITS*,
     320X,F5.0/9X,*CAPACITY PER UNIT*,4X,E11.4,* CU.M. (*,E10.4,* CU.FT.
     4)*/9X,*CHEMICAL REQUIREMENT *,E11.4,* KG/HR (*,E10.4,* TONS/HR)*/9
     5X,*POWER REQUIREMENT*,20X,E11.4,* HP*)
  815 IF(INSTR)820,821,822
  820 WRITE(6,1008)
 1008 FORMAT(/ 1X,*3. STORAGE-NONE USED*)
      GO TO 829
  821 WRITE(6,1009)
 1009 FORMAT(/ 1X,*3. STORAGE-NO MIXING*)
      GO TO 825
  822 WRITE(6,1010)
 1010 FORMAT(/ 1X,*3. STORAGE-WITH MIXING*)
  825 ACM(4)=ACP(4)*35.31
      WRITE(6,1011)TNMBR(4),ACP(4),ACM(4),WK(4)
 1011 FORMAT(/9X,*NUMBER OF UNITS*,20X,F5.0/9X,*CAPACITY PER UNIT*,4X,E1
     21.4,* CU.M. (*,E10.4,* CU FT.)*/9X,*POWER REQUIREMENT*,20X,E11.4,*
     3 HP*)
  829 ACM(5)=ACP(5)*35.31
      WRITE(6,1012)
 1012 FORMAT(/ 1X,*4. DIGESTION*)
      IF(INDIG)831,833,839
  831 WRITE(6,1013)
 1013 FORMAT(/5X,*CSTR*)
      GO TO 835
  839 IF(INDIG-1)832,832,838
  832 WRITE(6,1014)
 1014 FORMAT(/5X,*PLUG FLOW*)
      GO TO 835
  838 WRITE(6,1039)
 1039 FORMAT(/5X,*BATCH*)
      GO TO 835
  833 WRITE(6,1015)
 1015 FORMAT(/5X,*CSTR IN SERIES*)
  835 DGPM=DGP*0.03725
      DGPMM=DGP*0.03531
      WRITE(6,1016)TNMBR(5),ACP(5),ACM(5),TC,HRTDG,FBVSC,WK(5),GPD,DGPM,
     1DGPMM
 1016 FORMAT( 9X,*NUMBER OF UNITS*,20X,F5.0/9X,*CAPACITY PER UNIT*,4X,E1
     21.4,* CU.M. (*,E10.4,* CU.FT.)*/9X,*DIGESTER TEMPERATURE*,16X,
     3F4.0,*C*/9X,*RETENTION TIME*,15X,F6.1,*DAYS*/9X,*CONVERSION EFFICI
     4ENCY*,16X,F6.3/9X,*POWER REQUIREMENT*,20X,
     4E11.4,*HP*//5X,*GAS PRODUCTION*,26X,F7.3,* VVD*/41X,E11.4,* GJ/D*/
     541X,*(*,E10.4,* MM BTU/D)*)
```

```
      IF(INHEX)841,841,842
  842 ACM(6)=ACP(6)*10.76
      WRITE(6,1017)TNMBR(6),ACP(6),ACM(6)
 1017 FORMAT( 5X,*HEAT EXCHANGER*/ 9X,*NUMBER OF UNITS*,20X,F5.0/9X,*HEA
     1T EXCHANGER AREA*, 2X,E11.4,* SQ.M. (*,E11.4,* SQ.FT.)*)
  841 HREQG=HREQ/1.E9
      HREQM=HREQ/1055.E6
      WRITE(6,1018)HREQG,HREQM
 1018 FORMAT(/5X,*TOTAL HEAT REQUIREMENT*,3X,E11.4,* GJ/D (*,E10.4,* MM
     2BTU/D)*)
      IF(INDWT)850,850,851
  850 WRITE(6,1019)
 1019 FORMAT(/ 2X,*5. DEWATERING-NONE USED*)
      GO TO 852
  851 ACM(7)=ACP(7)/908.
      WRITE(6,1020)TNMBR(7),ACP(7),ACM(7),WK(7)
 1020 FORMAT(//1X,*5. DEWATERING */9X,*NUMBER OF UNITS*,20X,F5.0/9X,*CAP
     2ACITY PER UNIT*,4X,E11.4,* KG/HR (*,E10.4,* TONS/HR)*/9X,*POWER RE
     3QUIREMENT*,20X,E11.4,* HP*)
  852 IF(INGPU)860,860,861
  860 WRITE(6,1021)
 1021 FORMAT(/ 1X,*6. GAS PURIFICATION-NONE USED*)
      GO TO 862
  861 ACM(8)=ACP(8)*35.31
      WRITE(6,1022)ACP(8),ACM(8)
 1022 FORMAT(/1X,*6. GAS PURIFICATION*//9X,*CAPACITY PER UNIT *,E11.4,
     1* CU.M/D (*,E10.4,* CU.FT/D)*)
  862 CONTINUE

C     SYSTEM COSTS

      SUBCS=0.
      DO 100 J=1,KK
      CCOS(J)=CONS(J)*(ACP(J)**EX(J))*OINDX/OIND(J)
      TCOS(J)=CCOS(J)*TNMBR(J)
  100 SUBCS=SUBCS+TCOS(J)
      WRITE(6,2000)
C     SHREDDER COST
      IF(INSHR)500,500,501
  501 WRITE(6,2001)TCOS(1)
  500 IF(INDGR)505,505,506
  506 WRITE(6,2002)TCOS(2)
  505 IF(INCPF)510,510,511
  511 WRITE(6,2003)TCOS(3)
  510 IF(INSTR)515,516,516
  516 WRITE(6,2004)TCOS(4)
  515 WRITE(6,2005)TCOS(5)
      IF(INHEX)520,520,521
  521 WRITE (6,2006)TCOS(6)
  520 IF(INDWT)525,525,526
  526 WRITE(6,2007)TCOS(7)
  525 IF(INGPU)530,530,531
  531 WRITE(6,2008)TCOS(8)
  530 TCOS(KK+1)=(SUBCS-TCOS(8))*PPMP
      WRITE(6,2009)TCOS(KK+1)
      SUBCS=SUBCS+TCOS(KK+1)
```

```
C       TOTAL PLANT INVESTMENT

        CSEP=SUBCS*PSEP
        SUB1=SUBCS+CSEP
        CSF=SUB1*FSF
        SUB2=SUB1+CSF
        CCOP=SUB2*FCOP
        CED=SUB2*FED
        SUB3=SUB2+CCOP+CED
        CPC=SUB3*FPC
        SUB4=SUB3+CPC
        CIDC=SUB4*FIDC
        CWC=SUB4*PWC

C       ANNUAL OPERATING COSTS


C       MATERIALS

        TCFF=CRF*TSF/908.*TNDO
        IF(INCPR)550,550,551
  551   TCCHM=CCHM*TCHEM*TNDO/908.
        GO TO 555
  550   TCCHM=0.

C       UTILITIES

  555   TWK=0.
        IF(INDIG-1)130,130,131
  131   WK(8)=WK(8)*HRTDG/TNDO
  130   CONTINUE
        DO 120 J=1,KK
  120   TWK=TWK+WK(J)
        TCWK=TWK*TNDO*CEP/5.589
        TCST=CST*HREQ*TNDO*EFFS/1.055E9

C       LABOR

        TCOL=TMP*AW*WD*365.
        TCML=SUB4*FM
        TCSV=PSV*(TCOL+TCML)
        TCAH=PAH*(TCOL+TCML+TCSV)

C       SUPPLIES

        TCSO=POPS*TCOL

C       LOCAL TAXES + INSURANCE

        TCLTI=PLTI*SUB4
        IF(INGPU)150,150,151
  151   TOCGP=1.
        GO TO 155
  150   TOCGP=0.
  155   TAGOP=TCFF+TCCHM+TCAH*(1.+1./PAH)+TCSO+TCML+TCLTI+TOCGP+TCWK+TCST
        TOWCV=CDWV*TF5*FSDWT*TNDO/908.
        TWCV=CEFV*TF6*TNDO/908.
```

```
      TCPNC=TDWCV+TWCV
      TANOP=TAGCP+TCPNC
      CSUP=PSUP*TAGOP
      SUB5=SUB4+CSUP+CIDC+CWC
      RRB=(FDBT*CIR)+(1.-FDBT)*RE
      IF(INDIG-1)138,138,139
  139 TNDO=HRTDG
  138 CONTINUE
      AGP=TVD*GFD*TNDO*(1.055E-3)
      AGPM=AGP/1.055
      TACST=TANOP+DPRC*(SUB5-CWC)+0.5*(RRB+FITR/(1.-FITR)*(1.-FDBT)*RE)*
     1(SUB5+CWC)
      UGC=TACST/AGP
      UGCM=TACST/AGPM
      WRITE(6,2020)SUBCS,CSEP,SUB1,CSF,SUB2,CCOP,CED,SUB3,CPC,SUB4,CIDC,
     2CSUP,CWC,SUB5
      CRFM=CRF/980.
      WRITE(6,2050)CRFM,CRF,TCFF
      IF(INCPR)700,700,701
  701 CCHMM=CCHM/980.
      WRITE(6,2051)CCHMM,CCHM,TCCHM
  700 CSTM=CST/1.055
      WRITE(6,2052)CEP,TCWK,CSTM,CST,TCST,TMP,WD,AW,TCOL,TCML,TCSV,TCAH,
     2TCSO,TCML,TCLTI
      IF(INGPU)720,720,721
  721 WRITE(6,2053)TOCGP
  720 WRITE(6,2054)TAGOP,TCPNC,TANOP,TACST,AGP,AGPM,UGC,UGCM
 2000 FORMAT(1H1////,24X,*CAPITAL COST ESTIMATE*//50X,*AUG 1978 COSTS ($
     2)*///5X,*INSTALLED EQUIPMENT*/)
 2001 FORMAT(10X,*SHREDDERS*,30X,F20.0)
 2002 FORMAT(10X,*DEGRITTERS*,29X,F20.0)
 2003 FORMAT(10X,*CHEMICAL PRETREATMENT*,18X,F20.0)
 2004 FORMAT(10X,*STORAGE TANKS*,26X,F20.0)
 2005 FORMAT(10X,*DIGESTERS*,30X,F20.0)
 2006 FORMAT(10X,*HEAT EXCHANGERS*,24X,F20.0)
 2007 FORMAT(10X,*DEWATERING UNITS*,23X,F20.0)
 2008 FORMAT(10X,*GAS PURIFICATION UNITS*,17X,F20.0)
 2009 FORMAT(10X,*MATERIALS HANDLING EQUIPMENT*,11X,F20.0)
 2020 FORMAT(49X,*--------------------*/1X, *SUBTOTAL*,40X,F20.0//5X,* S
     2TRUCTURES-ELECT. + PIPING*,17X,F20.0/49X,*--------------------*/1X
     3,*SUBTOTAL*,40X,  F20.0//5X,* SUPPORT FACILITIES*,25X,F20.0/49X,*-
     4-------------------*/1X,*SUBTOTAL CAPITAL INVESTMENT*,21X,F20.0//5
     5X,* CONTRACTOR OVERHEAD + PROFIT*,15X,F20.0//5X,* ENGINEERING + DE
     6SIGN*,23X,F20.0/49X,*--------------------*/1X, *SUBTOTAL PLANT INV
     6ESTMENT*,23X,F20.0//5X,* PROJECT CONTINGENCY*,24X,F20.0/49X, *----
     7----------------*/1X, *TOTAL PLANT INVESTMENT*,26X,F20.0//5X,* INT
     8EREST DURING CONSTRUCTION*,15X,F20.0//5X,* START-UP COSTS*,29X,F20
     9.0//5X,* WORKING CAPITAL*,28X,F20.0/49X, *--------------------*/1X
     7, *TOTAL CAPITAL REQUIREMENT*,23X,F20.0)
 2050 FORMAT(1H1,////,23X,* ANNUAL OPERATING COSTS*//54X,*ANNUAL COST ($
     2)*///5X,* PRODUCTION MATERIALS*//10X,* FEED - $*,F7.2,*/KG ($*,F7.
     32,*/TON)*,6X,F20.0)
 2051 FORMAT(/10X,* PRETREATMENT CHEMICAL*/18X,*$*,F7.2,*/KG ($*,F7.2,*/
     2TON)*,6X,F20.0)
 2052 FORMAT(/5X,* UTILITIES*//10X,* ELECTRIC -* ,F4.1,* CTS./KW-HR*,14X
     2, F20.0//11X,*STEAM -  $*,F5.2,*/GJ*/19X,*($*,F5.2,*/MM BTU)*,16X,
     3F20.0//6X,*LABOR*//10X,* OPERATING LABOR - MEN PER SHIFT *,F4.0/29
```

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     4X,*HRS PER SHIFT  *,F4.1/29X,*HOURLY RATE   $*,F5.2,1X,F20.0//10X,
     5* MAINTENANCE LABOR*,22X,F20.0//10X,* SUPERVISION*,28X,F20.0//5X,*
     8 ADMINISTRATION + OVERHEAD*,19X,F20.0//5X,* SUPPLIES*//10X,* OPERA
     9TING*,30X,F20.0//10X,* MAINTENANCE*,28X,F20.0//5X,* LOCAL TAXES +
     7INSURANCE*,21X,F20.0)
2053 FORMAT(/5X,* OPERATING COST - GAS PURIFICATION*,11X,F20.0)
2054 FORMAT(49X,* --------------------*,/2X,*GROSS ANNUAL OPERATING COS
     2T*,21X,F20.0//5X,* CREDITS/PENALTIES COST*,22X,F20.0/49X,* -------
     3-------------*/1X,* NET ANNUAL OPERATING COST*,23X,F20.0//1X,*XXXX
     4XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX*
     5/* XX*,65X,*XX*/* XX*,9X,*TOTAL ANNUAL CAPITAL REQUIREMENT = $*,
     6F12.0,8X,*XX*/* XX*,65X,*XX*/* XX*,9X,*TOTAL ANNUAL GAS PRODUCTION
     7 = *,F10.0,* GJ*,13X,*XX*/* XX*,38X,*(*,F10.0,* MM BTU)*,8X,*XX*/
     8* XX*,65X,*XX*/* XX*,9X,*UNIT GAS COST = $*,F7.2,*/GJ*,29X,*XX*/
     9* XX*,24X,*($*,F7.2,*/MM BTU)*,24X,*XX*/* XX*,65X,*XX*/* XXXXXXXX
     1XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX*)
  98 CONTINUE
  90 TC=60.
 199 CONTINUE
  99 CONTINUE
     END
```

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LIST OF SYMBOLS

A	ft^3 CH_4/lb converted
ACM	Equipment size
ACP	Equipment size (SI units)
AGP	Gas production (GJ/yr)
AGPM	Gas production (MM BTU/yr)
AW	Labor cost (\$/hr)
CBE	Final bio. vol. sol. conc. (lb/ft^3)
CBI	Initial bio. vol. sol. conc. (lb/ft^3)
CCHM	Cost of pretreatment chemical (\$/ton)
CCHMM	Cost of pretreatment chemical (\$/kg)
CCOP	Cost of contractor's overhead and profit
CCOS	Cost of equipment (per unit)
CDWV	Credit/penalty for dewatered output (\$/ton)
CED	Cost of engineering and design
CEFV	Credit/penalty for digester effluent (\$/ton)
CEP	Cost of electricity (¢/kwh)
CF	Heat capacity of digester slurry (Btu/lb°F)
CHEM	Amount of pretreatment chemical per solids converted
CIDC	Cost of interest during construction
CONS	Cost constant for equipment
CPC	Cost of project contingency
CPRMC	Maximum capacity of chemical pretreatment equipment (ft^3)
CPRWK	Work for chemical pretreatment (hp/ft^3)
CRF	Cost of feedstock (\$/ton)
CRFM	Cost of feedstock (\$/kg)
CSEP	Cost of electricity and piping
CSF	Cost of support facilities
CST	Cost of steam (\$/MM BTU)
CSTM	Cost of steam (\$/GJ)
CSUP	Startup cost
CWC	Working capital

DGP	Gas production (m^3/day)
DGPM	Gas production (GJ/day)
DGPMM	Gas production (MM BTU/day)
DGRMC	Degritter maximum capacity (tons/hr)
DGRWK	Degritter work (HP)
DIGMC	Digester maximum capacity (ft^3)
DIGWK	Digester work (HP/ft^3)
DIR	Interest on debt
DNS	Digester slurry density (lb/ft^3)
DPRC	Rate of depreciation
DWTMC	Dewatering maximum capacity (T/hr)
DWTWK	Dewatering work (HP/ton/hr)
EA	Activation energy for Arrhenius rate equation (cal/mole)
ECHP	Efficiency of chemical pretreatment
EDG	Efficiency of degritting
EDWT	Efficiency of dewatering
EFFS	Efficiency of steam utilization
EX	Equipment cost scale exponent
FBVSC	Fraction of bio. vol. solids converted to CH$_4$
FBVSF	Fraction of bio. vol. solids in feed
FD	Feedstock rate (tons solids/day)
FDBT	Fraction debt
FITR	Federal income tax rate
FSDF	Fraction of solids in digester feed
FSDWT	Fraction of solids in dewatered stream
FSF	Fraction of solids in feed
FSVF	Fraction of vol. solids in feed
GPD	Gas production (VVD)
HEFF	Heat content of digester effluent (J/day)
HEVAP	Heat content of moisture in gas stream (J/day)
HEXMC	Heat exchanger maximum capacity (ft^2)
HFEED	Heat content of digester input stream (J/day)

HGAS	Heat content of gas stream (J/day)
HLOSS	Heat loss from digester walls (J/day)
HREQ	Digester heat requirement (J/day)
HREQG	Digester heat requirement (GJ/day)
HREQM	Digester heat requirement (MM BTU/day)
HRTCP	Retention time for chemical pretreatment (hr)
HRTDG	Retention time for digestion (days)
HRTST	Retention time for storage (days)
HRXN	Heat of reaction in digester (J/day)
HTC	Heat transfer coefficient in heat exchanger
HW	Heat content of make-up water
H2OEV	Evaporated water in gas stream (kg/day)
I	Dummy variable
IJ	Dummy variable
IJK	Dummy variable
INCPR	Dummy variable for use of chemical pretreatment
INDGR	Dummy variable for use of degritter
INDIG	Dummy variable for use of digester
INDWT	Dummy variable for use of dewatering
INGPU	Dummy variable for use of gas purification
INHEX	Dummy variable for use of heat exchanger
INSHR	Dummy varaible for use of shredder
INSTR	Dummy variable for use of storage
J	Dummy variable
JJ	Dummy variable
JTMP	Dummy variable
K	Dummy variable
KK	Dummy variable
NCPR	Number of chemical pretreatment units
NDGR	Number of degritters
NDIG	Number of digesters
NDWT	Number of dewaterers
NHEX	Number of heat exchangers

NSHR	Number of shredders
NSTR	Number of storage units
OIND	Cost index for year of equipment cost estimation
OINDX	Current cost index
PAH	Fraction of labor cost for administration and overhead
PCOP	Fraction of equipment cost for contractor's overhead and profit
PED	Fraction of equipment cost for engineering and design
PIDC	Fraction of total plant investment for interest during construction
PLTI	Fraction of total plant investment for local taxes and insurance
PM	Fraction of total plant investment for maintenance labor
--	Fraction of total plant investment for maintenance supplies
POPS	Fraction of operating labor cost for operating supplies
PPC	Fraction of subtotal plant investment for project contingency
PPMP	Fraction of equipment cost for materials handling equipment
PSEP	Fraction of equipment cost for electricity and piping
PSF	Fraction of equipment cost for support facility
PSUP	Fraction of annual gross operating cost for plant start-up
PSV	Fraction of labor cost for supervision
PWC	Fraction of total plant investment for working capital
RE	Return on equity
RK	Rate constant (day^{-1})
RKO	Rate equation constant (day^{-1})
RNUM	Number of digesters in series
RRB	Return on rate base
SHRMC	Shredder maximum capacity (tons/hr)
SHRWK	Shredder work (HP/tons/hr)
STF	Stream factor (fraction of year in operation)
STRMC	Storage maximum capacity (ft^3)
STRWK	Storage work (HP/ft^3)
SUBCS	Subtotal equipment cost
SUB1	Total equipment cost
SUB2	Cost subtotal

SUB3	Subtotal plant investment
SUB4	Total plant investment
SUB5	Total capital requirement
TA	Ambient temperature (°C)
TACST	Annual capital requirement
TAF	Total ash in feed (kg/day)
TAGOP	Annual gross operating cost
TAHEX	Total required heat exchanger area (m^2)
TANOP	Annual net operating cost
TAR	Total ash removed in degritter (kg/day)
TA1	Total ash leaving degritter (kg/day)
TA2	Total ash leaving chemical pretreatment (kg/day)
TBVSF	Total bio. vol. solids in feed (kg/day)
TBVS2	Total bio. vol. solids leaving chemical pretreatment (kg/day)
TBVS4	Total bio. vol. solids leaving digester (kg/day)
TC	Digester temperature (°C)
TCAH	Annual cost for administration and overhead
TCBVS	Vol. sol. converted to bio. vol. sol. in chem. pretreatment (kg/day)
TCCHM	Annual cost for pretreatment chemicals
TCFF	Annual cost for feedstock
TCHEM	Pretreatment chemical requirement (kg/hr)
TCHMM	Pretreatment chemical requirement (tons/hr)
TCLTI	Annual cost for local taxes and insurance
TCML	Annual cost for maintenance labor
TCOL	Annual cost for operating labor
TCOS	Total equipment cost for each process step
TCPNC	Annual credit/penalty cost
TCSO	Annual cost for operating supplies
TCST	Annual cost for steam
TCSV	Annual cost for supervision
TCWK	Annual cost for electricity
TDWCV	Annual credit/penalty for dewatered solids
TFF	Total feed stream (kg/day)
TFFM	Total solids in feed (lb/day)

TF1	Total flow from degritter (kg/day)
TF2	Total flow from chemical pretreatment (kg/day)
TF3	Total flow from storage (kg/day)
TF4	Total flow from digester (kg/day)
TF5	Total flow from dewatering, high solids stream (kg/day)
TF6	Total flow from dewatering, low solids stream (kg/day)
TGRAD	Temperature difference in heat exchanger
TMP	Manpower requirement
TNDV	Number of days of operation per year
TNMBR	Number of units per processing step
TOCGP	Annual operating cost for gas purification
TPEX	Temperature of digester effluent leaving heat exchanger (°C)
TPF	Temperature of feed streams (°C)
TSF	Solids in feedstock (kg/day)
TS1	Solids leaving degritter (kg/day)
TS2	Solids leaving chemical pretreatment (kg/day)
TS4	Solids leaving digester (kg/day)
TS5	Solids leaving dewaterer, high solids stream (kg/day)
TS6	Solids leaving dewaterer, low solids stream (kg/day)
TVD	Total digester volume (ft^3)
TVSF	Vol. solids in feed (kg/day)
TVS2	Vol. solids leaving chemical pretreatment (kg/day)
TVS4	Vol. solids leaving digester (kg/day)
TWCV	Annual credit/penalty for dewatered low solids stream
TWF	Water content in feedstock (kg/day)
TWK	Total horsepower
TW4	Water leaving digester (kg/day)
TW5	Water leaving dewaterer, high solids stream (kg/day)
TW6	Water leaving dewaterer, low solids stream (kg/day)
UGC	Unit gas cost ($/GJ)
UGCM	Unit gas cost ($/MM BTU)
W	Make-up water required (kg/day)

WD	Hours per work day
WK	Work per process step (HP)
XX	Mole fraction of water in digester gas stream

APPENDIX F

Computer Output

Table F.1

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                       .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                      .130
        VOLATILE SOLIDS FRACTION IN SOLIDS           .800
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS          .440

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT          .1362E+04 CU.M. ( .4808E+05 CU FT.)
        POWER REQUIREMENT                           .9615E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT          .5727E+04 CU.M. ( .2022E+06 CU.FT.)
        DIGESTER TEMPERATURE                  37.0
        RETENTION TIME                      8.4DAYS
        CONVERSION EFFICIENCY                       .500
        POWER REQUIREMENT                           .4845E+02HP

    GAS PRODUCTION                                1.218  VVD
                                               .2599E+03 GJ/D
                                             ( .2464E+03 MM BTU/D)

    HEAT EXCHANGER
        NUMBER OF UNITS                           1.
        HEAT EXCHANGER AREA        .3277E+03 SQ.M. (  .3526E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT         .9580E+01 GJ/D ( .9081E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	170955.
DIGESTERS	713568.
HEAT EXCHANGERS	19786.
MATERIALS HANDLING EQUIPMENT	45215.

SUBTOTAL	949525.
STRUCTURES-ELECT. + PIPING	37981.

SUBTOTAL	987506.
SUPPORT FACILITIES	49375.

SUBTOTAL CAPITAL INVESTMENT	1036881.
CONTRACTOR OVERHEAD + PROFIT	103688.
ENGINEERING + DESIGN	51844.

SUBTOTAL PLANT INVESTMENT	1192414.
PROJECT CONTINGENCY	178862.

TOTAL PLANT INVESTMENT	1371276.
INTEREST DURING CONSTRUCTION	123415.
START-UP COSTS	37760.
WORKING CAPITAL	27426.

TOTAL CAPITAL REQUIREMENT	1559876.

```
                         ANNUAL OPERATING COSTS

                                                          ANNUAL COST ($)

     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                            0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                   8827.

          STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                      9546.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE    $ 6.00                    35040.

          MAINTENANCE LABOR                                          20569.

          SUPERVISION                                                 8341.

     ADMINISTRATION + OVERHEAD                                       38370.

     SUPPLIES

          OPERATING                                                  10512.

          MAINTENANCE                                                20569.

     LOCAL TAXES + INSURANCE                                         37024.
                                                       --------------------
 GROSS ANNUAL OPERATING COST                                        188799.

     CREDITS/PENALTIES COST                                              0.
                                                       --------------------
 NET ANNUAL OPERATING COST                                          188799.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                          XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      376228.                XX
XX                                                                          XX
XX         TOTAL ANNUAL GAS PRODUCTION =      85394. GJ                     XX
XX                                            80942. MM BTU)                XX
XX                                                                          XX
XX         UNIT GAS COST = $    4.41/GJ                                     XX
XX                       ($    4.65/MM BTU)                                 XX
XX                                                                          XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.2

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

       TOTAL SOLIDS                     .4540E+05 KG/D ( .1000E+06 LB/D)
       SOLIDS FRACTION IN FEED                    .130
       VOLATILE SOLIDS FRACTION IN SOLIDS         .800
       BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .440

2. PRETREATMENT

     SHREDDING-NONE USED

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

       NUMBER OF UNITS                          1.
       CAPACITY PER UNIT            .6808E+03 CU.M. ( .2404E+05 CU FT.)
       POWER REQUIREMENT                          .4808E+01 HP

4. DIGESTION

     CSTR
       NUMBER OF UNITS                          1.
       CAPACITY PER UNIT            .9711E+03 CU.M. ( .3429E+05 CU.FT.)
       DIGESTER TEMPERATURE                   60.0
       RETENTION TIME                     2.9DAYS
       CONVERSION EFFICIENCY                      .650
       POWER REQUIREMENT                          .6858E+01HP

     GAS PRODUCTION                             4.671 VVD
                                            .1690E+03 GJ/D
                                          ( .1602E+03 MM BTU/D)

     HEAT EXCHANGER
       NUMBER OF UNITS                          1.
       HEAT EXCHANGER AREA          .3351E+03 SQ.M. (  .3606E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT         .6121E+01 GJ/D ( .5802E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	119219.
DIGESTERS	188550.
HEAT EXCHANGERS	20123.
MATERIALS HANDLING EQUIPMENT	16395.
SUBTOTAL	344286.
STRUCTURES-ELECT. + PIPING	13771.
SUBTOTAL	358058.
SUPPORT FACILITIES	17903.
SUBTOTAL CAPITAL INVESTMENT	375961.
CONTRACTOR OVERHEAD + PROFIT	37596.
ENGINEERING + DESIGN	18798.
SUBTOTAL PLANT INVESTMENT	432355.
PROJECT CONTINGENCY	64853.
TOTAL PLANT INVESTMENT	497208.
INTEREST DURING CONSTRUCTION	44749.
START-UP COSTS	23549.
WORKING CAPITAL	9944.
TOTAL CAPITAL REQUIREMENT	575450.

```
                     ANNUAL OPERATING COSTS

                                               ANNUAL COST ($)


     PRODUCTION MATERIALS

        FEED - $   0.00/KG ($   0.00/TON)                        0.

     UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                             2057.

        STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                              6099.

     LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00               35040.

        MAINTENANCE LABOR                                     7458.

        SUPERVISION                                           6375.

     ADMINISTRATION + OVERHEAD                               29324.

     SUPPLIES

        OPERATING                                            10512.

        MAINTENANCE                                           7458.

     LOCAL TAXES + INSURANCE                                 13425.
                                                  --------------------
  GROSS ANNUAL OPERATING COST                               117747.

     CREDITS/PENALTIES COST                                      0.
                                                  --------------------
  NET ANNUAL OPERATING COST                                 117747.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                            XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      186887.   XX
XX                                                            XX
XX        TOTAL ANNUAL GAS PRODUCTION =       55506. GJ       XX
XX                                    (       52613. MM BTU)  XX
XX                                                            XX
XX        UNIT GAS COST = $    3.37/GJ                        XX
XX                       ($    3.55/MM BTU)                   XX
XX                                                            XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.3

```
                       SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                    .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                   .130
        VOLATILE SOLIDS FRACTION IN SOLIDS        .800
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS       .440

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT       .1362E+04 CU.M. ( .4808E+05 CU FT.)
        POWER REQUIREMENT                         .9615E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT       .1942E+04 CU.M. ( .6858E+05 CU.FT.)
        DIGESTER TEMPERATURE                    60.C
        RETENTION TIME                     2.9DAYS
        CONVERSION EFFICIENCY                     .650
        POWER REQUIREMENT                         .1372E+02HP

    GAS PRODUCTION                               4.671 VVD
                                              .3379E+03 GJ/D
                                            ( .3203E+03 MM BTU/D)

    HEAT EXCHANGER
        NUMBER OF UNITS                          1.
        HEAT EXCHANGER AREA     .6702E+03 SQ.M. (  .7212E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT      .1002E+02 GJ/D ( .9499E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	170955.
DIGESTERS	317101.
HEAT EXCHANGERS	33842.
MATERIALS HANDLING EQUIPMENT	26095.
SUBTOTAL	547994.
STRUCTURES-ELECT. + PIPING	21920.
SUBTOTAL	569913.
SUPPORT FACILITIES	28496.
SUBTOTAL CAPITAL INVESTMENT	598409.
CONTRACTOR OVERHEAD + PROFIT	59841.
ENGINEERING + DESIGN	29920.
SUBTOTAL PLANT INVESTMENT	688171.
PROJECT CONTINGENCY	103226.
TOTAL PLANT INVESTMENT	791396.
INTEREST DURING CONSTRUCTION	71226.
START-UP COSTS	28333.
WORKING CAPITAL	15828.
TOTAL CAPITAL REQUIREMENT	907283.

```
                    ANNUAL OPERATING COSTS

                                                  ANNUAL COST ($)


   PRODUCTION MATERIALS

        FEED - $   0.00/KG ($   0.00/TON)                     0.

   UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                          4114.

        STEAM -  $ 3.79/GJ
                ($ 4.00/MM BTU)                            9986.

   LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00            35040.

        MAINTENANCE LABOR                                 11871.

        SUPERVISION                                        7037.

   ADMINTSTRATION + OVERHEAD                              32369.

   SUPPLIES

        OPERATING                                         10512.

        MAINTENANCE                                       11871.

   LOCAL TAXES + INSURANCE                                21368.
                                                   --------------------
GROSS ANNUAL OPERATING COST                              144166.

   CREDITS/PENALTIES COST                                     0.
                                                   --------------------
NET ANNUAL OPERATING COST                                144166.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                 XX
XX      TOTAL ANNUAL CAPITAL REQUIREMENT = $     253179.           XX
XX                                                                 XX
XX      TOTAL ANNUAL GAS PRODUCTION =      111013. GJ              XX
XX                                  (      105225. MM BTU)         XX
XX                                                                 XX
XX      UNIT GAS COST = $    2.28/GJ                               XX
XX                     ($    2.41/MM BTU)                          XX
XX                                                                 XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.4

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                       .4540E+06 KG/D ( .1000E+07 LB/D)
        SOLIDS FRACTION IN FEED                       .130
        VOLATILE SOLIDS FRACTION IN SOLIDS            .800
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS           .440

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                             1.
        CAPACITY PER UNIT            .6808E+04 CU.M. ( .2404E+06 CU FT.)
        POWER REQUIREMENT                             .4808E+02 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                             1.
        CAPACITY PER UNIT            .9711E+04 CU.M. ( .3429E+06 CU.FT.)
        DIGESTER TEMPERATURE                      60.C
        RETENTION TIME                         2.9DAYS
        CONVERSION EFFICIENCY                         .650
        POWER REQUIREMENT                             .6858E+02HP

    GAS PRODUCTION                                4.671 VVD
                                               .1690E+04 GJ/D
                                             ( .1602E+04 MM BTU/D)

    HEAT EXCHANGER
        NUMBER OF UNITS                             4.
        HEAT EXCHANGER AREA          .8376E+03 SQ.M. (  .9014E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT           .3238E+02 GJ/D ( .3069E+02 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	394773.
DIGESTERS	1060293.
HEAT EXCHANGERS	160029.
MATERIALS HANDLING EQUIPMENT	80755.
SUBTOTAL	1695849.
STRUCTURES-ELECT. + PIPING	67834.
SUBTOTAL	1763683.
SUPPORT FACILITIES	88184.
SUBTOTAL CAPITAL INVESTMENT	1851867.
CONTRACTOR OVERHEAD + PROFIT	185187.
ENGINEERING + DESIGN	92593.
SUBTOTAL PLANT INVESTMENT	2129647.
PROJECT CONTINGENCY	319447.
TOTAL PLANT INVESTMENT	2449094.
INTEREST DURING CONSTRUCTION	220418.
START-UP COSTS	59655.
WORKING CAPITAL	48982.
TOTAL CAPITAL REQUIREMENT	2778150.

```
                        ANNUAL OPERATING COSTS

                                                      ANNUAL COST ($)

    PRODUCTION MATERIALS

         FEED - $    0.00/KG ($    0.00/TON)                          0.

    UTILITIES

         ELECTRIC - 5.0 CTS./KW-HR                                20570.

         STEAM -  $ 3.79/GJ
                 ($ 4.00/MM BTU)                                  32263.

    LABOR

         OPERATING LABOR - MEN PER SHIFT    2.
                           HRS PER SHIFT    8.0
                           HOURLY RATE    $ 6.00                  35040.

         MAINTENANCE LABOR                                        36736.

         SUPERVISION                                              10766.

    ADMINISTRATION + OVERHEAD                                     49526.

    SUPPLIES

         OPERATING                                                10512.

         MAINTENANCE                                              36736.

    LOCAL TAXES + INSURANCE                                       66126.
                                                     --------------------
GROSS ANNUAL OPERATING COST                                      298275.

    CREDITS/PENALTIES COST                                            0.
                                                     --------------------
NET ANNUAL OPERATING COST                                        298275.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      632089.         XX
XX                                                                  XX
XX        TOTAL ANNUAL GAS PRODUCTION =      555063. GJ             XX
XX                                         (    526125. MM BTU)     XX
XX                                                                  XX
XX        UNIT GAS COST = $    1.14/GJ                              XX
XX                       ($    1.20/MM BTU)                         XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.5

```
                        SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                   .9080E+06 KG/D ( .2000E+07 LB/D)
        SOLIDS FRACTION IN FEED                    .130
        VOLATILE SOLIDS FRACTION IN SOLIDS         .800
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .440

2. PRETREATMENT

     SHREDDING-NONE USED

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                            2.
        CAPACITY PER UNIT         .6808E+04 CU.M. ( .2404E+06 CU FT.)
        POWER REQUIREMENT                          .9615E+02 HP

4. DIGESTION

     CSTR
        NUMBER OF UNITS                            1.
        CAPACITY PER UNIT         .1942E+05 CU.M. ( .6858E+06 CU.FT.)
        DIGESTER TEMPERATURE                   60.C
        RETENTION TIME                     2.9DAYS
        CONVERSION EFFICIENCY                       .650
        POWER REQUIREMENT                           .1372E+03HP

     GAS PRODUCTION                             4.671 VVD
                                             .3379E+04 GJ/D
                                          ( .3203E+04 MM BTU/D)

     HEAT EXCHANGER
        NUMBER OF UNITS                            8.
        HEAT EXCHANGER AREA       .8378E+03 SQ.M. (  .9014E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT       .5446E+02 GJ/D ( .5162E+02 MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	789545.
DIGESTERS	1783193.
HEAT EXCHANGERS	320058.
MATERIALS HANDLING EQUIPMENT	144640.
SUBTOTAL	3037435.
STRUCTURES-ELECT. + PIPING	121497.
SUBTOTAL	3158933.
SUPPORT FACILITIES	157947.
SUBTOTAL CAPITAL INVESTMENT	3316880.
CONTRACTOR OVERHEAD + PROFIT	331688.
ENGINEERING + DESIGN	165844.
SUBTOTAL PLANT INVESTMENT	3814411.
PROJECT CONTINGENCY	572162.
TOTAL PLANT INVESTMENT	4386573.
INTEREST DURING CONSTRUCTION	394792.
START-UP COSTS	95138.
WORKING CAPITAL	87731.
TOTAL CAPITAL REQUIREMENT	4964234.

```
                         ANNUAL OPERATING COSTS

                                                        ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                            0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                  41139.

          STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                                    54261.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                     35040.

          MAINTENANCE LABOR                                          65799.

          SUPERVISION                                                15126.

     ADMINISTRATION + OVERHEAD                                       69579.

     SUPPLIES

          OPERATING                                                  10512.

          MAINTENANCE                                                65799.

     LOCAL TAXES + INSURANCE                                        118437.
                                                      --------------------
  GROSS ANNUAL OPERATING COST                                       475691.

     CREDITS/PENALTIES COST                                              0.
                                                      --------------------
  NET ANNUAL OPERATING COST                                         475691.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                    XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $    1072182.           XX
XX                                                                    XX
XX         TOTAL ANNUAL GAS PRODUCTION =    1110125. GJ               XX
XX                                     (   1052251. MM BTU)           XX
XX                                                                    XX
XX         UNIT GAS COST = $     .97/GJ                               XX
XX                        ($   1.02/MM BTU)                           XX
XX                                                                    XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.6

```
                       SYSTEM PERFORMANCE

1. FEED MAKEUP

      TOTAL SOLIDS                    .9080E+05 KG/D ( .2000E+06 LB/D)
      SOLIDS FRACTION IN FEED                   .700
      VOLATILE SOLIDS FRACTION IN SOLIDS        .500
      BIODEGR. VOL. SOL. FRACT. IN SOLIDS       .220

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER
      EFFICIENCY OF ASH REMOVAL                1.000
      NUMBER OF UNITS                          1.
      CAPACITY PER UNIT      .5405E+04 KG/HR ( .5952E+01 TONS/HR)
      POWER REQUIREMENT                         .2000E+01 HP

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

      NUMBER OF UNITS                          1.
      CAPACITY PER UNIT      .5206E+03 CU.M. ( .1838E+05 CU FT.)
      POWER REQUIREMENT                         .3676E+01 HP

4. DIGESTION

    CSTR
      NUMBER OF UNITS                          1.
      CAPACITY PER UNIT      .4067E+04 CU.M. ( .1436E+06 CU.FT.)
      DIGESTER TEMPERATURE                   37.0
      RETENTION TIME                   15.6DAYS
      CONVERSION EFFICIENCY                     .650
      POWER REQUIREMENT                         .2872E+02HP

    GAS PRODUCTION                             1.115 VVD
                                             .1690E+03 GJ/D
                                           ( .1602E+03 MM BTU/D)
    HEAT EXCHANGER
      NUMBER OF UNITS                          1.
      HEAT EXCHANGER AREA    .1253E+03 SQ.M. (  .1348E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT   .3700E+01 GJ/D ( .3507E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	2243.
STORAGE TANKS	103697.
DIGESTERS	551973.
HEAT EXCHANGERS	9621.
MATERIALS HANDLING EQUIPMENT	33376.
SUBTOTAL	700906.
STRUCTURES-ELECT. + PIPING	28036.
SUBTOTAL	728942.
SUPPORT FACILITIES	36447.
SUBTOTAL CAPITAL INVESTMENT	765389.
CONTRACTOR OVERHEAD + PROFIT	76539.
ENGINEERING + DESIGN	38269.
SUBTOTAL PLANT INVESTMENT	880198.
PROJECT CONTINGENCY	132030.
TOTAL PLANT INVESTMENT	1012228.
INTEREST DURING CONSTRUCTION	91100.
START-UP COSTS	31038.
WORKING CAPITAL	20245.
TOTAL CAPITAL REQUIREMENT	1154610.

```
                  ANNUAL OPERATING COSTS

                                           ANNUAL COST ($)


   PRODUCTION MATERIALS

       FEED - $   0.00/KG ($   0.00/TON)                   0.

   UTILITIES

       ELECTRIC - 3.0 CTS./KW-HR                        6065.

       STEAM -  $ 3.79/GJ
              ($ 4.00/MM BTU)                           3687.

   LABOR

       OPERATING LABOR - MEN PER SHIFT    2.
                         HRS PER SHIFT    8.0
                         HOURLY RATE    $ 6.00         35040.

       MAINTENANCE LABOR                               15183.

       SUPERVISION                                      7534.

   ADMINISTRATION + OVERHEAD                           34654.

   SUPPLIES

       OPERATING                                       10512.

       MAINTENANCE                                     15183.

   LOCAL TAXES + INSURANCE                             27330.
                                              ----------------
GROSS ANNUAL OPERATING COST                           155189.

   CREDITS/PENALTIES COST                                  0.
                                              ----------------
NET ANNUAL OPERATING COST                             155189.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                      XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $     293921. XX
XX                                                      XX
XX       TOTAL ANNUAL GAS PRODUCTION =      55506. GJ     XX
XX                                    (     52613. MM BTU)XX
XX                                                      XX
XX       UNIT GAS COST = $   5.30/GJ                      XX
XX                      ($   5.59/MM BTU)                 XX
XX                                                      XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.7

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                     .4540E+05 KG/D ( .1000E+06 LB/D)
        SOLIDS FRACTION IN FEED                     .700
        VOLATILE SOLIDS FRACTION IN SOLIDS          .500
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS         .220

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER
        EFFICIENCY OF ASH REMOVAL                  1.000
        NUMBER OF UNITS                            1.
        CAPACITY PER UNIT       .2702E+04 KG/HR ( .2976E+01 TONS/HR)
        POWER REQUIREMENT                           .2000E+01 HP

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                            1.
        CAPACITY PER UNIT       .2603E+03 CU.M. ( .9191E+04 CU FT.)
        POWER REQUIREMENT                           .1838E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                            1.
        CAPACITY PER UNIT       .7997E+03 CU.M. ( .2824E+05 CU.FT.)
        DIGESTER TEMPERATURE                     60.C
        RETENTION TIME                      6.1DAYS
        CONVERSION EFFICIENCY                       .800
        POWER REQUIREMENT                           .5648E+01HP

    GAS PRODUCTION                                 3.490 VVD
                                               .1040E+03 GJ/D
                                             ( .9856E+02 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                            1.
        HEAT EXCHANGER AREA     .1281E+03 SQ.M. (  .1379E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT      .3480E+01 GJ/D ( .3299E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	1757.
STORAGE TANKS	72315.
DIGESTERS	162999.
HEAT EXCHANGERS	9784.
MATERIALS HANDLING EQUIPMENT	12343.
SUBTOTAL	259198.
STRUCTURES-ELECT. + PIPING	10368.
SUBTOTAL	269566.
SUPPORT FACILITIES	13478.
SUBTOTAL CAPITAL INVESTMENT	283045.
CONTRACTOR OVERHEAD + PROFIT	28304.
ENGINEERING + DESIGN	14152.
SUBTOTAL PLANT INVESTMENT	325501.
PROJECT CONTINGENCY	48825.
TOTAL PLANT INVESTMENT	374326.
INTEREST DURING CONSTRUCTION	33689.
START-UP COSTS	21236.
WORKING CAPITAL	7487.
TOTAL CAPITAL REQUIREMENT	436738.

```
                    ANNUAL OPERATING COSTS

                                              ANNUAL COST ($)

      PRODUCTION MATERIALS

           FEED - $    0.00/KG ($    0.00/TON)                  0.

      UTILITIES

           ELECTRIC - 3.0 CTS./KW-HR                         1673.

           STEAM -  $ 3.79/GJ
                   ($ 4.00/MM BTU)                           3468.

      LABOR

           OPERATING LABOR - MEN PER SHIFT    2.
                             HRS PER SHIFT    8.0
                             HOURLY RATE   $ 6.00           35040.

           MAINTENANCE LABOR                                 5615.

           SUPERVISION                                       6098.

      ADMINISTRATION + OVERHEAD                             28052.

      SUPPLIES

           OPERATING                                        10512.

           MAINTENANCE                                       5615.

      LOCAL TAXES + INSURANCE                               10107.
                                                -------------------
  GROSS ANNUAL OPERATING COST                              106179.

      CREDITS/PENALTIES COST                                    0.
                                                -------------------
  NET ANNUAL OPERATING COST                                106179.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $     158652.          XX
XX                                                                XX
XX        TOTAL ANNUAL GAS PRODUCTION =      34158. GJ              XX
XX                                      (    32377. MM BTU)         XX
XX                                                                XX
XX        UNIT GAS COST = $    4.64/GJ                              XX
XX                       ($    4.90/MM BTU)                         XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.8

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                      .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                    .700
        VOLATILE SOLIDS FRACTION IN SOLIDS         .500
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .220

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER
        EFFICIENCY OF ASH REMOVAL                 1.000
        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT         .5405E+04 KG/HR ( .5952E+01 TONS/HR)
        POWER REQUIREMENT                          .2000E+01 HP

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT         .5206E+03 CU.M. ( .1838E+05 CU FT.)
        POWER REQUIREMENT                          .3676E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT         .1599E+04 CU.M. ( .5648E+05 CU.FT.)
        DIGESTER TEMPERATURE                    60.C
        RETENTION TIME                        6.1DAYS
        CONVERSION EFFICIENCY                      .800
        POWER REQUIREMENT                          .1130E+02HP

    GAS PRODUCTION                                3.490 VVD
                                                .2080E+03 GJ/D
                                              ( .1971E+03 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                           1.
        HEAT EXCHANGER AREA       .2563E+03 SQ.M. (  .2757E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT        .5010E+01 GJ/D ( .4749E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	2240.
STORAGE TANKS	103697.
DIGESTERS	274131.
HEAT EXCHANGERS	16455.
MATERIALS HANDLING EQUIPMENT	19826.
SUBTOTAL	416348.
STRUCTURES-ELECT. + PIPING	16654.
SUBTOTAL	433002.
SUPPORT FACILITIES	21650.
SUBTOTAL CAPITAL INVESTMENT	454652.
CONTRACTOR OVERHEAD + PROFIT	45465.
ENGINEERING + DESIGN	22733.
SUBTOTAL PLANT INVESTMENT	522850.
PROJECT CONTINGENCY	78427.
TOTAL PLANT INVESTMENT	601277.
INTEREST DURING CONSTRUCTION	54115.
START-UP COSTS	24964.
WORKING CAPITAL	12026.
TOTAL CAPITAL REQUIREMENT	692382.

```
                        ANNUAL OPERATING COSTS

                                                    ANNUAL COST ($)


    PRODUCTION MATERIALS

        FEED - $   0.00/KG ($   0.00/TON)                         0.

    UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                              2993.

        STEAM -  $ 3.79/GJ
              ($ 4.00/MM BTU)                                  4992.

    LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00                35040.

        MAINTENANCE LABOR                                      9019.

        SUPERVISION                                            6609.

    ADMINISTRATION + OVERHEAD                                 30401.

    SUPPLIES

        OPERATING                                             10512.

        MAINTENANCE                                            9019.

    LOCAL TAXES + INSURANCE                                   16234.
                                                    --------------------
 GROSS ANNUAL OPERATING COST                                 124819.

    CREDITS/PENALTIES COST                                        0.
                                                    --------------------
 NET ANNUAL OPERATING COST                                   124819.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                    XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      208010.          XX
XX                                                                    XX
XX         TOTAL ANNUAL GAS PRODUCTION =      68315. GJ               XX
XX                                      (     64754. MM BTU)          XX
XX                                                                    XX
XX         UNIT GAS COST = $    3.04/GJ                               XX
XX                       ($    3.21/MM BTU)                           XX
XX                                                                    XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.9

```
                         SYSTEM PERFORMANCE

 1. FEED MAKEUP

         TOTAL SOLIDS                        .4540E+06 KG/D ( .1000E+07 LB/D)
         SOLIDS FRACTION IN FEED                       .700
         VOLATILE SOLIDS FRACTION IN SOLIDS            .500
         BIODEGR. VOL. SOL. FRACT. IN SOLIDS           .220

 2. PRETREATMENT

     SHREDDING-NONE USED

     DEGRITTER
         EFFICIENCY OF ASH REMOVAL                    1.000
         NUMBER OF UNITS                              1.
         CAPACITY PER UNIT      .2702E+05 KG/HR ( .2976E+02 TONS/HR)
         POWER REQUIREMENT                             .2000E+01 HP

     CHEMICAL PRETREATMENT-NONE USED

 3. STORAGE-WITH MIXING

         NUMBER OF UNITS                              1.
         CAPACITY PER UNIT      .2603E+04 CU.M. ( .9191E+05 CU FT.)
         POWER REQUIREMENT                             .1838E+02 HP

 4. DIGESTION

     CSTR
         NUMBER OF UNITS                              1.
         CAPACITY PER UNIT      .7997E+04 CU.M. ( .2824E+06 CU.FT.)
         DIGESTER TEMPERATURE                        60.C
         RETENTION TIME                         6.1DAYS
         CONVERSION EFFICIENCY                         .800
         POWER REQUIREMENT                             .5648E+02HP

     GAS PRODUCTION                                   3.490 VVD
                                                   .1040E+04 GJ/D
                                                ( .9856E+03 MM BTU/D)
     HEAT EXCHANGER
         NUMBER OF UNITS                              2.
         HEAT EXCHANGER AREA    .6407E+03 SQ.M. (  .6893E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT     .9473E+01 GJ/D ( .8980E+01 MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	3934.
STORAGE TANKS	239458.
DIGESTERS	916611.
HEAT EXCHANGERS	65432.
MATERIALS HANDLING EQUIPMENT	61272.
SUBTOTAL	1286706.
STRUCTURES-ELECT. + PIPING	51468.
SUBTOTAL	1338174.
SUPPORT FACILITIES	66909.
SUBTOTAL CAPITAL INVESTMENT	1405083.
CONTRACTOR OVERHEAD + PROFIT	140508.
ENGINEERING + DESIGN	70254.
SUBTOTAL PLANT INVESTMENT	1615845.
PROJECT CONTINGENCY	242377.
TOTAL PLANT INVESTMENT	1858222.
INTEREST DURING CONSTRUCTION	167240.
START-UP COSTS	45462.
WORKING CAPITAL	37164.
TOTAL CAPITAL REQUIREMENT	2108088.

```
                    ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


      PRODUCTION MATERIALS

           FEED - $    0.00/KG ($    0.00/TON)                   0.

      UTILITIES

           ELECTRIC - 3.0 CTS./KW-HR                         13552.

           STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                              9439.

      LABOR

           OPERATING LABOR - MEN PER SHIFT     2.
                             HRS PER SHIFT     8.0
                             HOURLY RATE    $ 6.00            35040.

           MAINTENANCE LABOR                                 27873.

           SUPERVISION                                        9437.

      ADMINISTRATION + OVERHEAD                              43410.

      SUPPLIES

           OPERATING                                         10512.

           MAINTENANCE                                       27873.

      LOCAL TAXES + INSURANCE                                50172.
                                                   --------------------
   GROSS ANNUAL OPERATING COST                              227310.

      CREDITS/PENALTIES COST                                     0.
                                                   --------------------
   NET ANNUAL OPERATING COST                                227310.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX      TOTAL ANNUAL CAPITAL REQUIREMENT = $      480611.           XX
XX                                                                  XX
XX      TOTAL ANNUAL GAS PRODUCTION =      341577. GJ               XX
XX                                    (    323770. MM BTU)          XX
XX                                                                  XX
XX      UNIT GAS COST = $    1.41/GJ                                XX
XX                     ($    1.48/MM BTU)                           XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.10

```
                      SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                    .9080E+06 KG/D ( .2000E+07 LB/D)
        SOLIDS FRACTION IN FEED                 .700
        VOLATILE SOLIDS FRACTION IN SOLIDS      .500
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS     .220

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER
        EFFICIENCY OF ASH REMOVAL              1.000
        NUMBER OF UNITS                        1.
        CAPACITY PER UNIT       .5405E+05 KG/HR ( .5952E+02 TONS/HR)
        POWER REQUIREMENT                       .2000E+01 HP

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                        1.
        CAPACITY PER UNIT       .5206E+04 CU.M. ( .1838E+06 CU FT.)
        POWER REQUIREMENT                       .3676E+02 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                        1.
        CAPACITY PER UNIT       .1599E+05 CU.M. ( .5648E+06 CU.FT.)
        DIGESTER TEMPERATURE                 60.C
        RETENTION TIME                    6.1DAYS
        CONVERSION EFFICIENCY                   .800
        POWER REQUIREMENT                       .1130E+03HP

    GAS PRODUCTION                             3.490 VVD
                                             .2080E+04 GJ/D
                                           ( .1971E+04 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                        3.
        HEAT EXCHANGER AREA     .8542E+03 SQ.M. (  .9191E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT      .9895E+01 GJ/D ( .9379E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	5014.
STORAGE TANKS	343371.
DIGESTERS	1541550.
HEAT EXCHANGERS	121782.
MATERIALS HANDLING EQUIPMENT	100586.
SUBTOTAL	2112303.
STRUCTURES-ELECT. + PIPING	84492.
SUBTOTAL	2196795.
SUPPORT FACILITIES	109840.
SUBTOTAL CAPITAL INVESTMENT	2306635.
CONTRACTOR OVERHEAD + PROFIT	230664.
ENGINEERING + DESIGN	115332.
SUBTOTAL PLANT INVESTMENT	2652630.
PROJECT CONTINGENCY	397895.
TOTAL PLANT INVESTMENT	3050525.
INTEREST DURING CONSTRUCTION	274547.
START-UP COSTS	64783.
WORKING CAPITAL	61011.
TOTAL CAPITAL REQUIREMENT	3450865.

```
                     ANNUAL OPERATING COSTS
                                                  ANNUAL COST ($)


   PRODUCTION MATERIALS

        FEED - $   0.00/KG ($   0.00/TON)                          0.

   UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                              26752.

        STEAM -   $ 3.79/GJ
               ($ 4.00/MM BTU)                                  9859.

   LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00                 35040.

        MAINTENANCE LABOR                                      45758.

        SUPERVISION                                            12120.

   ADMINISTRATION + OVERHEAD                                   55751.

   SUPPLIES

        OPERATING                                              10512.

        MAINTENANCE                                            45758.

   LOCAL TAXES + INSURANCE                                     82364.
                                                    --------------------
 GROSS ANNUAL OPERATING COST                                  323914.

   CREDITS/PENALTIES COST                                          0.
                                                    --------------------
 NET ANNUAL OPERATING COST                                    323914.
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      738562.          XX
XX                                                                  XX
XX        TOTAL ANNUAL GAS PRODUCTION =      683154. GJ             XX
XX                                      (    647539. MM BTU)        XX
XX                                                                  XX
XX        UNIT GAS COST = $    1.08/GJ                              XX
XX                       ($    1.14/MM BTU)                         XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.11

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                      .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                      .700
        VOLATILE SOLIDS FRACTION IN SOLIDS           .500
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS          .220

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER
        EFFICIENCY OF ASH REMOVAL                   1.000
        NUMBER OF UNITS                             1.
        CAPACITY PER UNIT           .5405E+04 KG/HR ( .5952E+01 TONS/HR)
        POWER REQUIREMENT                            .2000E+01 HP

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                             1.
        CAPACITY PER UNIT           .2603E+03 CU.M. ( .9191E+04 CU FT.)
        POWER REQUIREMENT                            .1838E+01 HP

4. DIGESTION

    PLUG FLOW
        NUMBER OF UNITS                             6.
        CAPACITY PER UNIT           .1093E+04 CU.M. ( .3861E+05 CU.FT.)
        DIGESTER TEMPERATURE                       37.C
        RETENTION TIME                      25.2DAYS
        CONVERSION EFFICIENCY                        .950
        POWER REQUIREMENT                           0.          HP

    GAS PRODUCTION                                  1.011 VVD
                                                  .2469E+03 GJ/D
                                               ( .2341E+03 MM BTU/D)

    HEAT EXCHANGER
        NUMBER OF UNITS                             1.
        HEAT EXCHANGER AREA         .1253E+03 SQ.M. (  .1348E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT          .1066E+02 GJ/D ( .1011E+02 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	2240.
STORAGE TANKS	72315.
DIGESTERS	85806.
HEAT EXCHANGERS	9621.
MATERIALS HANDLING EQUIPMENT	8499.
SUBTOTAL	178481.
STRUCTURES-ELECT. + PIPING	7139.
SUBTOTAL	185620.
SUPPORT FACILITIES	9281.
SUBTOTAL CAPITAL INVESTMENT	194901.
CONTRACTOR OVERHEAD + PROFIT	19490.
ENGINEERING + DESIGN	9745.
SUBTOTAL PLANT INVESTMENT	224136.
PROJECT CONTINGENCY	33620.
TOTAL PLANT INVESTMENT	257757.
INTEREST DURING CONSTRUCTION	23198.
START-UP COSTS	20845.
WORKING CAPITAL	5155.
TOTAL CAPITAL REQUIREMENT	306955.

```
                      ANNUAL OPERATING COSTS
                                                     ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                          0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                  677.

          STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                   10624.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE    $ 6.00                  35040.

          MAINTENANCE LABOR                                         3866.

          SUPERVISION                                               5836.

     ADMINISTRATION + OVERHEAD                                     26845.

     SUPPLIES

          OPERATING                                                10512.

          MAINTENANCE                                               3866.

     LOCAL TAXES + INSURANCE                                        6959.
                                                       --------------------
  GROSS ANNUAL OPERATING COST                                     104226.

     CREDITS/PENALTIES COST                                            0.
                                                       --------------------
  NET ANNUAL OPERATING COST                                       104226.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                XX
XX      TOTAL ANNUAL CAPITAL REQUIREMENT = $     141104.  XX
XX                                                XX
XX      TOTAL ANNUAL GAS PRODUCTION =      81125. GJ      XX
XX                                  (      76895. MM BTU) XX
XX                                                XX
XX      UNIT GAS COST = $    1.74/GJ                  XX
XX                     ($    1.84/MM BTU)             XX
XX                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.12

```
                        SYSTEM PERFORMANCE

1. FEED MAKEUP

          TOTAL SOLIDS                       .9080E+05 KG/D ( .2000E+06 LB/D)
          SOLIDS FRACTION IN FEED                      .700
          VOLATILE SOLIDS FRACTION IN SOLIDS           .500
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS          .220

2. PRETREATMENT

     SHREDDING-NONE USED

     DEGRITTER
          EFFICIENCY OF ASH REMOVAL                   1.000
          NUMBER OF UNITS                             1.
          CAPACITY PER UNIT          .5400E+04 KG/HR ( .5952E+01 TONS/HR)
          POWER REQUIREMENT                            .2000E+01 HP

     CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

          NUMBER OF UNITS                             1.
          CAPACITY PER UNIT          .2603E+03 CU.M. ( .9191E+04 CU FT.)
          POWER REQUIREMENT                            .1838E+01 HP

4. DIGESTION

     PLUG FLOW
          NUMBER OF UNITS                             2.
          CAPACITY PER UNIT          .5989E+03 CU.M. ( .2115E+05 CU.FT.)
          DIGESTER TEMPERATURE                      60.0
          RETENTION TIME                         4.6DAYS
          CONVERSION EFFICIENCY                        .950
          POWER REQUIREMENT                           0.           HP

     GAS PRODUCTION                                5.534 VVD
                                                .2469E+03 GJ/D
                                              ( .2341E+03 MM BTU/D)

     HEAT EXCHANGER
          NUMBER OF UNITS                             1.
          HEAT EXCHANGER AREA        .2563E+03 SQ.M. (  .2757E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT          .3768E+01 GJ/D ( .3598E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DEGRITTERS	2240.
STORAGE TANKS	72315.
DIGESTERS	15668.
HEAT EXCHANGERS	16455.
MATERIALS HANDLING EQUIPMENT	5334.
SUBTOTAL	112012.
STRUCTURES-ELECT. + PIPING	4480.
SUBTOTAL	116493.
SUPPORT FACILITIES	5825.
SUBTOTAL CAPITAL INVESTMENT	122317.
CONTRACTOR OVERHEAD + PROFIT	12232.
ENGINEERING + DESIGN	6116.
SUBTOTAL PLANT INVESTMENT	140665.
PROJECT CONTINGENCY	21100.
TOTAL PLANT INVESTMENT	161764.
INTEREST DURING CONSTRUCTION	14559.
START-UP COSTS	18139.
WORKING CAPITAL	3235.
TOTAL CAPITAL REQUIREMENT	197698.

```
                       ANNUAL OPERATING COSTS

                                                      ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                         0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                 677.

          STEAM -   $ 3.79/GJ
                   ($ 4.00/MM BTU)                                 3774.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                  35040.

          MAINTENANCE LABOR                                        2426.

          SUPERVISION                                              5620.

     ADMINISTRATION + OVERHEAD                                    25852.

     SUPPLIES

          OPERATING                                               10512.

          MAINTENANCE                                              2426.

     LOCAL TAXES + INSURANCE                                       4368.
                                                        -------------------
 GROSS ANNUAL OPERATING COST                                      90696.

     CREDITS/PENALTIES COST                                           0.
                                                        -------------------
 NET ANNUAL OPERATING COST                                        90696.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      114445.        XX
XX                                                                  XX
XX         TOTAL ANNUAL GAS PRODUCTION =       81125. GJ            XX
XX                                       (     76895. MM BTU)       XX
XX                                                                  XX
XX         UNIT GAS COST = $    1.41/GJ                             XX
XX                        ($    1.49/MM BTU)                        XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX)XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.13

```
                         SYSTEM PERFORMANCE

  1. FEED MAKEUP

          TOTAL SOLIDS                      .3314E+06 KG/D ( .7300E+06 LB/D)
          SOLIDS FRACTION IN FEED                     .700
          VOLATILE SOLIDS FRACTION IN SOLIDS          .500
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS         .220

  2. PRETREATMENT

      SHREDDING-NONE USED

      DEGRITTER-NONE USED

      CHEMICAL PRETREATMENT-NONE USED

  3. STORAGE-NONE USED

  4. DIGESTION

      BATCH
          NUMBER OF UNITS                            1.
          CAPACITY PER UNIT       .1654E+06 CU.M. ( .5840E+07 CU.FT.)
          DIGESTER TEMPERATURE                     37.C
          RETENTION TIME                     25.2DAYS
          CONVERSION EFFICIENCY                       .950
          POWER REQUIREMENT                          0.          HP

      GAS PRODUCTION                                  .146 VVD
                                                  .9014E+03 GJ/D
                                                ( .8544E+03 MM BTU/D)

      TOTAL HEAT REQUIREMENT     0.          GJ/D (0.          MM BTU/D)

   5. DEWATERING-NONE USED

  6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
DIGESTERS	865001.
MATERIALS HANDLING EQUIPMENT	43250.
SUBTOTAL	908251.
STRUCTURES-ELECT. + PIPING	36330.
SUBTOTAL	944582.
SUPPORT FACILITIES	47229.
SUBTOTAL CAPITAL INVESTMENT	991811.
CONTRACTOR OVERHEAD + PROFIT	99181.
ENGINEERING + DESIGN	49591.
SUBTOTAL PLANT INVESTMENT	1140582.
PROJECT CONTINGENCY	171087.
TOTAL PLANT INVESTMENT	1311670.
INTEREST DURING CONSTRUCTION	118050.
START-UP COSTS	33256.
WORKING CAPITAL	26233.
TOTAL CAPITAL REQUIREMENT	1489209.

```
                         ANNUAL OPERATING COSTS

                                                  ANNUAL COST ($)


    PRODUCTION MATERIALS

         FEED - $    0.00/KG ($    0.00/TON)                          0.

    UTILITIES

         ELECTRIC - 3.0 CTS./KW-HR                                    0.

         STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                      0.

    LABOR

         OPERATING LABOR - MEN PER SHIFT    2.
                           HRS PER SHIFT    8.0
                           HOURLY RATE   $ 6.00                   35040.

         MAINTENANCE LABOR                                        19675.

         SUPERVISION                                               8207.

    ADMINISTRATION + OVERHEAD                                     37753.

    SUPPLIES

         OPERATING                                                10512.

         MAINTENANCE                                              19675.

    LOCAL TAXES + INSURANCE                                       35415.
                                                        --------------------
GROSS ANNUAL OPERATING COST                                      166278.

    CREDITS/PENALTIES COST                                            0.
                                                        --------------------
NET ANNUAL OPERATING COST                                        166278.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $     345216.         XX
XX                                                                XX
XX       TOTAL ANNUAL GAS PRODUCTION =      81128. GJ             XX
XX                                   (     76899. MM BTU)         XX
XX                                                                XX
XX       UNIT GAS COST = $    4.26/GJ                             XX
XX                     ($    4.49/MM BTU)                         XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.14

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

          TOTAL SOLIDS                      .9080E+05 KG/D ( .2000E+06 LB/D)
          SOLIDS FRACTION IN FEED                     .900
          VOLATILE SOLIDS FRACTION IN SOLIDS          .825
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS         .400

2. PRETREATMENT

     SHREDDING
          NUMBER OF UNITS                           1.
          CAPACITY PER UNIT       .4204E+04 KG/HR ( .4630E+01 TONS/HR)
          POWER REQUIREMENT                          .6944E+02 HP

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

          NUMBER OF UNITS                           1.
          CAPACITY PER UNIT       .1041E+04 CU.M. ( .3676E+05 CU FT.)
          POWER REQUIREMENT                          .7353E+01 HP

4. DIGESTION

     CSTR
          NUMBER OF UNITS                           1.
          CAPACITY PER UNIT       .8134E+04 CU.M. ( .2872E+06 CU.FT.)
          DIGESTER TEMPERATURE                     37.C
          RETENTION TIME                       15.6DAYS
          CONVERSION EFFICIENCY                      .650
          POWER REQUIREMENT                          .5744E+02HP

     GAS PRODUCTION                                1.014 VVD
                                                .3072E+03 GJ/D
                                             ( .2912E+03 MM BTU/D)

     HEAT EXCHANGER
          NUMBER OF UNITS                           1.
          HEAT EXCHANGER AREA     .2506E+03 SQ.M. (  .2696E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT       .6002E+01 GJ/D ( .5689E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
STORAGE TANKS	148696.
DIGESTERS	928304.
HEAT EXCHANGERS	16180.
MATERIALS HANDLING EQUIPMENT	58510.
SUBTOTAL	1228704.
STRUCTURES-ELECT. + PIPING	49148.
SUBTOTAL	1277853.
SUPPORT FACILITIES	63893.
SUBTOTAL CAPITAL INVESTMENT	1341745.
CONTRACTOR OVERHEAD + PROFIT	134175.
ENGINEERING + DESIGN	67087.
SUBTOTAL PLANT INVESTMENT	1543007.
PROJECT CONTINGENCY	231451.
TOTAL PLANT INVESTMENT	1774458.
INTEREST DURING CONSTRUCTION	159701.
START-UP COSTS	45628.
WORKING CAPITAL	35489.
TOTAL CAPITAL REQUIREMENT	2015276.

```
                      ANNUAL OPERATING COSTS

                                                ANNUAL COST ($)

    PRODUCTION MATERIALS

         FEED - $    0.00/KG ($    0.00/TON)                     0.

    UTILITIES

         ELECTRIC - 3.0 CTS./KW-HR                           23670.

         STEAM -   $ 3.79/GJ
                  ($ 4.00/MM BTU)                             5980.

    LABOR

         OPERATING LABOR - MEN PER SHIFT    2.
                           HRS PER SHIFT    8.0
                           HOURLY RATE   $ 6.00              35040.

         MAINTENANCE LABOR                                   26617.

         SUPERVISION                                          9249.

    ADMINISTRATION + OVERHEAD                                42543.

    SUPPLIES

         OPERATING                                           10512.

         MAINTENANCE                                         26617.

    LOCAL TAXES + INSURANCE                                  47910.
                                                   ---------------------
 GROSS ANNUAL OPERATING COST                                228138.

    CREDITS/PENALTIES COST                                       0.
                                                   ---------------------
 NET ANNUAL OPERATING COST                                  228138.
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $     470286.         XX
XX                                                                XX
XX       TOTAL ANNUAL GAS PRODUCTION =     100920. GJ             XX
XX                                    (     95659. MM BTU)        XX
XX                                                                XX
XX       UNIT GAS COST = $    4.66/GJ                             XX
XX                      ($    4.92/MM BTU)                        XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.15

```
                         SYSTEM PERFORMANCE

 1. FEED MAKEUP

        TOTAL SOLIDS                   .4540E+05 KG/D ( .1000E+06 LB/D)
        SOLIDS FRACTION IN FEED                  .900
        VOLATILE SOLIDS FRACTION IN SOLIDS       .825
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS      .400

 2. PRETREATMENT

    SHREDDING
        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT           .2102E+04 KG/HR ( .2315E+01 TONS/HR)
        POWER REQUIREMENT                        .3472E+02 HP

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

 3. STORAGE-WITH MIXING

        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT           .5206E+03 CU.M. ( .1838E+05 CU FT.)
        POWER REQUIREMENT                        .3676E+01 HP

 4. DIGESTION

    CSTR
        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT           .7426E+03 CU.M. ( .2622E+05 CU.FT.)
        DIGESTER TEMPERATURE                   60.C
        RETENTION TIME                      2.9DAYS
        CONVERSION EFFICIENCY                    .650
        POWER REQUIREMENT                        .5244E+01HP

    GAS PRODUCTION                              5.553 VVD
                                             .1536E+03 GJ/D
                                          ( .1456E+03 MM BTU/D)

    HEAT EXCHANGER
        NUMBER OF UNITS                         1.
        HEAT EXCHANGER AREA         .2563E+03 SQ.M. (  .2757E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT          .4117E+01 GJ/D ( .3903E+01 MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	38507.
STORAGE TANKS	103697.
DIGESTERS	154187.
HEAT EXCHANGERS	16455.
MATERIALS HANDLING EQUIPMENT	15642.
SUBTOTAL	328488.
STRUCTURES-ELECT. + PIPING	13140.
SUBTOTAL	341627.
SUPPORT FACILITIES	17081.
SUBTOTAL CAPITAL INVESTMENT	358709.
CONTRACTOR OVERHEAD + PROFIT	35871.
ENGINEERING + DESIGN	17935.
SUBTOTAL PLANT INVESTMENT	412515.
PROJECT CONTINGENCY	61877.
TOTAL PLANT INVESTMENT	474392.
INTEREST DURING CONSTRUCTION	42695.
START-UP COSTS	23960.
WORKING CAPITAL	9488.
TOTAL CAPITAL REQUIREMENT	550536.

```
                         ANNUAL OPERATING COSTS
                                                        ANNUAL COST ($)

   PRODUCTION MATERIALS

        FEED - $    0.00/KG ($    0.00/TON)                          0.

   UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                                 7695.

        STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                  4103.

   LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00                   35040.

        MAINTENANCE LABOR                                         7116.

        SUPERVISION                                               6323.

   ADMINISTRATION + OVERHEAD                                     29088.

   SUPPLIES

        OPERATING                                                10512.

        MAINTENANCE                                               7116.

   LOCAL TAXES + INSURANCE                                       12809.
                                                     --------------------
GROSS ANNUAL OPERATING COST                                     119801.

   CREDITS/PENALTIES COST                                            0.
                                                     --------------------
NET ANNUAL OPERATING COST                                       119801.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      185948.          XX
XX                                                                  XX
XX        TOTAL ANNUAL GAS PRODUCTION =       50460. GJ              XX
XX                                    (       47830. MM BTU)         XX
XX                                                                  XX
XX        UNIT GAS COST = $    3.69/GJ                               XX
XX                       ($    3.89/MM BTU)                          XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.16

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

     TOTAL SOLIDS                     .9080E+05 KG/D ( .2000E+06 LB/D)
     SOLIDS FRACTION IN FEED                    .900
     VOLATILE SOLIDS FRACTION IN SOLIDS         .825
     BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .400

2. PRETREATMENT

   SHREDDING
     NUMBER OF UNITS                           1.
     CAPACITY PER UNIT       .4204E+04 KG/HR ( .4630E+01 TONS/HR)
     POWER REQUIREMENT                          .6944E+02 HP

   DEGRITTER-NONE USED

   CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

     NUMBER OF UNITS                           1.
     CAPACITY PER UNIT       .1041E+04 CU.M. ( .3676E+05 CU FT.)
     POWER REQUIREMENT                          .7353E+01 HP

4. DIGESTION

   CSTR
     NUMBER OF UNITS                           1.
     CAPACITY PER UNIT       .1485E+04 CU.M. ( .5244E+05 CU.FT.)
     DIGESTER TEMPERATURE                    60.C
     RETENTION TIME                      2.9DAYS
     CONVERSION EFFICIENCY                      .650
     POWER REQUIREMENT                          .1049E+02HP

   GAS PRODUCTION                              5.553  VVD
                                            .3072E+03  GJ/D
                                         ( .2912E+03  MM BTU/D)
   HEAT EXCHANGER
     NUMBER OF UNITS                           1.
     HEAT EXCHANGER AREA     .5125E+03 SQ.M. (  .5515E+04 SQ.FT.)

   TOTAL HEAT REQUIREMENT    .6379E+01 GJ/D ( .6046E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
STORAGE TANKS	148696.
DIGESTERS	259310.
HEAT EXCHANGERS	27674.
MATERIALS HANDLING EQUIPMENT	25635.
SUBTOTAL	538330.
STRUCTURES-ELECT. + PIPING	21533.
SUBTOTAL	559863.
SUPPORT FACILITIES	27993.
SUBTOTAL CAPITAL INVESTMENT	587856.
CONTRACTOR OVERHEAD + PROFIT	58786.
ENGINEERING + DESIGN	29393.
SUBTOTAL PLANT INVESTMENT	676034.
PROJECT CONTINGENCY	101405.
TOTAL PLANT INVESTMENT	777439.
INTEREST DURING CONSTRUCTION	69970.
START-UP COSTS	30168.
WORKING CAPITAL	15549.
TOTAL CAPITAL REQUIREMENT	893126.

```
                         ANNUAL OPERATING COSTS

                                                     ANNUAL COST ($)


    PRODUCTION MATERIALS

         FEED - $    0.00/KG ($    0.00/TON)                         0.

    UTILITIES

         ELECTRIC - 3.0 CTS./KW-HR                               15391.

         STEAM -   $ 3.79/GJ
                  ($ 4.00/MM BTU)                                 6356.

    LABOR

         OPERATING LABOR - MEN PER SHIFT    2.
                           HRS PER SHIFT    8.0
                           HOURLY RATE   $ 6.00                  35040.

         MAINTENANCE LABOR                                       11662.

         SUPERVISION                                              7005.

    ADMINISTRATION + OVERHEAD                                    32224.

    SUPPLIES

         OPERATING                                               10512.

         MAINTENANCE                                             11662.

    LOCAL TAXES + INSURANCE                                      20991.
                                                    --------------------
 GROSS ANNUAL OPERATING COST                                    150842.

    CREDITS/PENALTIES COST                                           0.
                                                    --------------------
 NET ANNUAL OPERATING COST                                      150842.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      258154.         XX
XX                                                                  XX
XX        TOTAL ANNUAL GAS PRODUCTION =      100920. GJ             XX
XX                                      (     95659. MM BTU)        XX
XX                                                                  XX
XX        UNIT GAS COST = $    2.56/GJ                              XX
XX                       ($    2.70/MM BTU)                         XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.17

```
                          SYSTEM PERFORMANCE

 1. FEED MAKEUP

        TOTAL SOLIDS                    .4540E+06 KG/D ( .1000E+07 LB/D)
        SOLIDS FRACTION IN FEED                   .900
        VOLATILE SOLIDS FRACTION IN SOLIDS        .825
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS       .400

 2. PRETREATMENT

     SHREDDING
        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT           .2102E+05 KG/HR ( .2315E+02 TONS/HR)
        POWER REQUIREMENT                         .3472E+03 HP

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT-NONE USED

 3. STORAGE-WITH MIXING

        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT           .5206E+04 CU.M. ( .1838E+06 CU FT.)
        POWER REQUIREMENT                         .3676E+02 HP

 4. DIGESTION

     CSTR
        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT           .7426E+04 CU.M. ( .2622E+06 CU.FT.)
        DIGESTER TEMPERATURE                   60.C
        RETENTION TIME                     2.9DAYS
        CONVERSION EFFICIENCY                     .650
        POWER REQUIREMENT                         .5244E+02HP

     GAS PRODUCTION                              5.553 VVD
                                              .1536E+04 GJ/D
                                           ( .1456E+04 MM BTU/D)

     HEAT EXCHANGER
        NUMBER OF UNITS                          3.
        HEAT EXCHANGER AREA         .8542E+03 SQ.M. (  .9191E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT         .1707E+02 GJ/D ( .1618E+02 MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	385072.
STORAGE TANKS	343371.
DIGESTERS	867055.
HEAT EXCHANGERS	121782.
MATERIALS HANDLING EQUIPMENT	85864.
SUBTOTAL	1803145.
STRUCTURES-ELECT. + PIPING	72126.
SUBTOTAL	1875271.
SUPPORT FACILITIES	93764.
SUBTOTAL CAPITAL INVESTMENT	1969034.
CONTRACTOR OVERHEAD + PROFIT	196903.
ENGINEERING + DESIGN	98452.
SUBTOTAL PLANT INVESTMENT	2264389.
PROJECT CONTINGENCY	339658.
TOTAL PLANT INVESTMENT	2604048.
INTEREST DURING CONSTRUCTION	234364.
START-UP COSTS	70038.
WORKING CAPITAL	52081.
TOTAL CAPITAL REQUIREMENT	2960531.

```
                         ANNUAL OPERATING COSTS

                                                      ANNUAL COST ($)


    PRODUCTION MATERIALS

        FEED - $   0.00/KG ($   0.00/TON)                            0.

    UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                                76955.

        STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                 17007.

    LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE    $ 6.00                  35040.

        MAINTENANCE LABOR                                        39061.

        SUPERVISION                                              11115.

    ADMINISTRATION + OVERHEAD                                    51129.

    SUPPLIES

        OPERATING                                                10512.

        MAINTENANCE                                              39061.

    LOCAL TAXES + INSURANCE                                      70309.
                                                    --------------------
 GROSS ANNUAL OPERATING COST                                    350189.

    CREDITS/PENALTIES COST                                           0.
                                                    --------------------
 NET ANNUAL OPERATING COST                                      350189.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                    XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $     705915.             XX
XX                                                                    XX
XX       TOTAL ANNUAL GAS PRODUCTION =     504602. GJ                 XX
XX                                       (    478296. MM BTU)         XX
XX                                                                    XX
XX       UNIT GAS COST = $    1.40/GJ                                 XX
XX                      ($    1.48/MM BTU)                            XX
XX                                                                    XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.18

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

          TOTAL SOLIDS                       .9080E+06 KG/D ( .2000E+07 LB/D)
          SOLIDS FRACTION IN FEED                   .900
          VOLATILE SOLIDS FRACTION IN SOLIDS        .825
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS       .400

2. PRETREATMENT

     SHREDDING
          NUMBER OF UNITS                          2.
          CAPACITY PER UNIT         .2102E+05 KG/HR ( .2315E+02 TONS/HR)
          POWER REQUIREMENT                     .6944E+03 HP

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT-NONE USED

 . STORAGE-WITH MIXING

          NUMBER OF UNITS                          2.
          CAPACITY PER UNIT         .5206E+04 CU.M. ( .1838E+06 CU FT.)
          POWER REQUIREMENT                     .7353E+02 HP

4. DIGESTION

     CSTR
          NUMBER OF UNITS                          1.
          CAPACITY PER UNIT         .1485E+05 CU.M. ( .5244E+06 CU.FT.)
          DIGESTER TEMPERATURE                 60.C
          RETENTION TIME                     2.9DAYS
          CONVERSION EFFICIENCY                     .650
          POWER REQUIREMENT                         .1049E+03HP

     GAS PRODUCTION                                5.553 VVD
                                               .3072E+04 GJ/D
                                             ( .2912E+04 MM BTU/D)

     HEAT EXCHANGER
          NUMBER OF UNITS                          6.
          HEAT EXCHANGER AREA       .8542E+03 SQ.M. (  .9191E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT         .2552E+02 GJ/D ( .2419E+02 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	770144.
STORAGE TANKS	686743.
DIGESTERS	1458207.
HEAT EXCHANGERS	243565.
MATERIALS HANDLING EQUIPMENT	157933.
SUBTOTAL	3316591.
STRUCTURES-ELECT. + PIPING	132664.
SUBTOTAL	3449255.
SUPPORT FACILITIES	172463.
SUBTOTAL CAPITAL INVESTMENT	3621718.
CONTRACTOR OVERHEAD + PROFIT	362172.
ENGINEERING + DESIGN	181086.
SUBTOTAL PLANT INVESTMENT	4164976.
PROJECT CONTINGENCY	624746.
TOTAL PLANT INVESTMENT	4789722.
INTEREST DURING CONSTRUCTION	431075.
START-UP COSTS	117538.
WORKING CAPITAL	95794.
TOTAL CAPITAL REQUIREMENT	5434129.

```
                    ANNUAL OPERATING COSTS
                                                  ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                        0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                             153910.

          STEAM -   $ 3.79/GJ
                  ($ 4.00/MM BTU)                                25429.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                 35040.

          MAINTENANCE LABOR                                      71846.

          SUPERVISION                                            16033.

      ADMINISTRATION + OVERHEAD                                  73751.

      SUPPLIES

          OPERATING                                              10512.

          MAINTENANCE                                            71846.

      LOCAL TAXES + INSURANCE                                   129322.
                                                      --------------------
   GROSS ANNUAL OPERATING COST                                  587689.

      CREDITS/PENALTIES COST                                         0.
                                                      --------------------
   NET ANNUAL OPERATING COST                                    587689.
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                   XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $    1240637.           XX
XX                                                                   XX
XX        TOTAL ANNUAL GAS PRODUCTION =    1009205. GJ               XX
XX                                    (    956592. MM BTU)           XX
XX                                                                   XX
XX        UNIT GAS COST = $    1.23/GJ                               XX
XX                       ($    1.30/MM BTU)                          XX
XX                                                                   XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.19

```
                              SYSTEM PERFORMANCE

 1. FEED MAKEUP

          TOTAL SOLIDS                        .9080E+05 KG/D ( .2000E+06 LB/D)
          SOLIDS FRACTION IN FEED                        .900
          VOLATILE SOLIDS FRACTION IN SOLIDS             .825
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS            .400

 2. PRETREATMENT

      SHREDDING
          NUMBER OF UNITS                            1.
          CAPACITY PER UNIT           .4204E+04 KG/HR ( .4630E+01 TONS/HR)
          POWER REQUIREMENT                              .6944E+02 HP

      DEGRITTER-NONE USED

      CHEMICAL PRETREATMENT
          RATIO OF CHEMICAL TO FEED                      .200
          CONVERSION EFFICIENCY                          .500
          NUMBER OF UNITS                            1.
          CAPACITY PER UNIT           .8194E+01 CU.M. ( .2893E+03 CU.FT.)
          CHEMICAL REQUIREMENT        .3859E+04 KG/HR ( .4250E+01 TONS/HR)
          POWER REQUIREMENT                          0.          HP

 3. STORAGE-WITH MIXING

          NUMBER OF UNITS                            1.
          CAPACITY PER UNIT           .1085E+04 CU.M. ( .3833E+05 CU FT.)
          POWER REQUIREMENT                              .7665E+01 HP

 4. DIGESTION

      CSTR
          NUMBER OF UNITS                            1.
          CAPACITY PER UNIT           .3335E+04 CU.M. ( .1178E+06 CU.FT.)
          DIGESTER TEMPERATURE                     60.C
          RETENTION TIME                        6.1DAYS
          CONVERSION EFFICIENCY                          .800
          POWER REQUIREMENT                              .2355E+02HP

      GAS PRODUCTION                                 4.661 VVD
                                                 .5790E+03 GJ/D
                                              ( .5488E+03 MM BTU/D)

      HEAT EXCHANGER
          NUMBER OF UNITS                            1.
          HEAT EXCHANGER AREA         .5343E+03 SQ.M. (  .5749E+04 SQ.FT.)

      TOTAL HEAT REQUIREMENT          .1399E+01 GJ/D ( .1326E+01 MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
CHEMICAL PRETREATMENT	15843.
STORAGE TANKS	151950.
DIGESTERS	475650.
HEAT EXCHANGERS	28552.
MATERIALS HANDLING EQUIPMENT	37450.
SUBTOTAL	786459.
STRUCTURES-ELECT. + PIPING	31458.
SUBTOTAL	817917.
SUPPORT FACILITIES	40896.
SUBTOTAL CAPITAL INVESTMENT	858813.
CONTRACTOR OVERHEAD + PROFIT	85881.
ENGINEERING + DESIGN	42941.
SUBTOTAL PLANT INVESTMENT	987635.
PROJECT CONTINGENCY	148145.
TOTAL PLANT INVESTMENT	1135780.
INTEREST DURING CONSTRUCTION	102220.
START-UP COSTS	118403.
WORKING CAPITAL	22716.
TOTAL CAPITAL REQUIREMENT	1379119.

```
                    ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                      0.

          PRETREATMENT CHEMICAL
                $    .31/KG ($ 286.00/TON)                    418837.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                            17749.

          STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                1394.

     LABOR

          OPERATING LABOR - MEN PER SHIFT     2.
                            HRS PER SHIFT     8.0
                            HOURLY RATE    $ 6.00              35040.

          MAINTENANCE LABOR                                    17037.

          SUPERVISION                                           7812.

     ADMINISTRATION + OVERHEAD                                 35933.

     SUPPLIES

          OPERATING                                            10512.

          MAINTENANCE                                          17037.

     LOCAL TAXES + INSURANCE                                   35666.
                                                     --------------------
   GROSS ANNUAL OPERATING COST                                592016.

     CREDITS/PENALTIES COST                                        0.
                                                     --------------------
   NET ANNUAL OPERATING COST                                  592016.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                 XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      757695.        XX
XX                                                                 XX
XX        TOTAL ANNUAL GAS PRODUCTION =      190196. GJ            XX
XX                                   (      180281. MM BTU)        XX
XX                                                                 XX
XX        UNIT GAS COST = $    3.98/GJ                             XX
XX                       ($    4.20/MM BTU)                        XX
XX                                                                 XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.20

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

          TOTAL SOLIDS                       .9083E+05 KG/D ( .2000E+06 LB/D)
          SOLIDS FRACTION IN FEED                      .900
          VOLATILE SOLIDS FRACTION IN SOLIDS           .825
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS          .400

2. PRETREATMENT

     SHREDDING
          NUMBER OF UNITS                             1.
          CAPACITY PER UNIT          .4204E+04 KG/HR ( .4630E+01 TONS/HR)
          POWER REQUIREMENT                            .6944E+02 HP

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT
          RATIO OF CHEMICAL TO FEED                    .200
          CONVERSION EFFICIENCY                        .500
          NUMBER OF UNITS                             1.
          CAPACITY PER UNIT          .8194E+01 CU.M. ( .2893E+03 CU.FT.)
          CHEMICAL REQUIREMENT       .3859E+04 KG/HR ( .4250E+01 TONS/HR)
          POWER REQUIREMENT                           0.          HP

3. STORAGE-WITH MIXING

          NUMBER OF UNITS                             1.
          CAPACITY PER UNIT          .5427E+03 CU.M. ( .1916E+05 CU FT.)
          POWER REQUIREMENT                            .3833E+01 HP

4. DIGESTION

     PLUG FLOW
          NUMBER OF UNITS                             3.
          CAPACITY PER UNIT          .8325E+03 CU.M. ( .2940E+05 CU.FT.)
          DIGESTER TEMPERATURE                      60.C
          RETENTION TIME                        4.6DAYS
          CONVERSION EFFICIENCY                        .950
          POWER REQUIREMENT                           0.         HP

     GAS PRODUCTION                                 7.390 VVD
                                                 .6875E+03 GJ/D
                                             ( .6517E+03 MM BTU/D)

     HEAT EXCHANGER
          NUMBER OF UNITS                             1.
          HEAT EXCHANGER AREA        .5343E+03 SQ.M. (  .5749E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT      0.           GJ/D (0.           MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
CHEMICAL PRETREATMENT	15843.
STORAGE TANKS	105965.
DIGESTERS	32669.
HEAT EXCHANGERS	28552.
MATERIALS HANDLING EQUIPMENT	13002.
SUBTOTAL	273045.
STRUCTURES-ELECT. + PIPING	10922.
SUBTOTAL	283967.
SUPPORT FACILITIES	14198.
SUBTOTAL CAPITAL INVESTMENT	298166.
CONTRACTOR OVERHEAD + PROFIT	29817.
ENGINEERING + DESIGN	14908.
SUBTOTAL PLANT INVESTMENT	342890.
PROJECT CONTINGENCY	51434.
TOTAL PLANT INVESTMENT	394324.
INTEREST DURING CONSTRUCTION	35489.
START-UP COSTS	106838.
WORKING CAPITAL	7886.
TOTAL CAPITAL REQUIREMENT	544537.

```
                    ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


   PRODUCTION MATERIALS

        FEED - $    0.00/KG ($    0.00/TON)                         0.

        PRETREATMENT CHEMICAL
                $     .31/KG ($ 300.00/TON)                    418837.

   UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                               12921.

        STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                    0.

   LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00                  35040.

        MAINTENANCE LABOR                                        5915.

        SUPERVISION                                              6143.

   ADMINISTRATION + OVERHEAD                                    28259.

   SUPPLIES

        OPERATING                                               10512.

        MAINTENANCE                                              5915.

   LOCAL TAXES + INSURANCE                                      15647.
                                                    --------------------
GROSS ANNUAL OPERATING COST                                    534189.

   CREDITS/PENALTIES COST                                           0.
                                                    --------------------
NET ANNUAL OPERATING COST                                      534189.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $      599585.          XX
XX                                                                  XX
XX       TOTAL ANNUAL GAS PRODUCTION =      225858. GJ              XX
XX                                     (    214083. MM BTU)         XX
XX                                                                  XX
XX       UNIT GAS COST = $    2.65/GJ                               XX
XX                      ($    2.80/MM BTU)                          XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.21

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

       TOTAL SOLIDS                         .3314E+06 KG/D ( .7300E+06 LB/D)
       SOLIDS FRACTION IN FEED                        .900
       VOLATILE SOLIDS FRACTION IN SOLIDS             .825
       BIODEGR. VOL. SOL. FRACT. IN SOLIDS            .400

2. PRETREATMENT

     SHREDDING
        NUMBER OF UNITS                              1.
        CAPACITY PER UNIT            .1534E+05 KG/HR ( .1690E+02 TONS/HR)
        POWER REQUIREMENT                             .2535E+03 HP

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                     .200
        CONVERSION EFFICIENCY                         .500
        NUMBER OF UNITS                              1.
        CAPACITY PER UNIT            .4254E+02 CU.M. ( .1502E+04 CU.FT.)
        CHEMICAL REQUIREMENT         .1409E+05 KG/HR ( .1551E+02 TONS/HR)
        POWER REQUIREMENT                            0.          HP

 3. STORAGE-NONE USED

 4. DIGESTION

     BATCH
        NUMBER OF UNITS                              1.
        CAPACITY PER UNIT            .1654E+06 CU.M. ( .5840E+07 CU.FT.)
        DIGESTER TEMPERATURE                        37.C
        RETENTION TIME                          25.2DAYS
        CONVERSION EFFICIENCY                         .950
        POWER REQUIREMENT                            0.          HP

     GAS PRODUCTION                                   .407 VVD
                                                 .2510E+04 GJ/D
                                               ( .2379E+04 MM BTU/D)

     TOTAL HEAT REQUIREMENT                      GJ/D (0.          MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	281103.
CHEMICAL PRETREATMENT	54486.
DIGESTERS	865001.
MATERIALS HANDLING EQUIPMENT	60029.
SUBTOTAL	1260619.
STRUCTURES-ELECT. + PIPING	50425.
SUBTOTAL	1311044.
SUPPORT FACILITIES	65552.
SUBTOTAL CAPITAL INVESTMENT	1376596.
CONTRACTOR OVERHEAD + PROFIT	137660.
ENGINEERING + DESIGN	68830.
SUBTOTAL PLANT INVESTMENT	1583086.
PROJECT CONTINGENCY	237463.
TOTAL PLANT INVESTMENT	1820549.
INTEREST DURING CONSTRUCTION	163849.
START-UP COSTS	126556.
WORKING CAPITAL	36411.
TOTAL CAPITAL REQUIREMENT	2147365.

```
                         ANNUAL OPERATING COSTS

                                                      ANNUAL COST ($)


       PRODUCTION MATERIALS

            FEED - $    0.00/KG ($    0.00/TON)                       0.

            PRETREATMENT CHEMICAL
                   $     .31/KG ($ 300.00/TON)                   418837.

       UTILITIES

            ELECTRIC - 3.0 CTS./KW-HR                              12245.

            STEAM -   $ 3.79/GJ
                   ($ 4.00/MM BTU)                                     0.

       LABOR

            OPERATING LABOR - MEN PER SHIFT    2.
                              HRS PER SHIFT    8.0
                              HOURLY RATE   $ 6.00                 35040.

            MAINTENANCE LABOR                                      27308.

            SUPERVISION                                             9352.

       ADMINISTRATION + OVERHEAD                                   43020.

       SUPPLIES

            OPERATING                                              10512.

            MAINTENANCE                                            27308.

       LOCAL TAXES + INSURANCE                                     49155.
                                                         --------------------
    GROSS ANNUAL OPERATING COST                                   632778.

       CREDITS/PENALTIES COST                                          0.
                                                         --------------------
    NET ANNUAL OPERATING COST                                    632778.

    XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
    XX                                                                 XX
    XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $     890770.         XX
    XX                                                                 XX
    XX        TOTAL ANNUAL GAS PRODUCTION =     225868. GJ             XX
    XX                                   (      214093. MM BTU)        XX
    XX                                                                 XX
    XX        UNIT GAS COST = $    3.94/GJ                             XX
    XX                       ($    4.16/MM BTU)                        XX
    XX                                                                 XX
    XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.22

```
                              SYSTEM PERFORMANCE

 1. FEED MAKEUP

         TOTAL SOLIDS                     .9080E+05 KG/D ( .2000E+06 LB/D)
         SOLIDS FRACTION IN FEED                     .500
         VOLATILE SOLIDS FRACTION IN SOLIDS          .830
         BIODEGR. VOL. SOL. FRACT. IN SOLIDS         .640

 2. PRETREATMENT

      SHREDDING-NONE USED

      DEGRITTER-NONE USED

      CHEMICAL PRETREATMENT-NONE USED

 3. STORAGE-WITH MIXING

         NUMBER OF UNITS                             1.
         CAPACITY PER UNIT          .1041E+04 CU.M. ( .3676E+05 CU FT.)
         POWER REQUIREMENT                           .7353E+01 HP

 4. DIGESTION

      CSTR
         NUMBER OF UNITS                             1.
         CAPACITY PER UNIT          .4380E+04 CU.M. ( .1546E+06 CU.FT.)
         DIGESTER TEMPERATURE                      37.C
         RETENTION TIME                       8.4DAYS
         CONVERSION EFFICIENCY                       .500
         POWER REQUIREMENT                           .3093E+02HP

      GAS PRODUCTION                               2.318 VVD
                                                .3781E+03 GJ/D
                                              ( .3584E+03 MM BTU/D)

      HEAT EXCHANGER
         NUMBER OF UNITS                             1.
         HEAT EXCHANGER AREA        .2506E+03 SQ.M. (  .2696E+04 SQ.FT.)

      TOTAL HEAT REQUIREMENT     0.         GJ/D (0.          MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	148696.
DIGESTERS	583521.
HEAT EXCHANGERS	16180.
MATERIALS HANDLING EQUIPMENT	37420.
SUBTOTAL	785817.
STRUCTURES-ELECT. + PIPING	31433.
SUBTOTAL	817249.
SUPPORT FACILITIES	40862.
SUBTOTAL CAPITAL INVESTMENT	858112.
CONTRACTOR OVERHEAD + PROFIT	85811.
ENGINEERING + DESIGN	42906.
SUBTOTAL PLANT INVESTMENT	986829.
PROJECT CONTINGENCY	148024.
TOTAL PLANT INVESTMENT	1134853.
INTEREST DURING CONSTRUCTION	102137.
START-UP COSTS	32144.
WORKING CAPITAL	22697.
TOTAL CAPITAL REQUIREMENT	1291831.

```
                    ANNUAL OPERATING COSTS
                                                  ANNUAL COST ($)

   PRODUCTION MATERIALS

        FEED - $    0.00/KG ($    0.00/TON)                    0.

   UTILITIES

        ELECTRIC - 3.0 CTS./KW-HR                           6750.

        STEAM -   $ 3.79/GJ                                    0.
                 ($ 4.00/MM BTU)

   LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE   $ 6.00             35040.

        MAINTENANCE LABOR                                  17023.

        SUPERVISION                                         7809.

    ADMINISTRATION + OVERHEAD                              35923.

    SUPPLIES

        OPERATING                                          10512.

        MAINTENANCE                                        17023.

    LOCAL TAXES + INSURANCE                                30641.
                                                 -----------------
 GROSS ANNUAL OPERATING COST                              160722.

    CREDITS/PENALTIES COST                                     0.
                                                 -----------------
 NET ANNUAL OPERATING COST                                160722.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      315942.      XX
XX                                                                XX
XX         TOTAL ANNUAL GAS PRODUCTION =      124210. GJ          XX
XX                                      (     117734. MM BTU)     XX
XX                                                                XX
XX         UNIT GAS COST = $    2.54/GJ                           XX
XX                        ($    2.68/MM BTU)                      XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.23

```
                        SYSTEM PERFORMANCE

1. FEED MAKEUP

      TOTAL SOLIDS                     .4540E+05 KG/D ( .1000E+06 LB/D)
      SOLIDS FRACTION IN FEED                  .500
      VOLATILE SOLIDS FRACTION IN SOLIDS       .830
      BIODEGR. VOL. SOL. FRACT. IN SOLIDS      .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

      NUMBER OF UNITS                          1.
      CAPACITY PER UNIT       .5206E+03 CU.M. ( .1838E+05 CU FT.)
      POWER REQUIREMENT                        .3676E+01 HP

4. DIGESTION

    CSTR
      NUMBER OF UNITS                          1.
      CAPACITY PER UNIT       .7426E+03 CU.M. ( .2622E+05 CU.FT.)
      DIGESTER TEMPERATURE                    60.C
      RETENTION TIME                       2.9DAYS
      CONVERSION EFFICIENCY                    .650
      POWER REQUIREMENT                        .5244E+01HP

    GAS PRODUCTION                             8.884 VVD
                                           .2450E+03 GJ/D
                                         ( .2330E+03 MM BTU/D)

    HEAT EXCHANGER
      NUMBER OF UNITS                          1.
      HEAT EXCHANGER AREA     .2563E+03 SQ.M. (  .2757E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT    .5366E+00 GJ/D ( .5086E+00 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	103697.
DIGESTERS	154187.
HEAT EXCHANGERS	16455.
MATERIALS HANDLING EQUIPMENT	13717.
SUBTOTAL	288055.
STRUCTURES-ELECT. + PIPING	11522.
SUBTOTAL	299577.
SUPPORT FACILITIES	14979.
SUBTOTAL CAPITAL INVESTMENT	314556.
CONTRACTOR OVERHEAD + PROFIT	31456.
ENGINEERING + DESIGN	15728.
SUBTOTAL PLANT INVESTMENT	361740.
PROJECT CONTINGENCY	54261.
TOTAL PLANT INVESTMENT	416001.
INTEREST DURING CONSTRUCTION	37440.
START-UP COSTS	21209.
WORKING CAPITAL	8320.
TOTAL CAPITAL REQUIREMENT	482970.

```
                      ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


    PRODUCTION MATERIALS

        FEED - $   0.00/KG ($   0.00/TON)                         0.

    UTILITIES

        ELECTRIC - 3.8 CTS./KW-HR                              1573.

        STEAM -  $ 3.79/GJ
                ($ 4.00/MM BTU)                                  535.

    LABOR

        OPERATING LABOR - MEN PER SHIFT    2.
                          HRS PER SHIFT    8.0
                          HOURLY RATE    $ 6.00               35040.

        MAINTENANCE LABOR                                      6240.

        SUPERVISION                                            6192.

    ADMINISTRATION + OVERHEAD                                 28483.

    SUPPLIES

        OPERATING                                             10512.

        MAINTENANCE                                            6240.

    LOCAL TAXES + INSURANCE                                   11232.
                                                   --------------------
 GROSS ANNUAL OPERATING COST                                 106047.

    CREDITS/PENALTIES COST                                        0.
                                                   --------------------
 NET ANNUAL OPERATING COST                                   106047.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $      164075.        XX
XX                                                                XX
XX       TOTAL ANNUAL GAS PRODUCTION =       80736. GJ             XX
XX                                      (     76527. MM BTU)       XX
XX                                                                XX
XX       UNIT GAS COST = $    2.03/GJ                              XX
XX                      ($    2.14/MM BTU)                         XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.24

```
                          SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                       .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                      .500
        VOLATILE SOLIDS FRACTION IN SOLIDS           .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS          .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                              1.
        CAPACITY PER UNIT          .1041E+04 CU.M. ( .3676E+05 CU FT.)
        POWER REQUIREMENT                            .7353E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                              1.
        CAPACITY PER UNIT          .1485E+04 CU.M. ( .5244E+05 CU.FT.)
        DIGESTER TEMPERATURE                     60.C
        RETENTION TIME                         2.9DAYS
        CONVERSION EFFICIENCY                        .650
        POWER REQUIREMENT                            .1049E+02HP

    GAS PRODUCTION                                  8.884 VVD
                                                 .4915E+03 GJ/D
                                               ( .4659E+03 MM BTU/D)

    HEAT EXCHANGER
        NUMBER OF UNITS                              1.
        HEAT EXCHANGER AREA        .5125E+03 SQ.M. (  .5515E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT     0.        GJ/D (0.          MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	148696.
DIGESTERS	259310.
HEAT EXCHANGERS	27674.
MATERIALS HANDLING EQUIPMENT	21784.
SUBTOTAL	457464.
STRUCTURES-ELECT. + PIPING	18299.
SUBTOTAL	475763.
SUPPORT FACILITIES	23788.
SUBTOTAL CAPITAL INVESTMENT	499551.
CONTRACTOR OVERHEAD + PROFIT	49955.
ENGINEERING + DESIGN	24978.
SUBTOTAL PLANT INVESTMENT	574484.
PROJECT CONTINGENCY	86173.
TOTAL PLANT INVESTMENT	660656.
INTEREST DURING CONSTRUCTION	59459.
START-UP COSTS	24823.
WORKING CAPITAL	13213.
TOTAL CAPITAL REQUIREMENT	758151.

```
                         ANNUAL OPERATING COSTS

                                                     ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $   0.00/KG ($   0.00/TON)                            0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                 3146.

          STEAM -  $ 3.79/GJ
                   ($ 4.00/MM BTU)                                     0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                   35040.

          MAINTENANCE LABOR                                         9910.

          SUPERVISION                                               6742.

     ADMINISTRATION + OVERHEAD                                     31015.

     SUPPLIES

          OPERATING                                                10512.

          MAINTENANCE                                               9910.

     LOCAL TAXES + INSURANCE                                       17838.
                                                        --------------------
   GROSS ANNUAL OPERATING COST                                    124113.

     CREDITS/PENALTIES COST                                            0.
                                                        --------------------
   NET ANNUAL OPERATING COST                                      124113.

  XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
  XX                                                                   XX
  XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      215207.          XX
  XX                                                                   XX
  XX        TOTAL ANNUAL GAS PRODUCTION =      161473. GJ              XX
  XX                                     (     153055. MM BTU)         XX
  XX                                                                   XX
  XX        UNIT GAS COST = $     1.33/GJ                              XX
  XX                      ($     1.41/MM BTU)                          XX
  XX                                                                   XX
  XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.25

```
                        SYSTEM PERFORMANCE

 1. FEED MAKEUP

        TOTAL SOLIDS                     .4540E+06 KG/D ( .1000E+07 LB/D)
        SOLIDS FRACTION IN FEED                    .500
        VOLATILE SOLIDS FRACTION IN SOLIDS         .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .640

 2. PRETREATMENT

     SHREDDING-NONE USED

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT-NONE USED

 3. STORAGE-WITH MIXING

        NUMBER OF UNITS                            1.
        CAPACITY PER UNIT          .5206E+04 CU.M. ( .1838E+06 CU FT.)
        POWER REQUIREMENT                          .3676E+02 HP

 4. DIGESTION

     CSTR
        NUMBER OF UNITS                            1.
        CAPACITY PER UNIT          .7426E+04 CU.M. ( .2622E+06 CU.FT.)
        DIGESTER TEMPERATURE                     60.C
        RETENTION TIME                       2.9DAYS
        CONVERSION EFFICIENCY                      .650
        POWER REQUIREMENT                          .5244E+02HP

     GAS PRODUCTION                                8.884 VVD
                                               .2458E+04 GJ/D
                                             ( .2330E+04 MM BTU/D)

     HEAT EXCHANGER
        NUMBER OF UNITS                            3.
        HEAT EXCHANGER AREA        .8542E+03 SQ.M. (  .9191E+04 SQ.FT.)

     TOTAL HEAT REQUIREMENT     0.         GJ/D (0.           MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	343371.
DIGESTERS	867055.
HEAT EXCHANGERS	121782.
MATERIALS HANDLING EQUIPMENT	66610.
SUBTOTAL	1398819.
STRUCTURES-ELECT. + PIPING	55953.
SUBTOTAL	1454772.
SUPPORT FACILITIES	72739.
SUBTOTAL CAPITAL INVESTMENT	1527511.
CONTRACTOR OVERHEAD + PROFIT	152751.
ENGINEERING + DESIGN	76376.
SUBTOTAL PLANT INVESTMENT	1756637.
PROJECT CONTINGENCY	263496.
TOTAL PLANT INVESTMENT	2020133.
INTEREST DURING CONSTRUCTION	181812.
START-UP COSTS	46263.
WORKING CAPITAL	40403.
TOTAL CAPITAL REQUIREMENT	2288611.

```
                     ANNUAL OPERATING COSTS

                                                ANNUAL COST ($)

     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                  0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                        15730.

          STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                              0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE    $ 6.00          35040.

          MAINTENANCE LABOR                                30302.

          SUPERVISION                                       9801.

     ADMINISTRATION + OVERHEAD                             45086.

     SUPPLIES

          OPERATING                                        10512.

          MAINTENANCE                                      30302.

     LOCAL TAXES + INSURANCE                               54544.
                                                --------------------
 GROSS ANNUAL OPERATING COST                              231317.

     CREDITS/PENALTIES COST                                    0.
                                                --------------------
 NET ANNUAL OPERATING COST                                231317.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      506310.         XX
XX                                                                  XX
XX         TOTAL ANNUAL GAS PRODUCTION =      807364. GJ             XX
XX                                      (     765274. MM BTU)        XX
XX                                                                  XX
XX         UNIT GAS COST = $      .63/GJ                             XX
XX                       ($      .66/MM BTU)                         XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.26

```
                         SYSTEM PERFORMANCE

 1. FEED MAKEUP

          TOTAL SOLIDS                       .9080E+06 KG/D ( .2000E+07 LB/D)
          SOLIDS FRACTION IN FEED                       .500
          VOLATILE SOLIDS FRACTION IN SOLIDS            .830
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS           .640

 2. PRETREATMENT

       SHREDDING-NONE USED

       DEGRITTER-NONE USED

       CHEMICAL PRETREATMENT-NONE USED

 3. STORAGE-WITH MIXING

          NUMBER OF UNITS                           2.
          CAPACITY PER UNIT         .5206E+04 CU.M. ( .1838E+06 CU FT.)
          POWER REQUIREMENT                         .7353E+02 HP

 4. DIGESTION

       CSTR
          NUMBER OF UNITS                           1.
          CAPACITY PER UNIT         .1485E+05 CU.M. ( .5244E+06 CU.FT.)
          DIGESTER TEMPERATURE                    60.C
          RETENTION TIME                      2.9DAYS
          CONVERSION EFFICIENCY                     .650
          POWER REQUIREMENT                         .1049E+03HP

       GAS PRODUCTION                               8.884 VVD
                                                 .4915E+04 GJ/D
                                               ( .4659E+04 MM BTU/D)
       HEAT EXCHANGER
          NUMBER OF UNITS                           6.
          HEAT EXCHANGER AREA       .8542E+03 SQ.M. (  .9191E+04 SQ.FT.)

       TOTAL HEAT REQUIREMENT     0.          GJ/D (0.          MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
STORAGE TANKS	686743.
DIGESTERS	1458207.
HEAT EXCHANGERS	243565.
MATERIALS HANDLING EQUIPMENT	119426.
SUBTOTAL	2507940.
STRUCTURES-ELECT. + PIPING	100318.
SUBTOTAL	2608258.
SUPPORT FACILITIES	130413.
SUBTOTAL CAPITAL INVESTMENT	2738671.
CONTRACTOR OVERHEAD + PROFIT	273867.
ENGINEERING + DESIGN	136934.
SUBTOTAL PLANT INVESTMENT	3149471.
PROJECT CONTINGENCY	472421.
TOTAL PLANT INVESTMENT	3621892.
INTEREST DURING CONSTRUCTION	325970.
START-UP COSTS	71706.
WORKING CAPITAL	72438.
TOTAL CAPITAL REQUIREMENT	4092006.

```
                     ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


    PRODUCTION MATERIALS

         FEED - $   0.00/KG ($   0.00/TON)                        0.

    UTILITIES

         ELECTRIC - 3.0 CTS./KW-HR                            31459.

         STEAM -  $ 3.79/GJ
                 ($ 4.00/MM BTU)                                  0.

    LABOR

         OPERATING LABOR - MEN PER SHIFT    2.
                           HRS PER SHIFT    8.0
                           HOURLY RATE   $ 6.00               35040.

         MAINTENANCE LABOR                                    54328.

         SUPERVISION                                          13405.

    ADMINISTRATION + OVERHEAD                                 61664.

    SUPPLIES

         OPERATING                                            10512.

         MAINTENANCE                                          54328.

    LOCAL TAXES + INSURANCE                                   97791.
                                                --------------------
 GROSS ANNUAL OPERATING COST                                 358529.

    CREDITS/PENALTIES COST                                        0.
                                                --------------------
 NET ANNUAL OPERATING COST                                   358529.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $    850217.        XX
XX                                                                XX
XX         TOTAL ANNUAL GAS PRODUCTION =  1614727. GJ             XX
XX                                      (  1530547. MM BTU)       XX
XX                                                                XX
XX         UNIT GAS COST = $    .53/GJ                            XX
XX                      ($    .56/MM BTU)                         XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.27

```
                        SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                         .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                         .500
        VOLATILE SOLIDS FRACTION IN SOLIDS              .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS             .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                       .200
        CONVERSION EFFICIENCY                           .500
        NUMBER OF UNITS                                1.
        CAPACITY PER UNIT       .1475E+02 CU.M. ( .5208E+03 CU.FT.)
        CHEMICAL REQUIREMENT    .1725E+04 KG/HR ( .1900E+01 TONS/HR)
        POWER REQUIREMENT                              0.         HP

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                                1.
        CAPACITY PER UNIT       .1061E+04 CU.M. ( .3746E+05 CU FT.)
        POWER REQUIREMENT                               .7493E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                                1.
        CAPACITY PER UNIT       .3260E+04 CU.M. ( .1151E+06 CU.FT.)
        DIGESTER TEMPERATURE                          60.C
        RETENTION TIME                           6.1DAYS
        CONVERSION EFFICIENCY                           .800
        POWER REQUIREMENT                               .2302E+02HP

    GAS PRODUCTION                                     5.722 VVD
                                                   .6947E+03 GJ/D
                                                ( .6586E+03 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                                1.
        HEAT EXCHANGER AREA     .5223E+03 SQ.M. (  .5620E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT    0.         GJ/D (0.           MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
CHEMICAL PRETREATMENT	24620.
STORAGE TANKS	150159.
DIGESTERS	467585.
HEAT EXCHANGERS	28068.
MATERIALS HANDLING EQUIPMENT	33522.
SUBTOTAL	703953.
STRUCTURES-ELECT. + PIPING	28158.
SUBTOTAL	732112.
SUPPORT FACILITIES	36606.
SUBTOTAL CAPITAL INVESTMENT	768717.
CONTRACTOR OVERHEAD + PROFIT	76872.
ENGINEERING + DESIGN	38436.
SUBTOTAL PLANT INVESTMENT	884025.
PROJECT CONTINGENCY	132604.
TOTAL PLANT INVESTMENT	1016628.
INTEREST DURING CONSTRUCTION	91497.
START-UP COSTS	67674.
WORKING CAPITAL	20333.
TOTAL CAPITAL REQUIREMENT	1196131.

```
                    ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $   0.00/KG ($   0.00/TON)                        0.

          PRETREATMENT CHEMICAL
               $    .31/KG ($ 300.00/TON)                     187245.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                             5380.

          STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                                  0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00               35040.

          MAINTENANCE LABOR                                    15249.

          SUPERVISION                                           7543.

     ADMINISTRATION + OVERHEAD                                 34700.

     SUPPLIES

          OPERATING                                            10512.

          MAINTENANCE                                          15249.

     LOCAL TAXES + INSURANCE                                   27449.
                                                    --------------------
  GROSS ANNUAL OPERATING COST                                 338368.

     CREDITS/PENALTIES COST                                        0.
                                                    --------------------
  NET ANNUAL OPERATING COST                                   338368.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX          TOTAL ANNUAL CAPITAL REQUIREMENT = $     482077.        XX
XX                                                                  XX
XX          TOTAL ANNUAL GAS PRODUCTION =     228235. GJ            XX
XX                                       (    216337. MM BTU)       XX
XX                                                                  XX
XX          UNIT GAS COST = $    2.11/GJ                            XX
XX                         ($    2.23/MM BTU)                       XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.28

```
                          SYSTEM PERFORMANCE

 1. FEED MAKEUP

        TOTAL SOLIDS                    .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                   .500
        VOLATILE SOLIDS FRACTION IN SOLIDS        .830
        BIODEGR. VOL. SOL. FRACT.  IN SOLIDS      .640

 2. PRETREATMENT

      SHREDDING-NONE USED

      DEGRITTER-NONE USED

      CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                  .200
        CONVERSION EFFICIENCY                      .500
        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT        .1475E+02 CU.M. ( .5208E+03 CU.FT.)
        CHEMICAL REQUIREMENT     .1725E+04 KG/HR ( .1900E+01 TONS/HR)
        POWER REQUIREMENT                       0.          HP

  3. STORAGE-WITH MIXING

        NUMBER OF UNITS                          1.
        CAPACITY PER UNIT        .5305E+03 CU.M. ( .1873E+05 CU FT.)
        POWER REQUIREMENT                         .3746E+01 HP

  4. DIGESTION

      PLUG FLOW
        NUMBER OF UNITS                          3.
        CAPACITY PER UNIT        .8138E+03 CU.M. ( .2873E+05 CU.FT.)
        DIGESTER TEMPERATURE                    60.C
        RETENTION TIME                      4.6DAYS
        CONVERSION EFFICIENCY                     .950
        POWER REQUIREMENT                       0.          HP

      GAS PRODUCTION                             9.072 VVD
                                               .8250E+03 GJ/D
                                             ( .7820E+03 MM BTU/D)

      HEAT EXCHANGER
        NUMBER OF UNITS                          1.
        HEAT EXCHANGER AREA     .5223E+03 SQ.M. (  .5620E+04 SQ.FT.)

      TOTAL HEAT REQUIREMENT   0.          GJ/D (0.          MM BTU/D)

   5. DEWATERING-NONE USED

  6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
CHEMICAL PRETREATMENT	24620.
STORAGE TANKS	104716.
DIGESTERS	31932.
HEAT EXCHANGERS	28068.
MATERIALS HANDLING EQUIPMENT	9467.
SUBTOTAL	198803.
STRUCTURES-ELECT. + PIPING	7952.
SUBTOTAL	206756.
SUPPORT FACILITIES	10338.
SUBTOTAL CAPITAL INVESTMENT	217093.
CONTRACTOR OVERHEAD + PROFIT	21709.
ENGINEERING + DESIGN	10855.
SUBTOTAL PLANT INVESTMENT	249657.
PROJECT CONTINGENCY	37449.
TOTAL PLANT INVESTMENT	287106.
INTEREST DURING CONSTRUCTION	25840.
START-UP COSTS	56575.
WORKING CAPITAL	5742.
TOTAL CAPITAL REQUIREMENT	375262.

```
                         ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                         0.

          PRETREATMENT CHEMICAL
                 $     .31/KG ($ 300.00/TON)                     187245.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                 661.

          STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                                     0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                  35040.

          MAINTENANCE LABOR                                        4307.

          SUPERVISION                                              5902.

     ADMINISTRATION + OVERHEAD                                    27149.

     SUPPLIES

          OPERATING                                               10512.

          MAINTENANCE                                              4307.

     LOCAL TAXES + INSURANCE                                       7752.
                                                    --------------------
  GROSS ANNUAL OPERATING COST                                    282874.

     CREDITS/PENALTIES COST                                           0.
                                                    --------------------
  NET ANNUAL OPERATING COST                                      282874.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $       327947.       XX
XX                                                                  XX
XX         TOTAL ANNUAL GAS PRODUCTION =      271030. GJ            XX
XX                                         (    256900. MM BTU)     XX
XX                                                                  XX
XX         UNIT GAS COST = $    1.21/GJ                             XX
XX                        ($    1.28/MM BTU)                        XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.29

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                         .3314E+06 KG/D ( .7300E+06 LB/D)
        SOLIDS FRACTION IN FEED                          .500
        VOLATILE SOLIDS FRACTION IN SOLIDS               .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS              .640

2. PRETREATMENT

     SHREDDING-NONE USED

     DEGRITTER-NONE USED

     CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                        .200
        CONVERSION EFFICIENCY                            .500
        NUMBER OF UNITS                                 1.
        CAPACITY PER UNIT        .7657E+02 CU.M. ( .2704E+04 CU.FT.)
        CHEMICAL REQUIREMENT     .6297E+04 KG/HR ( .6935E+01 TONS/HR)
        POWER REQUIREMENT                               0.          HP

3. STORAGE-NONE USED

4. DIGESTION

     BATCH
        NUMBER OF UNITS                                 1.
        CAPACITY PER UNIT        .1654E+06 CU.M. ( .5840E+07 CU.FT.)
        DIGESTER TEMPERATURE                          37.C
        RETENTION TIME                            25.2DAYS
        CONVERSION EFFICIENCY                            .950
        POWER REQUIREMENT                               0.          HP

     GAS PRODUCTION                                      .489 VVD
                                                       .3011E+04 GJ/D
                                                    ( .2855E+04 MM BTU/D)

     TOTAL HEAT REQUIREMENT    0.          GJ/D (0.            MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
CHEMICAL PRETREATMENT	84672.
DIGESTERS	865001.
MATERIALS HANDLING EQUIPMENT	47484.
SUBTOTAL	997157.
STRUCTURES-ELECT. + PIPING	39886.
SUBTOTAL	1037043.
SUPPORT FACILITIES	51852.
SUBTOTAL CAPITAL INVESTMENT	1088895.
CONTRACTOR OVERHEAD + PROFIT	108890.
ENGINEERING + DESIGN	54445.
SUBTOTAL PLANT INVESTMENT	1252229.
PROJECT CONTINGENCY	187834.
TOTAL PLANT INVESTMENT	1440064.
INTEREST DURING CONSTRUCTION	129606.
START-UP COSTS	72492.
WORKING CAPITAL	28801.
TOTAL CAPITAL REQUIREMENT	1670962.

```
                          ANNUAL OPERATING COSTS

                                                         ANNUAL COST ($)


      PRODUCTION MATERIALS

           FEED - $   0.00/KG ($   0.00/TON)                           0.

           PRETREATMENT CHEMICAL
                 $    .31/KG ($ 300.00/TON)                       187245.

      UTILITIES

           ELECTRIC - 3.0 CTS./KW-HR                                   0.

           STEAM -  $ 3.79/GJ
                   ($ 4.00/MM BTU)                                     0.

      LABOR

           OPERATING LABOR - MEN PER SHIFT    2.
                             HRS PER SHIFT    8.0
                             HOURLY RATE   $ 6.00                  35040.

           MAINTENANCE LABOR                                       21601.

           SUPERVISION                                              8496.

      ADMINISTRATION + OVERHEAD                                    39082.

      SUPPLIES

           OPERATING                                               10512.

           MAINTENANCE                                             21601.

      LOCAL TAXES + INSURANCE                                      38882.
                                                          ------------------
  GROSS ANNUAL OPERATING COST                                     362459.

      CREDITS/PENALTIES COST                                           0.
                                                          ------------------
  NET ANNUAL OPERATING COST                                       362459.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                        XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      563224.               XX
XX                                                                        XX
XX        TOTAL ANNUAL GAS PRODUCTION =     271042. GJ                    XX
XX                                    (     256912. MM BTU)               XX
XX                                                                        XX
XX        UNIT GAS COST = $    2.08/GJ                                    XX
XX                       ($    2.19/MM BTU)                               XX
XX                                                                        XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX)XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.30

```
                    SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                   .3314E+06 KG/D ( .7300E+06 LB/D)
        SOLIDS FRACTION IN FEED                  .500
        VOLATILE SOLIDS FRACTION IN SOLIDS       .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS      .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT-NONE USED

3. STORAGE-NONE USED

4. DIGESTION

    BATCH
        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT     .1654E+06 CU.M. ( .5840E+07 CU.FT.)
        DIGESTER TEMPERATURE                   37.C
        RETENTION TIME                 25.2DAYS
        CONVERSION EFFICIENCY                    .950
        POWER REQUIREMENT                       0.        HP

    GAS PRODUCTION                               .426 VVD
                                            .2622E+04 GJ/D
                                          ( .2486E+04 MM BTU/D)

    TOTAL HEAT REQUIREMENT    0.          GJ/D (0.          MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

```
                    CAPITAL COST ESTIMATE

                                                AUG 1978 COSTS ($)

      INSTALLED EQUIPMENT

          DIGESTERS                                          865001.
          MATERIALS HANDLING EQUIPMENT                        43250.
                                                --------------------
   SUBTOTAL                                                  908251.

        STRUCTURES-ELECT. + PIPING                            36330.
                                                --------------------
   SUBTOTAL                                                  944582.

        SUPPORT FACILITIES                                    47229.
                                                --------------------
   SUBTOTAL CAPITAL INVESTMENT                               991811.

        CONTRACTOR OVERHEAD + PROFIT                          99181.

        ENGINEERING + DESIGN                                  49591.
                                                --------------------
   SUBTOTAL PLANT INVESTMENT                                1140582.

        PROJECT CONTINGENCY                                  171087.
                                                --------------------
   TOTAL PLANT INVESTMENT                                   1311670.

        INTEREST DURING CONSTRUCTION                         118050.

        START-UP COSTS                                        33256.

        WORKING CAPITAL                                       26233.
                                                --------------------
   TOTAL CAPITAL REQUIREMENT                                1489209.
```

```
                    ANNUAL  OPERATING COSTS

                                                    ANNUAL  COST  ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                          0.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                    0.

          STEAM -  $ 3.79/GJ
                 ($ 4.00/MM BTU)                                       0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT   2.
                            HRS PER SHIFT   8.0
                            HOURLY RATE   $ 6.00                   35040.

          MAINTENANCE LABOR                                        19675.

          SUPERVISION                                               8207.

     ADMINISTRATION + OVERHEAD                                     37753.

     SUPPLIES

          OPERATING                                                10512.

          MAINTENANCE                                              19675.

     LOCAL TAXES + INSURANCE                                       35415.
                                                         ----------------
  GROSS ANNUAL OPERATING COST                                     166278.

     CREDITS/PENALTIES COST                                            0.
                                                         ----------------
  NET ANNUAL OPERATING COST                                       166278.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      345216.          XX
XX                                                                XX
XX         TOTAL ANNUAL GAS PRODUCTION =      236039. GJ            XX
XX                                       (    223706. MM BTU)       XX
XX                                                                XX
XX         UNIT GAS COST = $    1.46/GJ                             XX
XX                        ($    1.54/MM BTU)                        XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.31

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                         .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                        .500
        VOLATILE SOLIDS FRACTION IN SOLIDS             .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS            .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                      .200
        CONVERSION EFFICIENCY                          .500
        NUMBER OF UNITS                               1.
        CAPACITY PER UNIT         .1475E+02 CU.M. ( .5208E+03 CU.FT.)
        CHEMICAL REQUIREMENT      .1725E+04 KG/HR ( .1900E+01 TONS/HR)
        POWER REQUIREMENT                             0.         HP

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                               1.
        CAPACITY PER UNIT         .1061E+04 CU.M. ( .3746E+05 CU FT.)
        POWER REQUIREMENT                              .7493E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                               1.
        CAPACITY PER UNIT         .3260E+04 CU.M. ( .1151E+06 CU.FT.)
        DIGESTER TEMPERATURE                         60.C
        RETENTION TIME                          6.1DAYS
        CONVERSION EFFICIENCY                          .800
        POWER REQUIREMENT                              .2302E+02HP

    GAS PRODUCTION                                    5.722 VVD
                                                  .6947E+03 GJ/D
                                               ( .6586E+03 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                               1.
        HEAT EXCHANGER AREA       .5223E+03 SQ.M. (  .5620E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT     0.          GJ/D (0.          MM BTU/D)


5. DEWATERING
        NUMBER OF UNITS                               1.
        CAPACITY PER UNIT         .2021E+05 KG/HR ( .2226E+02 TONS/HR)
        POWER REQUIREMENT                              .7011E+02 HP

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
CHEMICAL PRETREATMENT	24620.
STORAGE TANKS	150159.
DIGESTERS	467585.
HEAT EXCHANGERS	28068.
DEWATERING UNITS	556931.
MATERIALS HANDLING EQUIPMENT	61368.
SUBTOTAL	1288731.
STRUCTURES-ELECT. + PIPING	51549.
SUBTOTAL	1340280.
SUPPORT FACILITIES	67014.
SUBTOTAL CAPITAL INVESTMENT	1407294.
CONTRACTOR OVERHEAD + PROFIT	140729.
ENGINEERING + DESIGN	70365.
SUBTOTAL PLANT INVESTMENT	1618388.
PROJECT CONTINGENCY	242758.
TOTAL PLANT INVESTMENT	1861147.
INTEREST DURING CONSTRUCTION	167503.
START-UP COSTS	81902.
WORKING CAPITAL	37223.
TOTAL CAPITAL REQUIREMENT	2147774.

```
                      ANNUAL OPERATING COSTS

                                                  ANNUAL COST ($)


   PRODUCTION MATERIALS

       FEED - $   0.00/KG ($   0.00/TON)                        0.

       PRETREATMENT CHEMICAL
             $    .31/KG ($ 300.00/TON)                    187245.

   UTILITIES

       ELECTRIC - 3.0 CTS./KW-HR                            17742.

       STEAM -  $ 3.79/GJ
             ($ 4.00/MM BTU)                                    0.

   LABOR

       OPERATING LABOR - MEN PER SHIFT    2.
                         HRS PER SHIFT    8.0
                         HOURLY RATE   $ 6.00               35040.

       MAINTENANCE LABOR                                    27917.

       SUPERVISION                                           9444.

   ADMINISTRATION + OVERHEAD                                43440.

   SUPPLIES

       OPERATING                                            10512.

       MAINTENANCE                                          27917.

   LOCAL TAXES + INSURANCE                                  50251.
                                                 --------------------
GROSS ANNUAL OPERATING COST                                409508.

   CREDITS/PENALTIES COST                                       0.
                                                 --------------------
NET ANNUAL OPERATING COST                                  409508.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX       TOTAL ANNUAL CAPITAL REQUIREMENT = $      667565.          XX
XX                                                                  XX
XX       TOTAL ANNUAL GAS PRODUCTION =      228235. GJ              XX
XX                                      (   216337. MM BTU)         XX
XX                                                                  XX
XX       UNIT GAS COST = $    2.92/GJ                               XX
XX                      ($    3.09/MM BTU)                          XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.32

```
                         SYSTEM PERF(RMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                    .9080E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                    .500
        VOLATILE SOLIDS FRACTION IN SOLIDS         .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                  .200
        CONVERSION EFFICIENCY                      .500
        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT       .1475E+02 CU.M. ( .5208E+03 CU.FT.)
        CHEMICAL REQUIREMENT    .1725E+04 KG/HR ( .1900E+01 TONS/HR)
        POWER REQUIREMENT                         0.          HP

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT       .1061E+04 CU.M. ( .3746E+05 CU FT.)
        POWER REQUIREMENT                          .7493E+01 HP

4. DIGESTION

    CSTR
        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT       .3260E+04 CU.M. ( .1151E+06 CU.FT.)
        DIGESTER TEMPERATURE                     60.C
        RETENTION TIME                     6.1DAYS
        CONVERSION EFFICIENCY                      .800
        POWER REQUIREMENT                          .2302E+02HP

    GAS PRODUCTION                                5.722 VVD
                                               .6947E+03 GJ/D
                                             ( .6586E+03 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                           1.
        HEAT EXCHANGER AREA     .5223E+03 SQ.M. (  .5620E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT    0.           GJ/D (0.          MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION

        CAPACITY PER UNIT   .3108E+05 CU.M/D ( .1098E+07 CU.FT/D)
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
CHEMICAL PRETREATMENT	24620.
STORAGE TANKS	150159.
DIGESTERS	467585.
HEAT EXCHANGERS	28068.
GAS PURIFICATION UNITS	453797.
MATERIALS HANDLING EQUIPMENT	33522.
SUBTOTAL	1157750.
STRUCTURES-ELECT. + PIPING	46310.
SUBTOTAL	1204060.
SUPPORT FACILITIES	60203.
SUBTOTAL CAPITAL INVESTMENT	1264264.
CONTRACTOR OVERHEAD + PROFIT	126426.
ENGINEERING + DESIGN	63213.
SUBTOTAL PLANT INVESTMENT	1453903.
PROJECT CONTINGENCY	218085.
TOTAL PLANT INVESTMENT	1671989.
INTEREST DURING CONSTRUCTION	150479.
START-UP COSTS	10676.
WORKING CAPITAL	33440.
TOTAL CAPITAL REQUIREMENT	1962676.

```
                    ANNUAL  OPERATING  COSTS

                                               ANNUAL  COST  ($)


    PRODUCTION  MATERIALS

         FEED  -  $    0.00/KG  ($    0.00/TON)                       0.

         PRETREATMENT  CHEMICAL
               $     .31/KG  ($ 300.00/TON)                      187245.

    UTILITIES

         ELECTRIC  -  3.0  CTS./KW-HR                              5380.

         STEAM  -   $ 3.79/GJ
                 ($ 4.00/MM  BTU)                                     0.

    LABOR

         OPERATING  LABOR  -  MEN  PER  SHIFT    2.
                              HRS  PER  SHIFT   :8.0
                              HOURLY  RATE    $ 6.00              35040.

         MAINTENANCE  LABOR                                       25080.

         SUPERVISION                                               9018.

    ADMINISTRATION  +  OVERHEAD                                   41483.

    SUPPLIES

         OPERATING                                                18512.

         MAINTENANCE                                              25080.

    LOCAL  TAXES  +  INSURANCE                                    45144.

    OPERATING  COST  -  GAS  PURIFICATION                        149860.
                                                     --------------------
GROSS  ANNUAL  OPERATING  COST                                   533841.

    CREDITS/PENALTIES  COST                                           0.
                                                     --------------------
NET  ANNUAL  OPERATING  COST                                     533841.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX        TOTAL  ANNUAL  CAPITAL  REQUIREMENT  =  $     769632.     XX
XX                                                                  XX
XX        TOTAL  ANNUAL  GAS  PRODUCTION  =      228235.  GJ        XX
XX                                       (      216337.  MM  BTU)   XX
XX                                                                  XX
XX        UNIT  GAS  COST  =  $    3.37/GJ                          XX
XX                          ($    3.56/MM  BTU)                     XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX)XXXXYXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.33

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                     .9980E+05 KG/D ( .2000E+06 LB/D)
        SOLIDS FRACTION IN FEED                    .500
        VOLATILE SOLIDS FRACTION IN SOLIDS         .830
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .640

2. PRETREATMENT

    SHREDDING-NONE USED

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED                  .200
        CONVERSION EFFICIENCY                      .500
        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT      .1475E+02 CU.M. ( .5208E+03 CU.FT.)
        CHEMICAL REQUIREMENT   .1725E+04 KG/HR ( .1900E+01 TONS/HR)
        POWER REQUIREMENT                         0.         HP

3. STORAGE-WITH MIXING

        NUMBER OF UNITS                           1.
        CAPACITY PER UNIT      .1061E+04 CU.M. ( .3746E+05 CU FT.)
        POWER REQUIREMENT                          .7493E+01 HP

4. DIGESTION

    CSTR IN SERIES
        NUMBER OF UNITS                           4.
        CAPACITY PER UNIT      .4037E+03 CU.M. ( .1425E+05 CU.FT.)
        DIGESTER TEMPERATURE                     60.C
        RETENTION TIME                       3.0DAYS
        CONVERSION EFFICIENCY                      .800
        POWER REQUIREMENT                          .1140E+02HP

    GAS PRODUCTION                               11.551 VVD
                                               .6947E+03 GJ/D
                                             ( .6586E+03 MM BTU/D)
    HEAT EXCHANGER
        NUMBER OF UNITS                           1.
        HEAT EXCHANGER AREA    .5223E+03 SQ.M. (  .5620E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT    0.         GJ/D (0.          MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
CHEMICAL PRETREATMENT	24620.
STORAGE TANKS	150159.
DIGESTERS	390444.
HEAT EXCHANGERS	28068.
MATERIALS HANDLING EQUIPMENT	29665.
SUBTOTAL	622955.
STRUCTURES-ELECT. + PIPING	24918.
SUBTOTAL	647874.
SUPPORT FACILITIES	32394.
SUBTOTAL CAPITAL INVESTMENT	680267.
CONTRACTOR OVERHEAD + PROFIT	68027.
ENGINEERING + DESIGN	34013.
SUBTOTAL PLANT INVESTMENT	782307.
PROJECT CONTINGENCY	117346.
TOTAL PLANT INVESTMENT	899654.
INTEREST DURING CONSTRUCTION	80969.
START-UP COSTS	65636.
WORKING CAPITAL	17993.
TOTAL CAPITAL REQUIREMENT	1064251.

```
                          ANNUAL OPERATING COSTS

                                                      ANNUAL COST ($)

     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                         0.

          PRETREATMENT CHEMICAL
                 $    .31/KG ($ 300.00/TON)                      187245.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                3332.

          STEAM -  $ 3.79/GJ
                ($ 4.00/MM BTU)                                       0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                  35040.

          MAINTENANCE LABOR                                       13495.

          SUPERVISION                                              7280.

     ADMINISTRATION + OVERHEAD                                    33489.

     SUPPLIES

          OPERATING                                               10512.

          MAINTENANCE                                             13495.

     LOCAL TAXES + INSURANCE                                      24291.
                                                   --------------------
  GROSS ANNUAL OPERATING COST                                    328178.

     CREDITS/PENALTIES COST                                           0.
                                                   --------------------
  NET ANNUAL OPERATING COST                                      328178.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                   XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $      456040.         XX
XX                                                                   XX
XX         TOTAL ANNUAL GAS PRODUCTION =      228235. GJ             XX
XX                                       (    216337. MM BTU)        XX
XX                                                                   XX
XX         UNIT GAS COST = $    2.00/GJ                              XX
XX                        ($    2.11/MM BTU)                         XX
XX                                                                   XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.34

```
                        SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                        .3314E+06 KG/D ( .7306E+06 LB/D)
        SOLIDS FRACTION IN FEED                    .900
        VOLATILE SOLIDS FRACTION IN SOLIDS         .825
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS        .400

2. PRETREATMENT

    SHREDDING
        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT        .1534E+05 KG/HR ( .1690E+02 TONS/HR)
        POWER REQUIREMENT                      .2535E+03 HP

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED              .200
        CONVERSION EFFICIENCY                  .500
        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT        .4254E+02 CU.M. ( .1502E+04 CU.FT.)
        CHEMICAL REQUIREMENT     .1409E+05 KG/HR ( .1551E+02 TONS/HR)
        POWER REQUIREMENT                       0.              HP

3. STORAGE-NONE USED

4. DIGESTION

    BATCH
        NUMBER OF UNITS                         1.
        CAPACITY PER UNIT        .1654E+06 CU.M. ( .5840E+07 CU.FT.)
        DIGESTER TEMPERATURE                  37.C
        RETENTION TIME                    25.2DAYS
        CONVERSION EFFICIENCY                  .950
        POWER REQUIREMENT                       0.              HP

    GAS PRODUCTION                             .407 VVD
                                          .2510E+04 GJ/D
                                        ( .2379E+04 MM BTU/D)

    TOTAL HEAT REQUIREMENT    0.          GJ/D (0.           MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	281103.
CHEMICAL PRETREATMENT	54486.
DIGESTERS	432501.
MATERIALS HANDLING EQUIPMENT	38404.
SUBTOTAL	806494.
STRUCTURES-ELECT. + PIPING	32260.
SUBTOTAL	838753.
SUPPORT FACILITIES	41938.
SUBTOTAL CAPITAL INVESTMENT	880691.
CONTRACTOR OVERHEAD + PROFIT	88069.
ENGINEERING + DESIGN	44035.
SUBTOTAL PLANT INVESTMENT	1012795.
PROJECT CONTINGENCY	151919.
TOTAL PLANT INVESTMENT	1164714.
INTEREST DURING CONSTRUCTION	104824.
START-UP COSTS	117426.
WORKING CAPITAL	23294.
TOTAL CAPITAL REQUIREMENT	1410259.

```
                 ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


      PRODUCTION MATERIALS

           FEED - $    0.00/KG ($    0.00/TON)                          0.

           PRETREATMENT CHEMICAL
                  $    .31/KG ($ 300.00/TON)                       418837.

      UTILITIES

           ELECTRIC - 3.0 CTS./KW-HR                                12245.

           STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                                       0.

      LABOR

           OPERATING LABOR - MEN PER SHIFT    2.
                             HRS PER SHIFT    8.0
                             HOURLY RATE   $ 6.00                   35040.

           MAINTENANCE LABOR                                        17471.

           SUPERVISION                                               7877.

      ADMINISTRATION + OVERHEAD                                     36232.

      SUPPLIES

           OPERATING                                                10512.

           MAINTENANCE                                              17471.

      LOCAL TAXES + INSURANCE                                       31447.
                                                      -------------------
   GROSS ANNUAL OPERATING COST                                     587132.

      CREDITS/PENALTIES COST                                            0.
                                                      -------------------
   NET ANNUAL OPERATING COST                                       587132.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $     756553.          XX
XX                                                                  XX
XX        TOTAL ANNUAL GAS PRODUCTION =     225868. GJ              XX
XX                                     (    214093. MM BTU)         XX
XX                                                                  XX
XX        UNIT GAS COST = $    3.35/GJ                              XX
XX                       ($    3.53/MM BTU)                         XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.35

```
                           SYSTEM PERFORMANCE

1. FEED MAKEUP

      TOTAL SOLIDS                          .3314E+06 KG/D ( .7306E+06 LB/D)
      SOLIDS FRACTION IN FEED                         .900
      VOLATILE SOLIDS FRACTION IN SOLIDS              .825
      BIODEGR. VOL. SOL. FRACT. IN SOLIDS             .400

2. PRETREATMENT

    SHREDDING
      NUMBER OF UNITS                                1.
      CAPACITY PER UNIT         .1534E+05 KG/HR ( .1690E+02 TONS/HR)
      POWER REQUIREMENT                               .2535E+03 HP

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
      RATIO OF CHEMICAL TO FEED                       .200
      CONVERSION EFFICIENCY                           .500
      NUMBER OF UNITS                                1.
      CAPACITY PER UNIT         .4254E+02 CU.M. ( .1502E+04 CU.FT.)
      CHEMICAL REQUIREMENT      .1409E+05 KG/HR ( .1551E+02 TONS/HR)
      POWER REQUIREMENT                              0.          HP

3. STORAGE-NONE USED

4. DIGESTION

    BATCH
      NUMBER OF UNITS                                1.
      CAPACITY PER UNIT         .1654E+06 CU.M. ( .5840E+07 CU.FT.)
      DIGESTER TEMPERATURE                          37.0
      RETENTION TIME                          25.2DAYS
      CONVERSION EFFICIENCY                           .950
      POWER REQUIREMENT                              0.          HP

    GAS PRODUCTION                                    .437 VVD
                                                   .2510E+04 GJ/D
                                                 ( .2379E+04 MM BTU/D)

    TOTAL HEAT REQUIREMENT     0.          GJ/D (0.          MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	281103.
CHEMICAL PRETREATMENT	54486.
DIGESTERS	173000.
MATERIALS HANDLING EQUIPMENT	25429.
SUBTOTAL	534018.
STRUCTURES-ELECT. + PIPING	21361.
SUBTOTAL	555379.
SUPPORT FACILITIES	27769.
SUBTOTAL CAPITAL INVESTMENT	583148.
CONTRACTOR OVERHEAD + PROFIT	58315.
ENGINEERING + DESIGN	29157.
SUBTOTAL PLANT INVESTMENT	670620.
PROJECT CONTINGENCY	100593.
TOTAL PLANT INVESTMENT	771213.
INTEREST DURING CONSTRUCTION	69409.
START-UP COSTS	111949.
WORKING CAPITAL	15424.
TOTAL CAPITAL REQUIREMENT	967995.

```
                    ANNUAL OPERATING COSTS

                                                    ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $   0.00/KG ($   0.00/TON)                          0.

          PRETREATMENT CHEMICAL
                 $    .31/KG ($ 300.00/TON)                     418837.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                               12245.

          STEAM -   $ 3.79/GJ
                ($ 4.00/MM BTU)                                      0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                  35040.

          MAINTENANCE LABOR                                       11568.

          SUPERVISION                                              6991.

     ADMINISTRATION + OVERHEAD                                    32160.

     SUPPLIES

          OPERATING                                               10512.

          MAINTENANCE                                             11568.

     LOCAL TAXES + INSURANCE                                      20823.
                                                     ---------------------
  GROSS ANNUAL OPERATING COST                                    559745.

     CREDITS/PENALTIES COST                                           0.
                                                     ---------------------
  NET ANNUAL OPERATING COST                                      559745.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                   XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      676023.          XX
XX                                                                   XX
XX        TOTAL ANNUAL GAS PRODUCTION =      225868. GJ              XX
XX                                      (    214093. MM BTU)         XX
XX                                                                   XX
XX        UNIT GAS COST = $    2.99/GJ                               XX
XX                       ($    3.16/MM BTU)                          XX
XX                                                                   XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.36

```
                        SYSTEM PERFORMANCE

1. FEED MAKEUP

        TOTAL SOLIDS                .3314E+06 KG/D ( .7300E+06 LB/D)
        SOLIDS FRACTION IN FEED                .900
        VOLATILE SOLIDS FRACTION IN SOLIDS     .825
        BIODEGR. VOL. SOL. FRACT. IN SOLIDS    .400

2. PRETREATMENT

    SHREDDING
        NUMBER OF UNITS                       1.
        CAPACITY PER UNIT      .1534E+05 KG/HR ( .1690E+02 TONS/HR)
        POWER REQUIREMENT                      .2535E+03 HP

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
        RATIO OF CHEMICAL TO FEED              .200
        CONVERSION EFFICIENCY                  .500
        NUMBER OF UNITS                       1.
        CAPACITY PER UNIT      .4254E+02 CU.M. ( .1502E+04 CU.FT.)
        CHEMICAL REQUIREMENT   .1409E+05 KG/HR ( .1551E+02 TONS/HR)
        POWER REQUIREMENT                     0.          HP

3. STORAGE-NONE USED

4. DIGESTION

    BATCH
        NUMBER OF UNITS                       1.
        CAPACITY PER UNIT      .1654E+06 CU.M. ( .5840E+07 CU.FT.)
        DIGESTER TEMPERATURE                 37.C
        RETENTION TIME               25.2DAYS
        CONVERSION EFFICIENCY                  .950
        POWER REQUIREMENT                     0.         HP

    GAS PRODUCTION                             .407 VVD
                                          .2510E+04 GJ/D
                                        ( .2379E+04 MM BTU/D)

    TOTAL HEAT REQUIREMENT    0.          GJ/D (0.         MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	281103.
CHEMICAL PRETREATMENT	54486.
DIGESTERS	86500.
MATERIALS HANDLING EQUIPMENT	21104.
SUBTOTAL	443193.
STRUCTURES-ELECT. + PIPING	17728.
SUBTOTAL	460921.
SUPPORT FACILITIES	23046.
SUBTOTAL CAPITAL INVESTMENT	483967.
CONTRACTOR OVERHEAD + PROFIT	48397.
ENGINEERING + DESIGN	24198.
SUBTOTAL PLANT INVESTMENT	556562.
PROJECT CONTINGENCY	83484.
TOTAL PLANT INVESTMENT	640046.
INTEREST DURING CONSTRUCTION	57604.
START-UP COSTS	110123.
WORKING CAPITAL	12801.
TOTAL CAPITAL REQUIREMENT	820574.

```
                        ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                    0.

          PRETREATMENT CHEMICAL
                 $     .31/KG ($ 300.00/TON)                418837.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                          12245.

          STEAM -   $ 3.79/GJ
                 ($ 4.00/MM BTU)                                 0.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00             35040.

          MAINTENANCE LABOR                                   9601.

          SUPERVISION                                         6696.

     ADMINISTRATION + OVERHEAD                               30802.

     SUPPLIES

          OPERATING                                          10512.

          MAINTENANCE                                         9601.

     LOCAL TAXES + INSURANCE                                 17281.
                                                   --------------------
 GROSS ANNUAL OPERATING COST                                550615.

     CREDITS/PENALTIES COST                                      0.
                                                   --------------------
 NET ANNUAL OPERATING COST                                  550615.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                  XX
XX         TOTAL ANNUAL CAPITAL REQUIREMENT = $     649180.          XX
XX                                                                  XX
XX         TOTAL ANNUAL GAS PRODUCTION =      225868. GJ             XX
XX                                      (     214093. MM BTU)       XX
XX                                                                  XX
XX         UNIT GAS COST = $     2.87/GJ                            XX
XX                        ($     3.03/MM BTU)                       XX
XX                                                                  XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.37

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

       TOTAL SOLIDS                      .9080E+05 KG/D ( .2000E+06 LB/D)
       SOLIDS FRACTION IN FEED                      .900
       VOLATILE SOLIDS FRACTION IN SOLIDS           .825
       BIODEGR. VOL. SOL. FRACT. IN SOLIDS          .400

2. PRETREATMENT

    SHREDDING
       NUMBER OF UNITS                            1.
       CAPACITY PER UNIT           .4204E+04 KG/HR ( .4630E+01 TONS/HR)
       POWER REQUIREMENT                            .6944E+02 HP

    DEGRITTER-NONE USED

    CHEMICAL PRETREATMENT
       RATIO OF CHEMICAL TO FEED                    .200
       CONVERSION EFFICIENCY                        .500
       NUMBER OF UNITS                            1.
       CAPACITY PER UNIT           .8194E+01 CU.M. ( .2893E+03 CU.FT.)
       CHEMICAL REQUIREMENT        .3859E+04 KG/HR ( .4250E+01 TONS/HR)
       POWER REQUIREMENT                          0.          HP

3. STORAGE-WITH MIXING

       NUMBER OF UNITS                            1.
       CAPACITY PER UNIT           .1085E+04 CU.M. ( .3833E+05 CU FT.)
       POWER REQUIREMENT                            .7665E+01 HP

4. DIGESTION

    CSTR
       NUMBER OF UNITS                            1.
       CAPACITY PER UNIT           .3335E+04 CU.M. ( .1178E+06 CU.FT.)
       DIGESTER TEMPERATURE                     60.C
       RETENTION TIME                       6.1DAYS
       CONVERSION EFFICIENCY                        .800
       POWER REQUIREMENT                            .2355E+02HP

    GAS PRODUCTION                                4.661 VVD
                                              .5790E+03 GJ/D
                                         ( .5488E+03 MM BTU/D)

    HEAT EXCHANGER
       NUMBER OF UNITS                            1.
       HEAT EXCHANGER AREA         .5343E+03 SQ.M. (  .5749E+04 SQ.FT.)

    TOTAL HEAT REQUIREMENT         .1399E+01 GJ/D ( .1326E+01 MM BTU/D)

 5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
CHEMICAL PRETREATMENT	15843.
STORAGE TANKS	151950.
DIGESTERS	475650.
HEAT EXCHANGERS	28552.
MATERIALS HANDLING EQUIPMENT	37450.
SUBTOTAL	786459.
STRUCTURES-ELECT. + PIPING	31458.
SUBTOTAL	817917.
SUPPORT FACILITIES	40896.
SUBTOTAL CAPITAL INVESTMENT	858813.
CONTRACTOR OVERHEAD + PROFIT	85881.
ENGINEERING + DESIGN	42941.
SUBTOTAL PLANT INVESTMENT	987635.
PROJECT CONTINGENCY	148145.
TOTAL PLANT INVESTMENT	1135780.
INTEREST DURING CONSTRUCTION	102220.
START-UP COSTS	76520.
WORKING CAPITAL	22716.
TOTAL CAPITAL REQUIREMENT	1337236.

```
                      ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)


     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                       0.

          PRETREATMENT CHEMICAL
                 $     .15/KG ($ 150.00/TON)                   209419.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                              17749.

          STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                                 1394.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                 35040.

          MAINTENANCE LABOR                                      17037.

          SUPERVISION                                             7812.

     ADMINISTRATION + OVERHEAD                                   35933.

     SUPPLIES

          OPERATING                                              10512.

          MAINTENANCE                                            17037.

     LOCAL TAXES + INSURANCE                                     30666.
                                                     --------------------
 GROSS ANNUAL OPERATING COST                                    382598.

     CREDITS/PENALTIES COST                                          0.
                                                     --------------------
 NET ANNUAL OPERATING COST                                      382598.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $     543259.          XX
XX                                                                XX
XX        TOTAL ANNUAL GAS PRODUCTION =     190196. GJ              XX
XX                                    (    180281. MM BTU)          XX
XX                                                                XX
XX        UNIT GAS COST = $    2.86/GJ                              XX
XX                       ($    3.01/MM BTU)                         XX
XX                                                                XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.38

```
                         SYSTEM PERFORMANCE

 1. FEED MAKEUP

          TOTAL SOLIDS               .9080E+05 KG/D ( .2000E+06 LB/D)
          SOLIDS FRACTION IN FEED                 .900
          VOLATILE SOLIDS FRACTION IN SOLIDS      .825
          BIODEGR. VOL. SOL. FRACT. IN SOLIDS     .400

 2. PRETREATMENT

      SHREDDING
          NUMBER OF UNITS                        1.
          CAPACITY PER UNIT        .4204E+04 KG/HR ( .4630E+01 TONS/HR)
          POWER REQUIREMENT                       .6944E+02 HP

      DEGRITTER-NONE USED

      CHEMICAL PRETREATMENT
          RATIO OF CHEMICAL TO FEED               .200
          CONVERSION EFFICIENCY                   .500
          NUMBER OF UNITS                        1.
          CAPACITY PER UNIT        .8194E+01 CU.M. ( .2893E+03 CU.FT.)
          CHEMICAL REQUIREMENT     .3859E+04 KG/HR ( .4250E+01 TONS/HR)
          POWER REQUIREMENT                      0.            HP

 3. STORAGE-WITH MIXING

          NUMBER OF UNITS                        1.
          CAPACITY PER UNIT        .1085E+04 CU.M. ( .3833E+05 CU FT.)
          POWER REQUIREMENT                       .7665E+01 HP

 4. DIGESTION

      CSTR
          NUMBER OF UNITS                        1.
          CAPACITY PER UNIT        .3335E+04 CU.M. ( .1178E+06 CU.FT.)
          DIGESTER TEMPERATURE                 60.C
          RETENTION TIME                   5.1DAYS
          CONVERSION EFFICIENCY                   .800
          POWER REQUIREMENT                       .2355E+02HP

      GAS PRODUCTION                             4.661 VVD
                                              .5790E+03 GJ/D
                                            ( .5488E+03 MM BTU/D)
      HEAT EXCHANGER
          NUMBER OF UNITS                        1.
          HEAT EXCHANGER AREA      .5343E+03 SQ.M. (  .5749E+04 SQ.FT.)

      TOTAL HEAT REQUIREMENT       .1399E+01 GJ/D ( .1326E+01 MM BTU/D)

  5. DEWATERING-NONE USED

 6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
CHEMICAL PRETREATMENT	15843.
STORAGE TANKS	151950.
DIGESTERS	475650.
HEAT EXCHANGERS	28552.
MATERIALS HANDLING EQUIPMENT	37450.
SUBTOTAL	786459.
STRUCTURES-ELECT. + PIPING	31458.
SUBTOTAL	817917.
SUPPORT FACILITIES	40896.
SUBTOTAL CAPITAL INVESTMENT	858813.
CONTRACTOR OVERHEAD + PROFIT	85881.
ENGINEERING + DESIGN	42941.
SUBTOTAL PLANT INVESTMENT	987635.
PROJECT CONTINGENCY	148145.
TOTAL PLANT INVESTMENT	1135780.
INTEREST DURING CONSTRUCTION	102220.
START-UP COSTS	51389.
WORKING CAPITAL	22716.
TOTAL CAPITAL REQUIREMENT	1312105.

```
                         ANNUAL OPERATING COSTS

                                                   ANNUAL COST ($)

     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                     0.

          PRETREATMENT CHEMICAL
                $     .06/KG ($  60.00/TON)                   83767.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                           17749.

          STEAM -  $ 3.79/GJ
                  ($ 4.00/MM BTU)                              1394.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00              35040.

          MAINTENANCE LABOR                                   17037.

          SUPERVISION                                          7812.

     ADMINISTRATION + OVERHEAD                                35933.

     SUPPLIES

          OPERATING                                           10512.

          MAINTENANCE                                         17037.

     LOCAL TAXES + INSURANCE                                  30666.
                                                     --------------------
   GROSS ANNUAL OPERATING COST                               256946.

     CREDITS/PENALTIES COST                                       0.
                                                     --------------------
   NET ANNUAL OPERATING COST                                 256946.

  XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
  XX                                                                  XX
  XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $      414597.         XX
  XX                                                                  XX
  XX        TOTAL ANNUAL GAS PRODUCTION =      190196. GJ             XX
  XX                                      (    180281. MM BTU)        XX
  XX                                                                  XX
  XX        UNIT GAS COST = $    2.18/GJ                              XX
  XX                     ($    2.30/MM BTU)                           XX
  XX                                                                  XX
  XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
```

Table F.39

```
                         SYSTEM PERFORMANCE

1. FEED MAKEUP

      TOTAL SOLIDS                       .9089E+05 KG/D ( .2000E+06 LB/D)
      SOLIDS FRACTION IN FEED                       .900
      VOLATILE SOLIDS FRACTION IN SOLIDS            .825
      BIODEGR. VOL. SOL. FRACT. IN SOLIDS           .400

2. PRETREATMENT

   SHREDDING
      NUMBER OF UNITS                             1.
      CAPACITY PER UNIT            .4204E+04 KG/HR ( .4630E+01 TONS/HR)
      POWER REQUIREMENT                             .6944E+02 HP

   DEGRITTER-NONE USED

   CHEMICAL PRETREATMENT
      RATIO OF CHEMICAL TO FEED                     .200
      CONVERSION EFFICIENCY                         .500
      NUMBER OF UNITS                             1.
      CAPACITY PER UNIT            .8194E+01 CU.M. ( .2893E+03 CU.FT.)
      CHEMICAL REQUIREMENT         .3859E+04 KG/HR ( .4250E+01 TONS/HR)
      POWER REQUIREMENT                           0.          HP

3. STORAGE-WITH MIXING

      NUMBER OF UNITS                             1.
      CAPACITY PER UNIT            .1085E+04 CU.M. ( .3833E+05 CU FT.)
      POWER REQUIREMENT                             .7665E+01 HP

4. DIGESTION

   CSTR
      NUMBER OF UNITS                             1.
      CAPACITY PER UNIT            .3335E+04 CU.M. ( .1178E+06 CU.FT.)
      DIGESTER TEMPERATURE                      60.C
      RETENTION TIME                       6.1DAYS
      CONVERSION EFFICIENCY                         .800
      POWER REQUIREMENT                             .2355E+02HP

   GAS PRODUCTION                                 4.661 VVD
                                             .5790E+03 GJ/D
                                           ( .5488E+03 MM BTU/D)

   HEAT EXCHANGER
      NUMBER OF UNITS                             1.
      HEAT EXCHANGER AREA          .5343E+03 SQ.M. (  .5749E+04 SQ.FT.)

   TOTAL HEAT REQUIREMENT          .1399E+01 GJ/D ( .1326E+01 MM BTU/D)

 5. DEWATERING-NONE USED

6. GAS PURIFICATION-NONE USED
```

CAPITAL COST ESTIMATE

	AUG 1978 COSTS ($)
INSTALLED EQUIPMENT	
SHREDDERS	77014.
CHEMICAL PRETREATMENT	15843.
STORAGE TANKS	151950.
DIGESTERS	475650.
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SUBTOTAL CAPITAL INVESTMENT	858813.
CONTRACTOR OVERHEAD + PROFIT	85881.
ENGINEERING + DESIGN	42941.
SUBTOTAL PLANT INVESTMENT	987635.
PROJECT CONTINGENCY	148145.
TOTAL PLANT INVESTMENT	1135780.
INTEREST DURING CONSTRUCTION	102220.
START-UP COSTS	38824.
WORKING CAPITAL	22716.
TOTAL CAPITAL REQUIREMENT	1299540.

```
                         ANNUAL OPERATING COSTS

                                                         ANNUAL COST ($)

     PRODUCTION MATERIALS

          FEED - $    0.00/KG ($    0.00/TON)                             0.

          PRETREATMENT CHEMICAL
                 $    .02/KG ($  15.00/TON)                           20942.

     UTILITIES

          ELECTRIC - 3.0 CTS./KW-HR                                   17749.

          STEAM -  $ 3.79/GJ
                 ($ 4.00/MM BTU)                                       1394.

     LABOR

          OPERATING LABOR - MEN PER SHIFT    2.
                            HRS PER SHIFT    8.0
                            HOURLY RATE   $ 6.00                      35040.

          MAINTENANCE LABOR                                           17037.

          SUPERVISION                                                  7812.

     ADMINISTRATION + OVERHEAD                                        35933.

     SUPPLIES

          OPERATING                                                   10512.

          MAINTENANCE                                                 17037.

     LOCAL TAXES + INSURANCE                                          30666.
                                                            --------------------
   GROSS ANNUAL OPERATING COST                                       194121.

     CREDITS/PENALTIES COST                                               0.
                                                            --------------------
   NET ANNUAL OPERATING COST                                         194121.

XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
XX                                                                      XX
XX        TOTAL ANNUAL CAPITAL REQUIREMENT = $       350266.            XX
XX                                                                      XX
XX        TOTAL ANNUAL GAS PRODUCTION =      190196. GJ                 XX
XX                                       (   180281. MM BTU)            XX
XX                                                                      XX
XX        UNIT GAS COST = $    1.84/GJ                                  XX
XX                        ($   1.94/MM BTU)                             XX
XX                                                                      XX
XXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXXX
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