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ANAEROBIC FERMENTATION OF BEEF CATTLE MANURE

FINAL REPORT

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FOREWORD

This report describes the results of a research project to produce energy in the form of methane and a high protein feed supplement from livestock manure. This work was jointly funded by the US Department of Agriculture, through the Science and Education Administration, and the US Department of Energy, through the Solar Energy Research Institute.

The results of this research indicates that there are many livestock operations where thermophillic fermentation of livestock manure would be both technically feasible and economically attractive. The development of this technology has now reached the point where a significant commercialization effort is needed, aimed at integrating such fermentation units into livestock production operations.

The USDA and DOE are currently working out arrangements for such a program.

Dan Jantzen
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SUMMARY

This report summarizes the research being conducted at the Roman L. Hruska U.S. Meat Animal Research Center to convert livestock manure and crop residues into methane and a high protein feed ingredient by thermophilic anaerobic fermentation. The major biological and operational factors involved in methanogenesis were discussed, and a kinetic model that describes the fermentation process was presented. Substrate biodegradability, fermentation temperature, and influent substrate concentration were shown to have significant effects on CH_4 production rate. The kinetic model predicted methane production rates of existing pilot and full-scale fermentation systems to within 15%.

The 5.7 m^3 fermentor was operated at: temperatures of 45, 50 and 55°C; hydraulic retention times ranging from 12 to 4 days; mixed continuously or 2 hr/day; and fed once/day or 22 times/day. No difference in methane production rate was observed when the fermentor was mixed 2 hr/day versus continuously. The methane production rate was about 10% higher when the fermentor was fed 22 times/day compared with once/day. The highest methane production rate achieved by the fermentor was 4.7 L CH_4/L fermentor·day. This is the highest rate reported in the literature and about 4 times higher than other pilot or full-scale systems fermenting livestock manures.

Assessment of the energy requirements for anaerobic fermentation systems showed that the major energy requirement for a thermophilic system was for maintaining the fermentor temperature. Of the total heating energy required, about 89 to 94% was for heating the influent slurry at an ambient temperature of 10°C. The next major energy consumption was due to the mixing of the influent slurry and fermentor liquor. Mixing amounted to 7.3% of the gross methane energy production, assuming continuous mixing. The least energy was consumed in pumping. The total energy required for mixing and pumping accounted for 10.8 to 11.3% of the gross thermal energy production.

An approach to optimizing anaerobic fermentor designs by selecting design criteria that maximize the net energy production per unit cost was presented. Using this optimization technique, we estimated that a farmer-constructed and operated system would be economically feasible for beef feedlots between 1,000 to 2,000 head without a feed credit assumed for the effluent, and about 300 head with a feed credit of \$60/Mg effluent total solids. Commercial "turn-key" systems are only feasible for feedlots larger than 8,000 head with no effluent credit, and feedlots between 1,000 to 2,000 head with an effluent credit of \$60/Mg. Based on these results, we believe that the economics of anaerobic fermentation is sufficiently favorable for farm-scale demonstration of this technology.

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SECTION 1.0

INTRODUCTION

This report summarizes the research being conducted at the Roman L. Hruska U.S. Meat Animal Research Center to assess the technical and economic feasibility of recovering methane and high protein biomass from the thermophilic fermentation of beef cattle and crop residues. Specific objectives are to:

1. Develop design criteria for optimum production of methane and/or biomass from anaerobic fermentation of livestock and crop residues,

2. Develop efficient methods to recover high protein biomass from the fermented residue,

3. Evaluate the nutritional value of the biomass as a livestock feed,

4. Determine the capital and operational costs, and energy, manpower and safety requirements for methane fermentation systems associated with livestock operations.

This project was initiated in 1976 and is jointly funded by the U.S. Department of Agriculture, Science and Education Administration, Agricultural Research and the U.S. Department of Energy, Biomass Energy Systems Branch/ Solar Energy Research Institute. The specific objectives of interest to the Department of Energy are Objectives 1 and 4 listed above. This report summarizes the completed research on thermophilic, anaerobic fermentation of beef cattle manure. Work is continuing on fermentation of crop residue.

SECTION 2.0

PRINCIPLES OF METHANE PRODUCTION

A. G. Hashimoto, Y. R. Chen, and V. H. Varel

2.1 INTRODUCTION

Methane (CH_4) is produced in nature through the anaerobic decomposition of organic matter. Since bacteria are the predominant species involved in methanogenesis, this discussion on the principles of CH_4 production deals with the bacteria involved in methanogenesis and the factors that affect both the rate of CH_4 production and the amount of organic matter that can be converted to CH_4.

2.2 MICROBIOLOGY

Methanogenesis has traditionally been viewed as a two-stage process -- the acid-forming and CH_4-forming stages (Kirsch and Sykes, 1971; Torien and Hattingh, 1969). In the first stage, acid-forming bacteria were thought to ferment organic materials, like carbohydrates, lipids, and proteins to formate, acetate, propionate, butyrate, ethanol, hydrogen (H_2) and carbon dioxide (CO_2). Bryant (1976, 1979) and McInerny and Bryant (1978) proposed a three-stage scheme that attempts to synthesize more current information on methanogenesis from organic matter. In general, the first stage involves species of fermentative bacteria which, as a metabolic group, hydrolyze complex carbohydrates, proteins, and lipids and ferment these products to fatty acids, H_2, and CO_2. The second metabolic group, called the "H_2-producing acetogenic bacteria" produce acetate, CO_2 and H_2 from the fatty acids generated in the first stage. The third stage involves the methanogenic bacteria that utilize the products of the first two stages -- mainly acetate, CO_2, and H_2 to produce CH_4 and CO_2. Recently, an additional stage was added to this scheme, as shown in Figure 2.1. This metabolic group is called the homoacetogenic bacteria which are reported to synthesize acetate using H_2, CO_2, and formate (Zeikus, 1979; Wolfe, 1979). Methanogenesis in the gastrointestinal tract of animals involves only the first metabolic group and H_2 utilization by methanogens (Hungate, 1966). Acetogenic bacteria are not significantly involved due to the short retention times in these ecosystems. Acetate and other volatile acids accumulate in rumen, fecal, and colon fermentations and are utilized as major energy sources by herbivorous animals.

Most of the information concerning extracellular intermediates important in methanogenesis comes from studies of rumen and sewage sludge fermentations (Hobson et al., 1974; Hungate, 1966, Torien and Hattingh, 1969; Wolin, 1974). Acetate is an important precursor in nature because about 70% of the CH_4 produced in sludge is produced via the methyl group of acetate (Kugelman and McCarty, 1965; Smith and Mah, 1966). Mountfort and Asher (1978) found that during the first few hours after a beef-manure fermentor is fed, up to 90% of the CH_4 produced comes from acetate. Reduction of CO_2 by H_2, and to some extent by other intermediate electron donors, accounts for the rest of the CH_4 production. Winfrey et al. (1977) showed that H_2 is an important intermediate and a rate-limiting factor in lake sediment methanogenesis. Formate is rapidly converted to H_2 and CO_2 by nonmethanogens or is directly utilized by methanogens (Hungate, 1966).

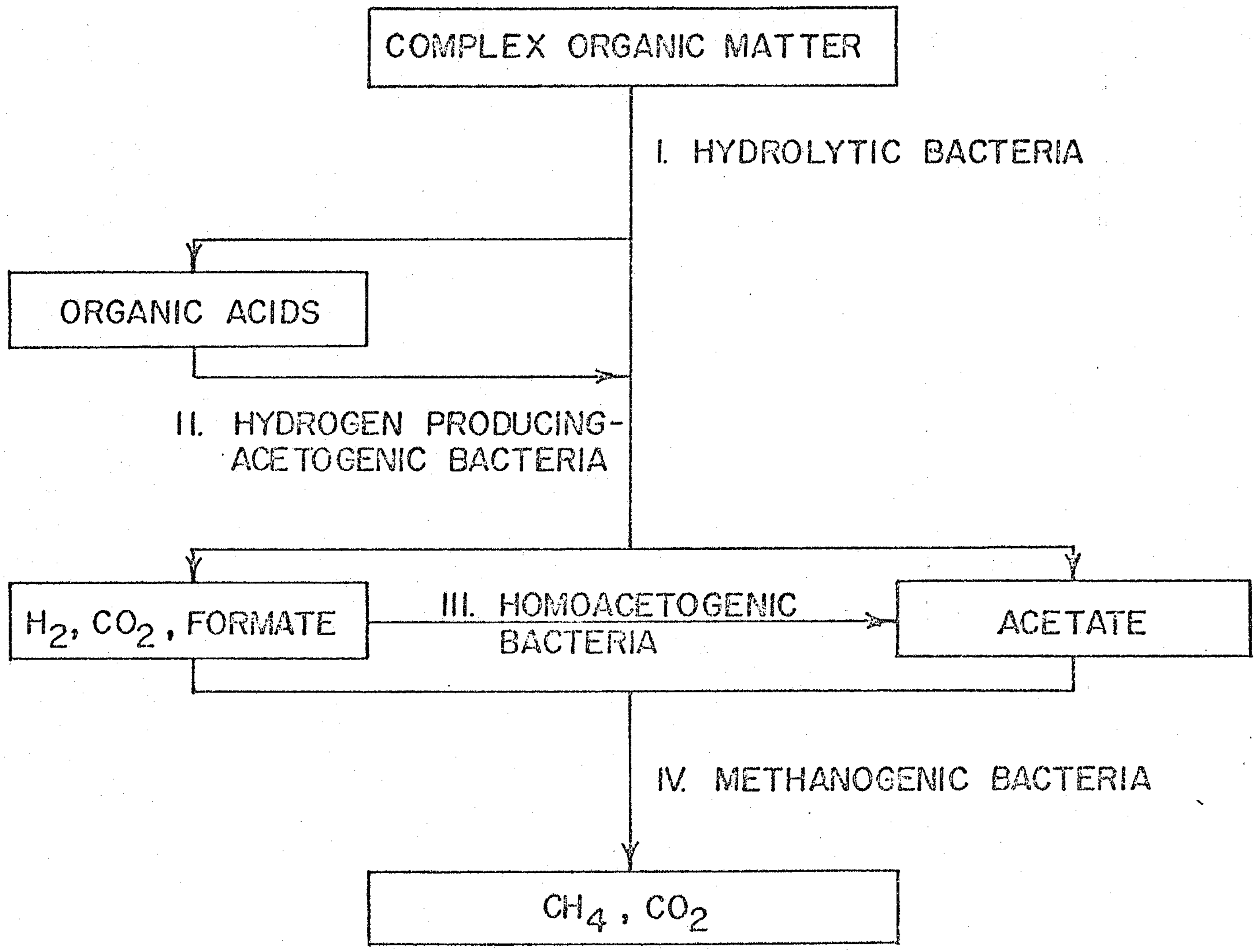

Figure 2.1. The four bacterial groups involved in the complete anaerobic degradation of organic matter.

Succinate is a major extracellular intermediate in the rumen, which is rapidly decarboxylated to propionate (Hungate, 1966; Scheifinger et al., 1973). Other than acetate and H_2, propionate is probably the most important intermediate in methanogenesis (McCarty, 1964c; Smith and Mah, 1966). Kaspar and Wuhrman (1978) calculated that 15% of the total steady-state CH_4 production is derived from propionate. Definitive kinetic studies, such as those of Smith and Mah (1966) on acetate, have not been reported on butyrate or longer carbon-chained acids.

Ethanol and lactate are probably not important intermediates. Organisms produce these products to dispose of electrons generated in glycolysis, but they also produce H_2. In the natural system, H_2-using methanogenic bacteria rapidly use the H_2, which allows the fermentative bacteria to produce more H_2 and acetate and less lactate and ethanol. Thus, in the rumen, ethanol is neither produced nor used, though many bacterial species produce ethanol in pure culture (Hungate, 1966). Only under stress of feeding high substrate levels does lactate become an important intermediate in the rumen. Wolin (1974, 1976) and Bryant (1976, 1979) discussed in detail the research dealing with altered electron flow in the direction of H_2 production caused by the metabolic interactions of methanogens and nonmethanogens.

2.3 ENVIRONMENTAL CONSIDERATIONS

Environmental factors influence the rate and amount of CH_4 produced during methanogenesis. Some of the major environmental factors are pH, alkalinity, volatile acids, temperature, nutrients, and toxic materials. Several authors have reviewed the influence of these factors on methanogenesis (Mah et al., 1977; Wolfe, 1971; Zeikus, 1977; Hobson et al., 1974; Torien and Hattingh, 1969; Speece and McCarty, 1964; Kirsch and Sykes, 1971).

2.3.1 pH

The methanogenic and acetogenic bacteria seem to be sensitive to pH. The pH, in turn, is a function of the bicarbonate alkalinity, the CO_2 partial pressure, and the volatile acids concentration. McCarty (1964a) reported that CH_4 production proceeds quite well as long as the pH is maintained between 6.6 and 7.6, with an optimum range between 7.0 and 7.2. At pH values below 6.2, toxicity is acute. Alkali should be added to maintain the pH above 6.6. High pH can be a problem with CH_4 production from animal manure because of the high levels of ammonia generated at high organic loading rates (Jewell et al., 1976).

2.3.2 Alkalinity

Alkalinity is a measure of the buffering capacity of the fermentor contents and consists of the bicarbonate, carbonate, ammonia, and hydroxide components. Organic acids and acid salts may also contribute to the buffering capacity (Am. Public Health Assoc., 1975). McCarty (1964a) indicated that a bicarbonate alkalinity in the range of 2.5 to 5.0 g $CaCO_3$/L provides a safe buffering capacity for anaerobic treatment of waste. Sievers and Brune (1978) and Kroeker et al. (1979) reported on the importance of ammonia in buffering animal manure fermentations. The relatively low carbon:nitrogen ratio of animal manures was reported as a major factor in the stability of animal manure fermentations. Ammonia was reported to contribute to the process stability by increasing the bicarbonate buffering capacity and increasing the pH.

2.3.3 Volatile Acids

McCarty and McKinney (1961) found that volatile acid levels should remain below 2.0 g acetate/L for efficient fermentation. Above this level, the acids were toxic. This seems to hold true for thermophilic temperatures also, as Varel et al. (1977) reported less efficient CH_4 production from cattle manure when the level of organic acids rose above 2.0 g/L. Kroeker et al. (1979) showed acute methanogenic toxicity at unionized volatile acid concentrations between 30 to 60 mg/L as acetic acid. This corresponded to total volatile acid concentrations between 1.65 to 2.6 g/L as acetic acid.

2.3.4 Temperature

Temperature is an important environmental parameter in anaerobic fermentation processes. Faster fermentation rates, faster solid-liquid separation and minimization of bacterial and viral pathogens are some benefits attributed to thermophilic fermentation (Pfeffer, 1974; Cooney and Wise, 1975). Pfeffer (1974) used shredded municipal refuse to establish two optimum temperatures. The optimum in the mesophilic and thermophilic range was 42 and 60°C, respectively. He also concluded that it was less expensive to produce CH_4 at the higher temperature. A definite acclimation period was required to initiate thermophilic fermentation. Buhr and Andrews (1977) stated that although the literature is contradictory, minor fluctuations in temperature can cause problems for thermophilic fermentors. Golueke (1958) found that the total volatile acids increased as temperature increased between 35 and 65°C.

Although the rates of reaction in the thermophilic range are much faster than those in the mesophilic range, most sewage sludge fermentation systems have operated under mesophilic conditions (McCarty, 1964a). In the past, energy requirements to maintain thermophilic temperatures were thought to be excessive due to the high water content of sewage sludges. Studies on urban refuse indicate that thermophilic temperatures are more economical and efficient for CH_4 production (Pfeffer and Liebman, 1976; Pfeffer, 1974). Results published in this report show that thermophilic fermentation of beef cattle manure is more economical than mesophilic fermentation.

2.3.5 Nutrients

Another important environmental condition is the presence of the nutrients, like nitrogen, phosphorous, sulfur, and trace nutrients, needed by bacteria (Bryant, 1974; Bryant et al., 1971; McCarty, 1964a). Animal manures and municipal sewage sludges usually contain all the required nutrients in adequate quantities, but other substrates may not. Pfeffer and Liebman (1976) found that municipal refuse was deficient in nitrogen and phosphorous. McCarty (1964b) reported that other elements having stimulatory effects at low concentrations include sodium, potassium, calcium, magnesium, and iron. All of these elements can exhibit inhibitory effects at higher concentrations. In general, the bacteria involved in methanogenesis have simple nutrient requirements and, although various individual species may require growth factors (e.g., B-vitamins, fatty acids, amino acids), these are supplied by other bacterial species (Bryant, 1974; Bryant et al., 1971).

The relative proportion of nutrients is also important in methanogenesis. Hills (1979) reported a 60 to 70% increase in CH_4 yield when the carbon:nitrogen ratio was increased from 8 to 25 by adding glucose or cellulose. Since most animal manures have carbon:nitrogen ratios between 6 to 10, the potential to increase CH_4 yields by adding carbonaceous materials to manures is apparent. The practical limitation of this concept, however, is that most crop residues are even less biodegradable than animal manures. Thus, pretreatment of crop residues is necessary to increase their biodegradability.

2.3.6 Toxic Materials

Other environmental factors involve toxicities resulting from excessive quantities of organic or inorganic substances. The threshold toxic levels of inorganic substances vary depending on whether the substrates act singly or in combination. Certain combinations have synergistic effects, whereas others display antagonistic effects (McCarty, 1946b; Kugelman and McCarty, 1965). Several investigators have implicated high concentrations of sulfate in retarding CH_4 production. But recently, Bryant et al. (1977) and Winfrey and Zeikus (1977) have independently proposed that competition for available H_2 is the mechanism by which sulfate inhibits methanogenesis in natural ecosystems. The sulfate-reducing bacteria apparently scavenge the available H_2 faster than the methanogens.

Inhibition by ammonia is a significant problem with some high rate fermentation processes, particularly when ammonia-rich manure from swine and poultry are fermented, and a proper acclimation period is not permitted (Lapp et al., 1975; Stevens and Schulte, 1979; Sievers and Brune, 1978; Kroeker et al., 1979; Converse et al., 1977a). McCarty (1964b) reported that at concentrations between 1.5 and 3.0 g/L of total ammonia nitrogen and at a pH greater than 7.4, the unionized ammonia may inhibit methanogenesis. At concentrations above 3.0 g/L, ammonia becomes toxic regardless of pH. However, Lapp et al. (1975), Converse et al. (1977a) and Fischer et al. (1979) have reported stable CH_4 production with ammonia concentrations in excess of 3.0 g/L (2.2 to 8.0 g/L). Kroeker et al. (1979) used a urea and acetic-acid substrate to investigate the effect of ammonia inhibition on CH_4 production. They concluded that CH_4 was progressively inhibited as the ammonia nitrogen concentration increased above 2 g/L; however, toxicity (i.e., complete cessation of CH_4 production) did not occur even at ammonia nitrogen concentrations of 7.0 g/L.

Antibiotics and growth promoters used in livestock rations can inhibit or even completely stop methanogenesis. Turnocliff and Custer (1978) reported that operating an anaerobic fermentation system where the antibiotic lincomycin is used is probably futile. Fischer et al. (1978) also reported severe fermentor instability when lincomycin was used in swine rations to control dysentery. Hashimoto et al. (1979) reported that chlortetracycline had no adverse effect on methanogenesis, but that monensin nearly doubled the time (from 20 to 40 days) for the start of CH_4 production in batch fermentations. After the bacteria adapted to the monensin, however, the fermentation proceeded at rates comparable to batch fermentations without monensin. Three possible mechanisms may explain the apparent adaptation of the bacteria to monensin or any other antibiotic: a) mutant strains of bacteria develop resistance to the antibiotic; b) microbial populations shift as the result of inhibition of some bacteria and increase in others; and/or c) the antibiotic is deactivated during the lag period. Chen and Wolin (1979) have evidence suggesting that the first two mechanisms listed above explain the role of monensin in the rumen. The Rumensin Technical Manual (Eli Lilly Co., 1975) shows that one part per million of monensin in soil samples is deactivated in 14 days when incubated with animal feces, and in 25 days when incubated without feces. Experiments on daily feeding of manure containing monensin to fermentors show unstable fermentation except at very long hydraulic retention times (30 to 40 days) (Varel and Hashimoto, 1981). More research on the effects of antibiotics on methanogenesis is necessary since antibiotics are widely used in livestock production.

2.4 FERMENTATION KINETICS

2.4.1 Kinetic Models

It is important to understand the kinetics of CH_4 fermentation to design and operate optimum systems. Several kinetic models have been used to describe the anaerobic fermentation process. The Monod (1950) kinetic model has been adapted to describe the anaerobic digestion kinetics of sewage sludge (O'Rourke, 1968; Lawrence and McCarty, 1969; Andrews and Pearson, 1965) and animal manures (Morris, 1976; Hill and Barth, 1977). The advantages of the Monod type model are that the kinetic parameters (the microorganism maximum specific growth rate and half-velocity constant) have deterministic connotations that describe the microbial processes, and the model can predict the conditions when maximum biological activity occurs and when activity ceases (i.e., wash-out). Disadvantages of the Monod model are that one set of kinetic parameters cannot describe the biological process at short and long retention times (Garrett and Sawyer, 1952; Chiu et al., 1972a,b), and that the kinetic parameters cannot be obtained for certain complex substrates (Pfeffer, 1974).

To overcome the disadvantages of the Monod model, various forms of the first-order kinetic model have been used (McKinney, 1962; Eckenfelder, 1963; Grau et al., 1975; Grady et al., 1972; Pfeffer, 1974; Morris, 1976). The advantages of the first-order models are that they are simple to use and give good fit of experimental data. Disadvantages are that they do not predict the conditions for maximum biological activity and system failure.

The Contois (1959) kinetic model has the advantages and generally avoids the disadvantages inherent in the Monod model. The Contois model was adapted to describe the kinetics of CH_4 fermentation as follows (Chen and Hashimoto, 1978):

$$\gamma_V = \frac{B_o\, S_o}{\Theta} \left[1 - \frac{K}{\Theta\mu_m - 1 + K}\right] \qquad (2.1)$$

where:

γ_V = volumetric CH_4 production rate, L CH_4/L fermentor·day;

S_o = influent total volatile solids (VS) concentration, g/L;

B_o = ultimate CH_4 yield, L CH_4/g VS added as $\Theta \longrightarrow \infty$;

Θ = hydraulic retention time, day;

μ_m = maximum specific growth rate of microorganisms, day^{-1};

K = kinetic parameter, dimensionless.

Equation 2.1 states that for a given loading rate (S_o/Θ), the daily volume of

CH_4 per volume of fermentor depends on the biodegradability of the material (B_o) and the kinetic parameters μ_m and K.

2.4.2 Ultimate Methane Yield (B_o)

Equation 2.1 shows that the amount of CH_4 produced is directly proportional to the ultimate CH_4 yield (B_o). B_o can be determined by two methods: 1) plotting the steady-state CH_4 yield (L CH_4/g VS fed) versus the reciprocal of the retention time and extrapolating to an infinite hydraulic retention time (i.e., $1/\Theta = 0$); or 2) incubating a known amount of substrate until a negligible amount of CH_4 is produced (long-term batch fermentation). These two methods gave similar estimates of B_o for beef cattle manure fermented at temperatures ranging from 30 to 65°C at 5°C intervals (Hashimoto et al., 1979). There was no effect of temperature on B_o, and B_o averaged 0.32 ± 0.01 L CH_4/g VS fed for the steady-state method and 0.328 ± 0.022 L CH_4/g VS fed for the batch method.

For livestock manures, B_o depends on the specie, ration, the age of the manure, the collection and storage method, and the amount of foreign material (like dirt and bedding) incorporated in the manure. Table 2.1 shows some values of B_o determined for beef cattle manure (Hashimoto et al., 1979). Table 2.1 shows that the manure from cattle fed higher grain rations had greater B_o values than that from animals fed higher roughage rations. This is an expected result since rations containing higher levels of roughage would contain greater amounts of lignin complexed with cellulose. Table 2.1 also shows that chlortetracycline and monensin do not affect B_o, but 6 to 8 week old manure from a dirt feedlot has a lower B_o than fresh manure. Based upon the trends noted above, we have estimated B_o (L CH_4/g VS fed) for confined beef to be 0.35 ± 0.05, beef manure from dirt lots to be 0.25 ± 0.05; dairy manure to be 0.20 ± 0.05; and swine manure to be 0.50 ± 0.05. More studies are needed to refine these estimates and to determine other factors that affect methane yield.

2.4.3 Maximum Specific Growth Rate (μ_m)

Figure 2.2 shows the relationship between temperature and μ_m. The values of μ_m shown in Figure 2.2 were estimated by Chen and Hashimoto (1978) from data on anaerobic fermentations of sewage sludge (O'Rourke, 1968), municipal refuse (Pfeffer, 1974), dairy cattle manure (Morris, 1976; Bryant et al., 1976) and beef cattle manure (Varel et al., 1977).

Figure 2.2 shows that a straight line can be drawn between most of the data between 20 and 60°C. This relationship is described by the following equation:

$$\mu_m = 0.013\ (T) - 0.129 \tag{2.2}$$

where T is the temperature between 20 and 60°C. Temperatures above 60°C sharply decrease μ_m. The data that do not conform to Equation 2.2 are those of Pfeffer at 40 and 45°C and those of Bryant et al. and Varel et al. at 60°C. Analysis of Pfeffer's data shows a large variation in B_o and K with temperature, indicating a variation in composition of the refuse fed to the various fermentors. The high values of μ_m for the data of Bryant et al. and Varel et al. may have resulted from the limited amount of data (three relatively short hydraulic retention times: 3, 6 and 9 days) available to

TABLE 2.1. EFFECT OF MANURE TYPE AND RATION CONSTITUENTS ON ULTIMATE METHANE YIELD (B_o)[a]

	Ration, % Dry Matter			
Manure Type	Corn Silage	Corn	Antibiotic	B_o L CH_4/g VS fed
1 day old	91.5	0	none	0.173[b]
1 day old	40.0	53.4	none	0.232[c]
1 day old	7.0	87.6	none	0.290[d]
1 day old	7.0	87.6	Chlortetracycline	0.294[d]
1 day old	7.0	87.6	Monensin	0.267[d]
6-8 weeks old from dirt lot	7.0	87.6	Chlortetracycline & Monensin	0.210[b]

[a]From Hashimoto et al., 1979.

[b,c,d]Means without a common superscript differ ($P<0.05$).

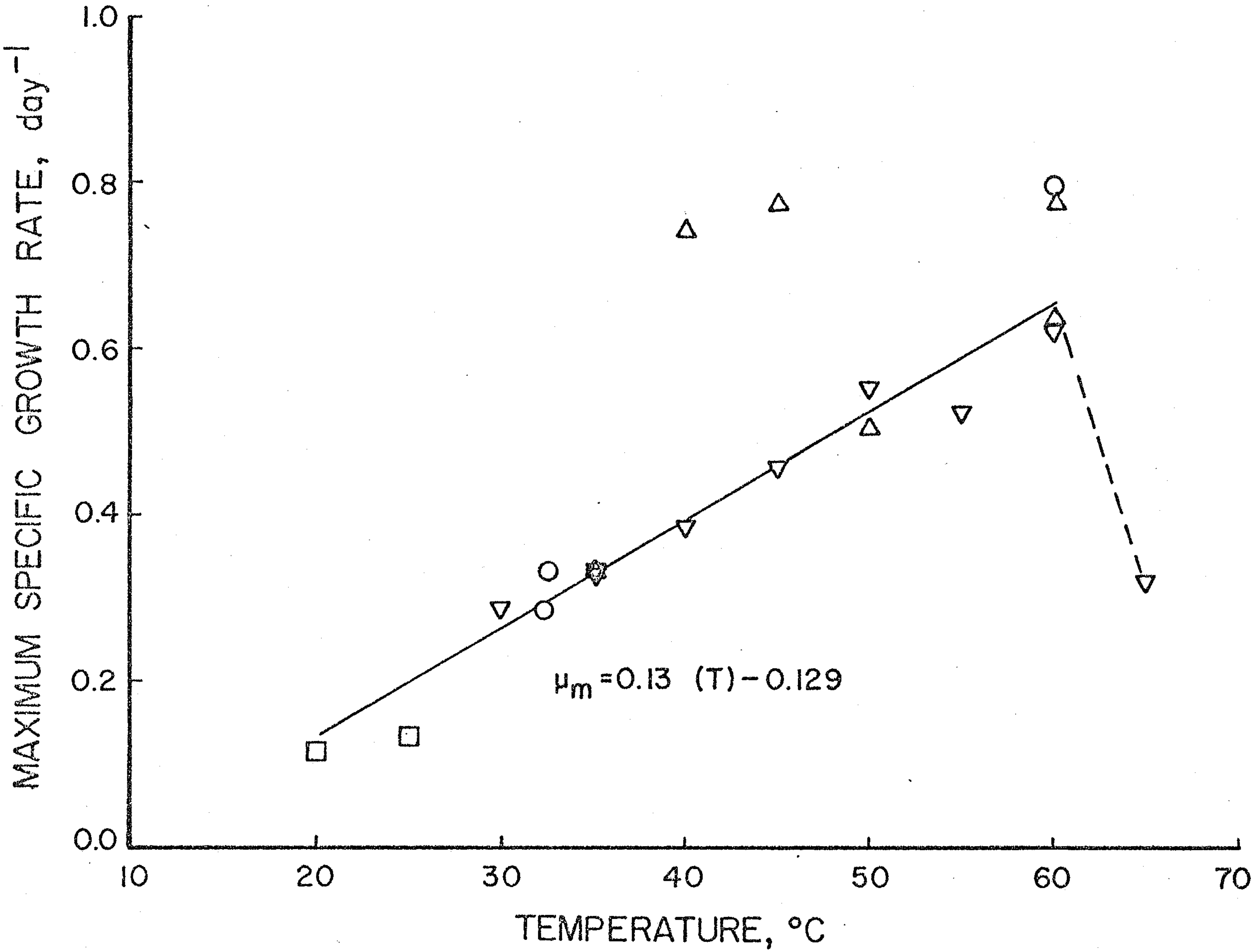

Figure 2.2. Effect of temperature on maximum specific growth rate (□ - O'Rourke, 1968; ● - Morris, 1976; ▽ - Chen et al., 1979; △ - Pfeffer, 1974; ▲ - Varel et al., 1978; ○ - Bryant et al., 1976).

estimate B_0, μ_m and K. The problems experienced in estimating these parameters are discussed elsewhere (Chen and Hashimoto, 1978).

2.4.4 Kinetic Parameter (K)

Equation 2.1 shows that when B_0, S_0, Θ, and μ_m are constant and K increases, the CH_4 production rate (γ_V) decreases. Thus, an increase in K indicates some type of inhibition has occurred. This inhibition may be caused by one or more of the following: overloading (i.e., more substrate is being added to the system than the bacteria can effectively use); inhibitory substances (e.g., volatile acids, ammonia, heavy metals, and salts) exceeding threshold levels; or reduced mass transfer of substrate, products, or both, because of the higher solids concentration.

Figure 2.3 shows the effect of influent volatile solids (VS) concentration on K for swine manure at 35°C, and cattle manure at 32.5 and 60°C. The K values for swine manure were calculated from the data of Summers and Bousfield (1980). We estimated the B_0 for their manure to be 0.36 L CH_4/g VS fed by plotting the CH_4 yield (L CH_4/g VS fed) versus $1/\Theta$ and extrapolating to an infinite Θ. This B_0 is lower than what we suggest for U.S. swine manure (0.50 L CH_4/g VS fed), which may have been caused by the diet (barley rather than corn) and the use of bedding (sawdust) to house the swine. Also, the lower VS content (70% rather than the 80 to 85% for fresh swine manure in the U.S.) of their manure suggests that some VS were destroyed before fermentation or that a larger portion of the VS in the ration was used by the swine. Both of these factors would decrease B_0.

The values for B_0 were 0.245 L CH_4/g VS fed for dairy cattle manure at 32.5°C (data of Morris, 1976), 0.169 L CH_4/g VS fed for dairy cattle manure at 60°C (data of Bryant et al., 1976), and 0.280 L CH_4/g VS fed for beef cattle manure at 60°C (data of Chen and Hashimoto, 1978).

We estimated the values for μ_m using Equation 2.2; we calculated K by substituting the cited values of B_0, μ_m, S_0 and Θ for each data set into Equation 2.1 and solving for K.

Figure 2.3 shows that K is relatively constant (about 0.6) at low S_0, but increases at different S_0 depending upon the fermentation temperature and manure type. The value of K begins increasing at 35 g VS/L for swine manure at 35°C, 40 g VS/L for cattle manure at 32.5°C and 60 g VS/L for cattle manure at 60°C. This behavior of K seems logical, because overloading a fermentor inhibits CH_4 formation, and thermophilic fermentors can sustain a higher loading rate than mesophilic fermentors before onset of inhibition (Varel et al., 1980). The effect of manure type on K may be caused by differences in ration digestible energy, differences in digestion (rumen versus monogastric) and/or the presence of inhibitory substances in the swine ration (e.g., the swine ration contained 200 ppm of copper).

Figure 2.3 should be used with caution because several data sources were used and these experiments were not planned to evaluate the kinetic parameters. A systematic study using identical apparatus and procedures is necessary to verify the preliminary results shown in Figure 2.3. Also, the presence of inhibitory substances in the manure would cause K to increase at a lower S_0 than shown in Figure 2.3.

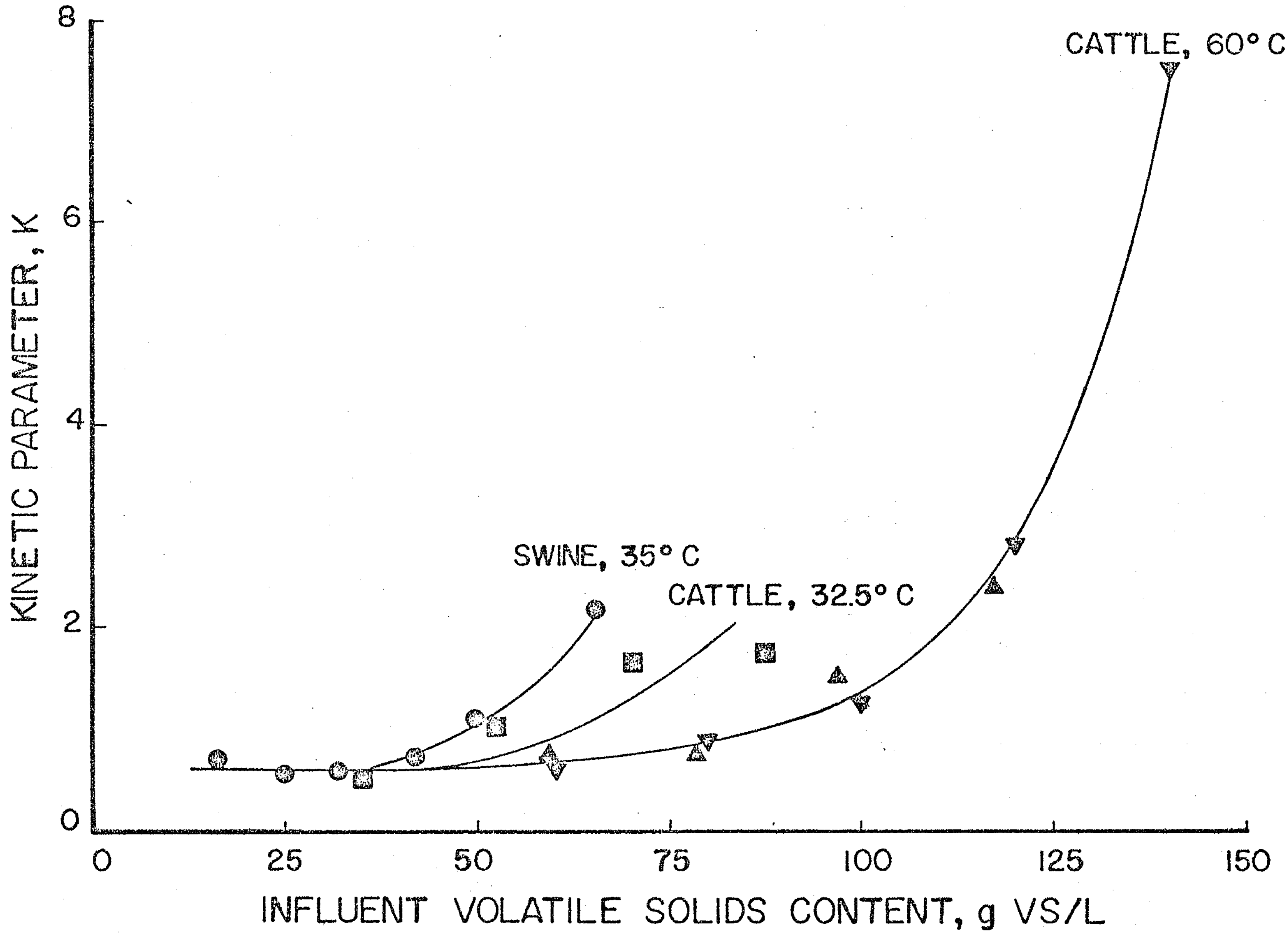

Figure 2.3. Effect of influent volatile solids content on the kinetic parameter K (● - Summers et al., 1980; ■ - Morris, 1976; ▼ - Bryant et al., 1976; ▲ - Varel et al., 1977).

2.4.5 Application of the Kinetic Model

Equation 2.1 was used to predict γ_V of various pilot- and full-scale systems fermenting livestock manures at 35, 55 and 60°C (Table 2.2). Figure 2.2 was used to estimate μ_m at each temperature and Figure 2.3 was used to estimate K at each S_o. Values for B_o were assumed to be 0.20 L CH_4/g VS fed for dairy cattle manure and 0.50 L CH_4/g VS fed for swine manure except when B_o could be calculated (the data of Summers and Bousfield, 1980).

Table 2.2 shows the experimental and predicted γ_V along with the operational and kinetic parameters used to estimate γ_V. It also shows the ratio of the predicted to experimental γ_V. Most of the predicted values are within 15% of the experimental value of γ_V except for the dairy manure fermented at 60°C. This predictive capacity is quite good, considering that μ_m and K, and B_o in most instances, were independently determined and had not been adjusted to fit the experimental data.

2.5 SUMMARY

This Section summarizes the major biological and operational factors involved in methanogenesis. A kinetic model that describes the fermentation process was presented and applied as a starting point in understanding and optimizing the fermentation process. Substrate biodegradability, fermentation temperature, and influent substrate concentration were shown to have significant effects on CH_4 production rate. The CH_4 production rates of existing pilot and full-scale fermentation systems were predicted to within 15% using this kinetic model.

TABLE 2.2. EXPERIMENTAL AND PREDICTED VOLUMETRIC METHANE PRODUCTION RATES

Specie	B_o[a] L CH_4/g VS fed	Temp °C	μ_m[a] day^{-1}	S_o[a] g VS/L	K[a]	Θ[a] day	Y_V,[a] L CH_4/L·day Exp	Pred	Ratio Pred/Exp	Source of Data
Dairy	0.20	35	0.326	64.7	1.05	10.4	0.94	0.86	0.92	Converse et al., 1977b
Dairy	0.20	60	0.651	65.2	0.60	6.2	1.41	1.76	1.25	Converse et al., 1977b
Swine	0.50	35	0.326	31.5	0.60	15	0.95	0.90	0.96	Kroeker et al., 1975
Swine	0.50	35	0.326	31.5	0.60	15	0.89	0.90	1.01	Kroeker et al., 1975
Swine	0.50	35	0.326	31.5	0.60	30	0.57	0.49	0.86	Kroeker et al., 1975
Swine	0.50	35	0.326	31.5	0.60	30	0.50	0.49	0.98	Kroeker et al., 1975
Swine	0.50	35	0.326	43.5	0.75	15	1.08	1.22	1.13	Fischer et al., 1975
Swine	0.50	35	0.326	39.2	0.70	15	1.07	1.11	1.03	Fischer et al., 1975
Swine	0.50	35	0.326	46.8	0.90	15	1.17	1.27	1.08	Fischer et al., 1975
Swine	0.50	35	0.326	60.0	1.70	15	1.36	1.39	1.02	Fischer et al., 1975
Swine	0.36	35	0.326	23.1	0.60	10	0.69	0.66	0.95	Summers et al., 1980

[a]Symbols are defined in Equation 2.1

SECTION 3.0

PILOT-SCALE THERMOPHILIC FERMENTOR OPERATION

A. G. Hashimoto

3.1 INTRODUCTION

Thermophilic anaerobic fermentation of livestock manures has several advantages that make it attractive for more detailed investigation. This system has the potential for significantly higher CH_4 production rate, with resultant savings in capital expenditures. Also, the residue is sanitized; therefore, disease transmission is minimized. This is especially important if the product is to be refed to livestock. Laboratory studies have demonstrated higher CH_4 production rates at thermophilic than mesophilic temperatures. Augenstein et al. (1976) showed about four times higher CH_4 production rates at 60°C than at 37°C for anaerobic cultures being fed CO_2 and H_2. Likewise, Pfeffer (1974) showed a four-fold increase in reaction rate at 60°C compared to 35°C for cultures fed domestic refuse. Varel et al. (1977) reported the highest CH_4 production rate (4.5 L CH_4/L fermentor·day) for beef cattle manure fermented at 60°C.

Converse et al. (1977b) compared the pilot-scale anaerobic fermentation of dairy waste at mesophilic (37°C) and thermophilic (60°C) temperatures. Their thermophilic CH_4 production rate was lower than the laboratory results reported by Varel et al. (1977) and close to those obtained by their mesophilic fermentor. They proposed the following possible explanations for the unexpectedly low gas yields of their thermophilic fermentor: insufficient mixing; wide temperature fluctuations in fermentor; less efficient microflora in their system; less biodegradable manure in their system; lower system efficiency because of improper scale-up factors.

The need for improved design criteria and scale-up factors for thermophilic, anaerobic fermentation systems is apparent. One of the objectives of this project was to determine the design factors necessary to achieve the high gas yields obtained by laboratory-scale thermophilic fermentors.

3.2 EQUIPMENT AND PROCEDURES

3.2.1 Pilot-Plant Facilities

Figure 3.1 is a schematic diagram of the pilot-scale fermentation system. The pilot-scale facilities were constructed under contract with Hamilton Standard Division of United Technologies, Inc. Manure (1 to 10 days old) was gathered daily from steers housed on partially roofed, concrete-floored pens. The steers weighed from 340 to 570 kg, depending on the season. Table 3.1 shows the rations fed to the cattle over the 1319 days of fermentor operation.

The manure was transported to the pilot plant by a small front-end loader and dumped into the slurry tank. Water was added to the material to form a slurry of 12 to 14% total solids (TS). The slurry was mixed by a 1-kW variable speed mixer. Based upon the TS and volatile solids (VS) analyses, a given amount of slurry was pumped into a 1-m^3 tank on a platform scale, the weight of the slurry transferred was recorded, and water was added to dilute the slurry to a specified VS concentration.

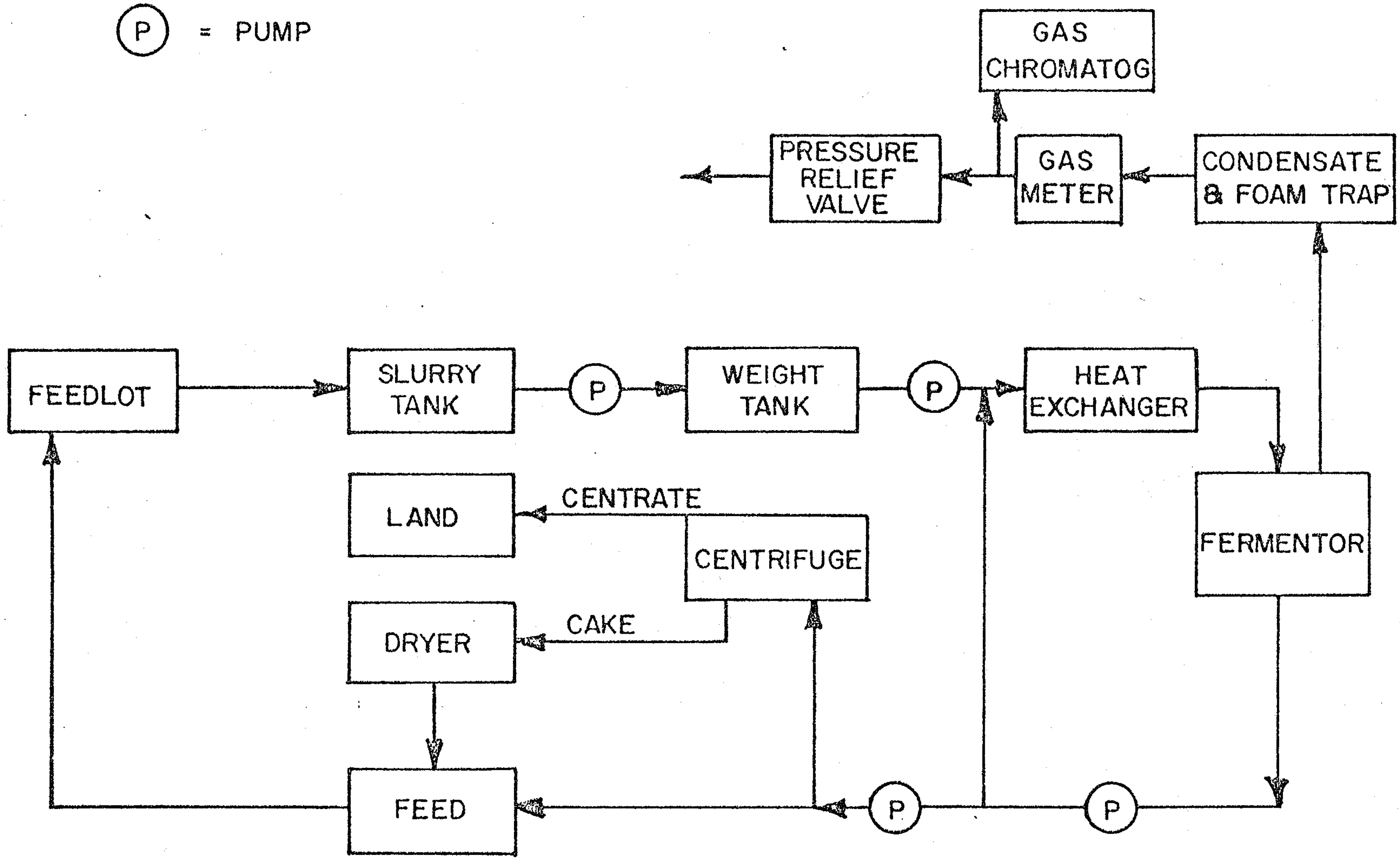

Figure 3.1. Schematic Diagram of Pilot-Scale Anaerobic Fermentation System

TABLE 3.1. BEEF CATTLE RATIONS[a] USED THROUGHOUT THE COURSE OF THE FERMENTOR OPERATION

	Day of Operation (inclusive dates)				
Item	0-125 (11/30/76-4/4/77)	126-436 (4/5/77-4/9/78)	437-463 (4/10/77-5/5/78)	464-809 (5/6/78-2/17/79)	810-1319 (2/18/79-7/11/80)
Yellow Corn	83.6	90.8	90.8	90.8	85.0
Corn Silage	4.0	6.9	6.9	6.9	13.0
Alfalfa Haylage	4.2	---	---	---	---
Soybean Meal	7.0	2.3	1.9	1.9	1.6
Limestone	0.9	---	0.3	0.3	0.2
Dicalcium Phosphate	0.1	---	0.1	0.1	0.1
Salt	---	---	0.1	0.1	0.1
Trace Minerals[b]	+	-	+	+	+
Vitamin ADE[c]	+	-	+	+	+
Chlortetracycline[d]	+	-	+	-	-

[a]Expressed on a dry matter basis

[b]9.9 g Arizona-chelated trace minerals per kg dry ration

[c]29.3 g (ADE supplement of 8.8 x 10^6 IU Vit. A/lb) per kg of dry ration

[d]10.8 g chlortetracycline (110 g chlortetracycline/kg carrier) per kg of dry ration

The slurry in the weight tank was mixed by a 0.25-kW dual-propeller mixer while the slurry was being pumped into the heat exchange loop and into the fermentor. The heat exchanger consisted of three, 6-m-long concentric tubes connected in series such that the slurry was pumped through the inner tube while hot water was pumped through the outer tube. Slurry from the fermentor was continuously pumped through the heat exchanger at 0.0032 m^3/sec.

A schematic diagram of the fermentor and mixer is shown in Figure 3.2. The fermentor volume was 5.7 m^3 with a working volume of 5.4 m^3 during the first 248 days of operation and 5.1 m^3 for the remainder of the study. The fermentor had four baffles equally spaced around the tank and the mixer consisted of a 1.5 kW variable-speed motor and two, 3-blade, stainless steel, marine propellers on a stainless steel shaft. The following geometric relationships were used: fermentor diameter (T) to propeller diameter (D) ratio, T/D = 5.6; propeller spacing of 2.5D; baffle width (W) of T/W = 14; and spacing between baffle and fermentor wall W/2.

The gas produced during the fermentation passed through condensate-foam traps, a temperature-compensated gas meter, and a pressure relief valve. The condensate-foam trap consisted of a cylindrical tank, 0.53 m in diameter and 1.73 m high, with a siphon calibrated to discharge when the pressure exceeds 0.25 m of water column. This has reduced the frequency of draining condensate from the gas flow meters and has eliminated the need to disassemble and clean the gas line after excessive foaming. The CH_4 and CO_2 concentration is measured by a gas chromatograph several times each day.

The pilot-plant facilities are housed in a 14 x 8.5 m building which also contains an office and laboratory facilities for determining solids, pH and alkalinity.

3.2.2 Methods

Before adding fresh slurry to the fermentor, a specified volume of fermented slurry, corresponding to the desired hydraulic retention time (HRT), was removed. The fermented slurry was either mixed directly with other feed ingredients for livestock feeding trials or centrifuged. The centrate flowed to a lagoon for ultimate land application, and the centrifuge cake was dried at 70°C, then used as a feed ingredient for livestock feeding trials.

Samples of slurries fed and withdrawn from the fermentor were routinely analyzed for various constituents. Total, volatile, fixed and suspended solids, ammonia (distillation method), chemical oxygen demand, alkalinity (to pH 3.7), pH, and total volatile acids (TVA, silicic acid method) were determined by the methods outlined in Standard Methods (APHA, 1975). Total Kjeldahl nitrogen was determined using Technicon block digestors and Auto-Analyzer II as described by Wael and Gehrke (1975).

Daily gas production was measured by an American AL-175 gas meter with temperature compensation capability. Gas volume was corrected to standard temperature (0°C) and pressure (1 atmosphere). CH_4 and CO_2 concentrations were measured using an on-line, Gow Mac Series 550 gas chromatograph with thermal conductivity detectors. The stainless steel column (0.64 by 183 cm) was packed with 60/80 mesh chromosorb 102. Injector, oven and detector temperatures were 102, 100 and 131°C, respectively, with a bridge current of 100 m.a. Helium carrier gas flow was 60 ml/min.

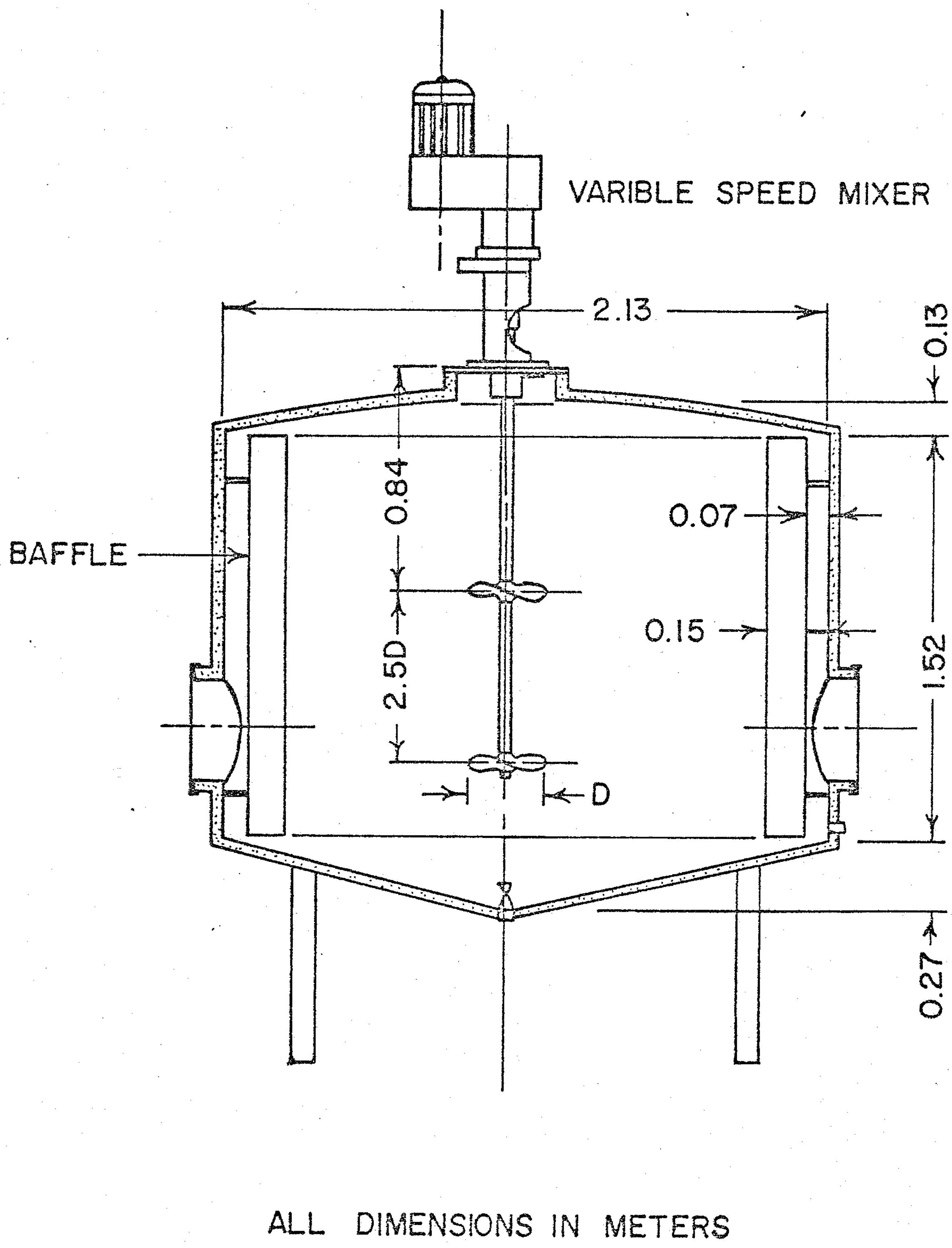

Figure 3.2. Schematic Drawing of Pilot-Scale Fermentor.

3.3 FERMENTOR OPERATION

3.3.1 Start-Up

Start-up commenced on November 30, 1976 with a charge of 50 kg of VS in 3.2 m^3 of water previously heated to 52°C. The manure charged to the system was 1 to 7 days old and contained 30% TS and 84% VS. Slaked lime (13.7 kg) was added during the first 5 days of operation to maintain the pH at 7. After day 6 of operation, the pH began to increase with a concomitant increase in gas production and decrease in TVA. Daily charging of manure began on day 9 with a loading of 1.6 kg VS/m^3 of tank contents.

Figure 3.3 shows the change in pH and accumulated gas production during start-up. After 6 days, the gas production increased dramatically. Figure 3.4 shows that the alkalinity increased during start-up and that the TVA increased to 3.5 g/L as HOAc at day 6 of operation, then steadily decreased to below 1 g/L after 10 days. Within 9 days, significant gas production was achieved. This agrees with the experiences of Varel et al. (1977) for their laboratory-scale fermentors.

The fermentor loading was gradually increased from 1.6 to 2.4 kg VS/m^3 between days 9 and 37 of operation. On day 38, the fermentor reached the desired operating volume (5.4 m^3) and daily effluent withdrawal commenced. The hydraulic retention time (HRT) was gradually decreased to 20 days on day 56. The temperature was raised to 55°C and loading of 5.4 kg VS/m^3 on day 63. The loading was decreased to 3.4 kg VS/m^3 on day 73 because the TVA began to increase.

3.3.2 Steady-State Operation

The fermentor performance was evaluated at various operating conditions to define optimum design criteria. The fermentor was operated at each condition for at least four HRT before steady-state data were recorded.

Tables 3.2, 3.3, 3.4 and 3.5 summarize the steady-state performance of the fermentor under different operating conditions. These tables show that the volumetric CH_4 production rate increases as the loading rate increases; the FS in the effluent was close to the influent FS, indicating that the fermentor contents were completely mixed; and that little nitrogen was lost during fermentation.

Table 3.2 summarizes the fermentor performance at 55°C, fed once daily and mixed continuously. It shows that the CH_4 yield (L CH_4/g VS fed (VS_f)) decreased as the HRT decreased, and that the L CH_4/g VS used (VS_u) averaged 0.54.

Table 3.3 summarizes the fermentor performance at 45 and 50°C, fed once daily and mixed continuously. The first three steady-states in Table 3.3 (S_0 = 65.3, 61.5 and 77.1 g VS/L) showed unusually low yields of 0.36, 0.39 and 0.30 L CH_4/g VS_u. These low yields prompted an intensive search for gas leaks from the fermentor and gas handling system. The search revealed a small leak around the packed bearing of the propeller shaft and a significant leak through the secondary gas-relief valve. The last steady-state in Table 3.3 (S_0 = 80.2 g VS/L) shows the fermentor performance after the gas leaks were sealed. The CH_4 yields were much higher than the three previous steady-state yields.

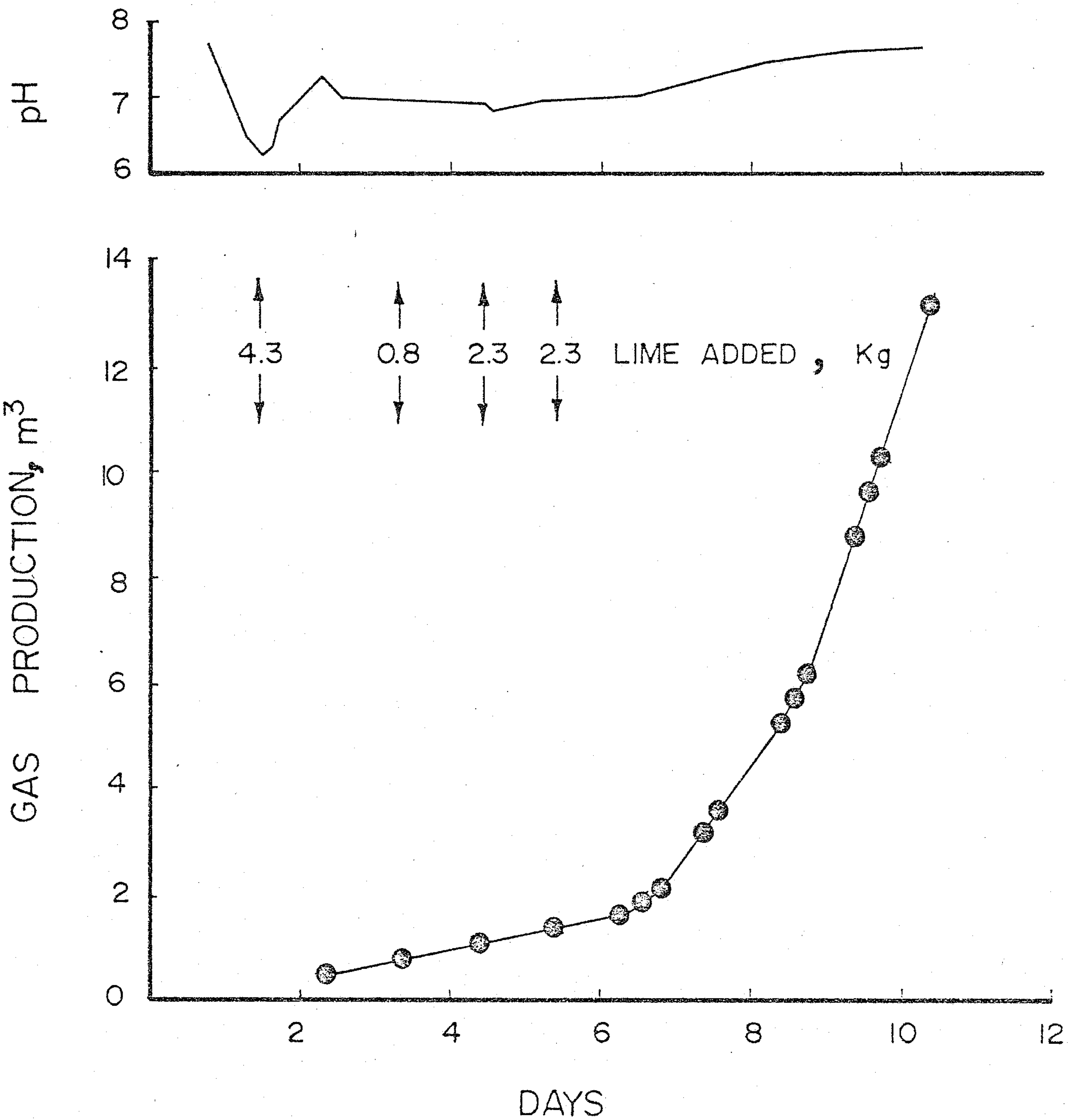

Figure 3.3. Gas Production and pH Profiles During Start-Up.

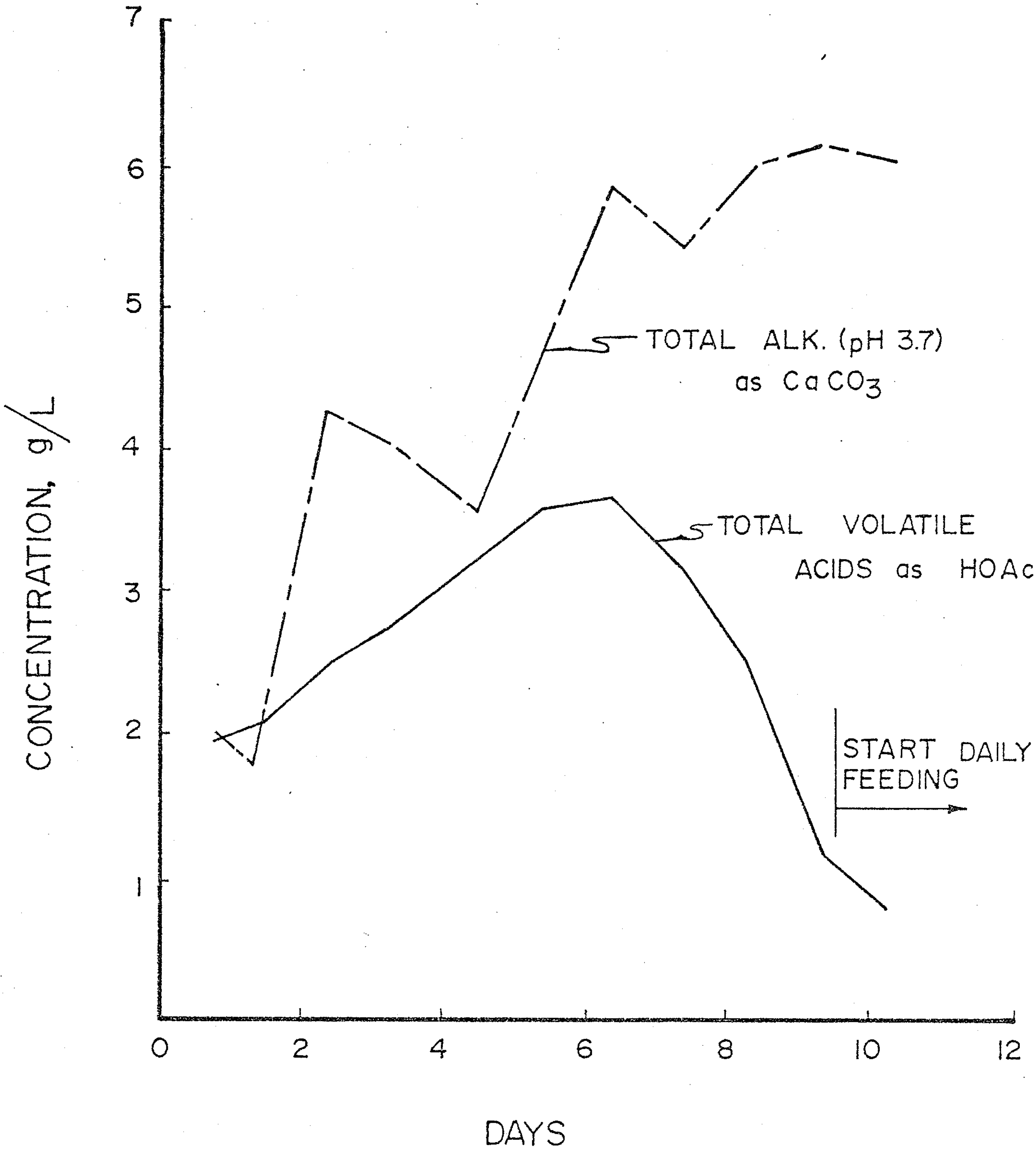

Figure 3.4. Total Alkalinity and Volatile Acids Profiles During Start-Up.

TABLE 3.2. SUMMARY OF STEADY-STATE PERFORMANCE OF THE PILOT-SCALE FERMENTOR OPERATED AT 55°C, MIXED CONTINUOUSLY, AND AT DIFFERENT HRT[a]

PARAMETER	Hydraulic Retention Time, Days			
	12	6	4	7
Total Solids				
Inf., g/L	70.1±5.4	74.4±7.3	67.7±4.7	92.4±2.7
Eff., g/L	36.6±2.7	43.0±3.8	43.8±1.7	47.0±0.8
Volatile Solids				
Inf., g/L	61.8±5.3	68.7±8.7	59.5±4.5	82.6±2.1
Eff., g/L	29.2±2.8	37.0±4.6	35.8±1.5	37.1±0.5
Fixed Solids				
Inf., g/L	8.3	5.7	8.2	9.8
Eff., g/L	7.4	6.0	7.9	9.9
COD				
Inf., g/L	74.9±13.2	73.8±3.1	73.0±3.6	93.1±11
Eff., g/L	40.2±7.1	47.2±3.5	47.8±2.4	55.0±3.6
Total Nitrogen				
Inf., g/L	4.32±0.37	3.69±0.41	3.81±0.10	4.25±0.32
Eff., g/L	3.93±0.38	3.82±0.03	4.14±0.23	4.19±0.03
Ammonia-N				
Inf., g/L	1.13±0.12	1.02±0.28	1.50±0.26	0.93±0.07
Eff., g/L	1.89±0.05	1.82±0.07	1.90±0.07	1.61±0.16
Volatile Acids				
Inf., g/L	6.95±0.79	6.75±0.74	4.56±0.85	7.85±0.66
Eff., g/L	1.15±0.23	1.82±0.21	2.55±0.19	1.27±0.07
Alkalinity				
Inf., g/L	4.06±1.18	3.26±0.54	5.43±0.45	3.73±0.48
Eff., g/L	8.59±0.42	8.53±0.55	9.23±0.11	8.26±0.51
pH				
Inf.	5.2±0.26	4.8±0.29	7.65±0.36	4.85±0.21
Eff.	7.9±0.08	7.9±0.07	7.93±0.12	7.87±0.04
Methane, %	55.0±4.9	52.1±3.0	52.2±2.1	49.9±0.7
Methane Production				
L/L·day	1.59±0.30	2.73±0.12	3.28±0.24	3.47±0.20
L/g VS_f	0.31	0.23	0.22	0.29
L/g VS_u	0.58	0.50	0.55	0.53

[a]Data presented as mean ± 1 standard deviation, steady-state assumed after 4 HRT

TABLE 3.3. SUMMARY OF STEADY-STATE PERFORMANCE OF THE PILOT-SCALE FERMENTOR OPERATED AT DIFFERENT TEMPERATURES AND HRT[a]

Parameter	Temperature/Hydraulic Retention Time			
	45°C/9d	50°C/6d	50°C/6d	50°C/6d
Total Solids				
Inf., g/L	74.8±9.4	70.1±3.9	85.1±11.2	92.0±3.1
Eff., g/L	38.5±6.8	39.5±0.5	42.3±5.9	53.8±5.6
Volatile Solids				
Inf., g/L	65.3±9.4	61.5±3.6	77.1±10.2	80.2±2.9
Eff., g/L	30.0±0.6	30.8±0.5	33.8±0.5	42.1±0.5
Fixed Solids				
Inf., g/L	9.5	8.6	8.0	11.8
Eff., g/L	8.5	8.7	8.5	11.7
COD				
Inf., g/L	72.4±4.8	73.9±3.3	76.5±10.4	94.3±5.2
Eff., g/L	42.7±2.7	42.8±1.6	43.6±5.7	56.5±5.5
Total Nitrogen				
Inf., g/L	2.68±0.10	2.81±0.16	2.97±0.38	3.44±0.21
Eff., g/L	2.80±0.06	2.98±0.10	3.20±0.07	3.66±0.08
Ammonia-N				
Inf., g/L	0.58±0.04	0.62±0.05	0.72±0.06	1.23±0.05
Eff., g/L	1.21±0.02	1.33±0.03	1.37±0.01	1.49±0.02
Volatile Acids				
Inf., g/L	6.44±0.51	6.60±0.62	8.06±0.39	6.41±1.38
Eff., g/L	1.47±0.09	0.87±0.05	1.17±0.13	1.68±0.07
Alkalinity				
Inf., g/L	2.88±0.56	2.88±0.35	3.98±1.42	5.05±0.38
Eff., g/L	6.93±0.09	7.34±0.15	7.53±0.20	10.23±0.34
pH				
Inf.	4.88±0.28	4.71±0.11	4.81±0.35	5.44±0.22
Eff.	7.61±0.04	7.76±0.07	7.78±0.10	7.91±0.06
Methane, %	52.7±3.7	58.1±1.3	53.9±1.4	59.4±0.7
Methane Production				
L/L·day	1.43±0.15	2.01±0.11	2.14±0.21	3.85±0.06
L/g VS_f	0.20	0.20	0.17	0.29
L/g VS_u	0.36	0.39	0.30	0.60

[a]Data presented as mean standard deviation, steady-state assumed after 4 HRT

TABLE 3.4. SUMMARY OF STEADY-STATE PERFORMANCE OF THE PILOT-SCALE FERMENTOR OPERATED AT 50°C, 6 DAYS HRT AND MIXED CONTINUOUSLY AND 2 HOURS PER DAY[a]

	Mixing Duration, h/d	
Parameter	24	2
Total Solids		
Inf., g/L	67.7±3.3	69.6±4.1
Eff., g/L	34.4±0.4	33.1±0.8
Volatile Solids		
Inf., g/L	59.8±3.0	61.4±3.6
Eff., g/L	26.5±0.3	25.1±0.8
Change, %	-55.7	-59.1
Fixed Solids		
Inf., g/L	7.9	8.2
Eff., g/L	7.9	8.0
COD		
Inf., g/L	68.9±3.5	70.2±6.9
Eff., g/L	34.0±4.3	34.8±5.1
Total Nitrogen		
Inf., g/L	2.42±0.17	2.61±0.24
Eff., g/L	2.65±0.06	2.54±0.03
Ammonia-N		
Inf., g/L	0.73±0.02	0.78±0.04
Eff., g/L	1.24±0.06	1.29±0.02
Volatile Acids		
Inf., g/L	5.07±0.70	6.72±0.82
Eff., g/L	0.62±0.10	0.92±0.35
Alkalinity		
Inf., g/L	3.33±0.15	3.19±0.26
Eff., g/L	6.57±0.22	6.79±0.27
pH		
Inf.	5.45±0.37	4.80±0.04
Eff.	7.50±0.04	7.51±0.05
Methane, %	52.5±0.8	53.9±4.7
Methane Production		
L/L·day	2.59±0.06	2.60±0.19
L/g VS_f	0.26	0.25
L/g VS_u	0.47	0.43

[a]Data presented as mean ± 1 standard deviation, steady-state assumed after 4 HRT

TABLE 3.5. SUMMARY OF STEADY-STATE PERFORMANCE OF THE PILOT-SCALE FERMENTOR OPERATED AT 55°C AND FED ONCE-PER-DAY OR 22 TIMES PER DAY[a]

	Hydraulic Retention Time, days (times fed per day)			
Parameter	5 (1x/day)	5 (22x/day)	5 (22x/day)	4.5(22x/day)
Total Solids				
Inf., g/L	92.8±8.9	94.7±5.9	95.0±9.9	88.8±2.6
Eff., g/L	46.4±1.9	51.6±2.0	50.2±1.6	51.2±0.7
Volatile Solids				
Inf., g/L	84.9±8.4	83.8±5.1	82.3±8.6	76.0±5.1
Eff., g/L	39.8±1.7	41.5±1.7	38.8±1.2	37.8±0.4
Fixed Solids				
Inf., g/L	7.9	10.9	12.7	12.8
Eff., g/L	6.6	10.1	11.4	13.4
COD				
Inf., g/L	93.7±11	96.1±13.1	102.2±9.7	95.9±12.1
Eff., g/L	52.9±5.2	56.9±9.2	53.0±5.3	55.3±4.6
Total Nitrogen				
Inf., g/L	3.62±0.31	4.25±0.20	3.95±0.12	4.25±0.10
Eff., g/L	3.88±0.29	4.27±0.18	4.01±0.08	4.22±0.08
Ammonia-N				
Inf., g/L	0.94±0.13	1.12±0.16	0.79±0.02	0.92±0.02
Eff., g/L	1.44±0.03	1.85±0.14	1.72±0.07	2.07±0.02
Volatile Acids				
Inf., g/L	6.89±0.34	7.70±1.14	9.04±1.01	11.42±0.42
Eff., g/L	1.64±0.12	2.39±0.33	2.12±0.85	3.34±0.08
Alkalinity				
Inf., g/L	2.95±0.59	4.37±0.25	3.24±0.24	4.35±0.46
Eff., g/L	6.12±0.32	8.63±0.63	9.50±0.48	10.08±1.31
pH				
Inf.	4.61±0.31	5.65±0.30	4.41±0.04	4.71±0.24
Eff.	7.70±0.05	7.71±0.12	7.70±0.16	7.84±0.06
Methane, %	51.9±1.4	55±2.3	56.6±0.3	57.3±0.4
Methane Production				
L/L·day	4.23±0.49	4.65±0.22	4.70±0.32	4.30±0.16
L/g VS_f	0.25	0.28	0.29	0.25
L/g VS_u	0.47	0.55	0.54	0.51

[a]Data presented as mean ± 1 standard deviation, steady-state assumed after 4 HRT

Table 3.4 summarizes the fermentor performance at 50°C, 6 days HRT, once daily feeding and mixed continuously or 2 hr/day. Table 3.4 shows that there is no difference in performance when the fermentor is mixed continuously or only 2 hr/day. Based on these results, it is difficult to justify the increased energy needed to continuously mix the fermentor when there is no apparent increase in CH_4 production rates. However, these steady-state trials were not long enough to assess the long-term effect that intermittent mixing may have on sediment accumulation in the fermentor. If intermittent mixing allows solids deposition in the fermentor, the fermentor volume would decrease. This decrease in effective fermentor volume affects important operational parameters such as HRT and loading rate. Thus, the mixing requirement for fermentation systems may be based on the materials handling and fermentor design aspects rather than maximum CH_4 production rates. More research is needed on the materials handling function of mixing systems in anaerobic fermentors.

Table 3.5 compares the fermentor performance when fed once daily or 22 times per day. At a HRT of 5 days and similar influent VS concentration, the CH_4 production rate was about 10% higher when the fermentor was fed 22 times per day compared to being fed once per day. The lower CH_4 production rate at once per day feeding may have resulted from the daily shock loading of the fermentor, especially at the short HRT of 5 days. Figure 3.5 and 3.6 show the variation in TVA, and percent CH_4 and CO_2, respectively, with time after feeding when the fermentor was operated at 55°C and HRT of 12 days. There was a 250% increase in TVA 2 hr after feeding, then a gradual decrease in TVA. The CH_4 concentration decreased to about 46% 4 hr after feeding, increased to 66% 14 hr after feeding, and remained at that concentration for the rest of the day. Figure 3.7 shows the change in hourly total gas and CH_4 production rate with time after feeding. The hourly total gas production rate was 7 times higher and the CH_4 production rate was 6 times higher 1 hr after feeding compared to 22 hr after feeding. Since the results shown in Figures 3.5 to 3.7 were for the fermentor operated at 12 days HRT and loading rate of 5.2 kg VS/m^3·day, we expect that the magnitude of a daily shock loading at 5 days HRT and 16.5 kg VS/m^3·day loading rate would be much greater, and this shock loading may be the reason for the lower CH_4 production rate for the daily fed operation compared to the 22 hr/day feeding.

3.3.3 Comparison of Experimental to Predicted CH_4 Production Rates

One of the major reasons for operating the pilot-scale fermentor was to obtain data that could be used to design full-scale systems.

In the design and scale-up of the fermentation systems, it is important to be able to predict the performance of the fermentor under different operating conditions in order to optimize the systems. Equation 2.1 was used to predict the CH_4 production rate (γ_V) of the fermentor. Figure 2.2 was used to estimate μ_m at 50 and 55°C and Figure 2.3 was used to estimate K. Since Figure 2.3 has relationships between K and S_0 only at 32.5 and 60°C, the values for K at 50 and 55°C were assumed to vary as they do at 60°C (this assumption seems to be valid based on preliminary results from our laboratory). The value of B_0 was assumed to be 0.35 L CH_4/g VS_f, since long-term (114 to 186 day) batch fermentations of the fermentor influent yielded B_0 ranging from 0.32 to 0.40 L CH_4/g VS_f.

Table 3.6 shows the experimental and predicted γ_V from the fermentor, along with the operational and kinetic parameters used to predict γ_V. The mean

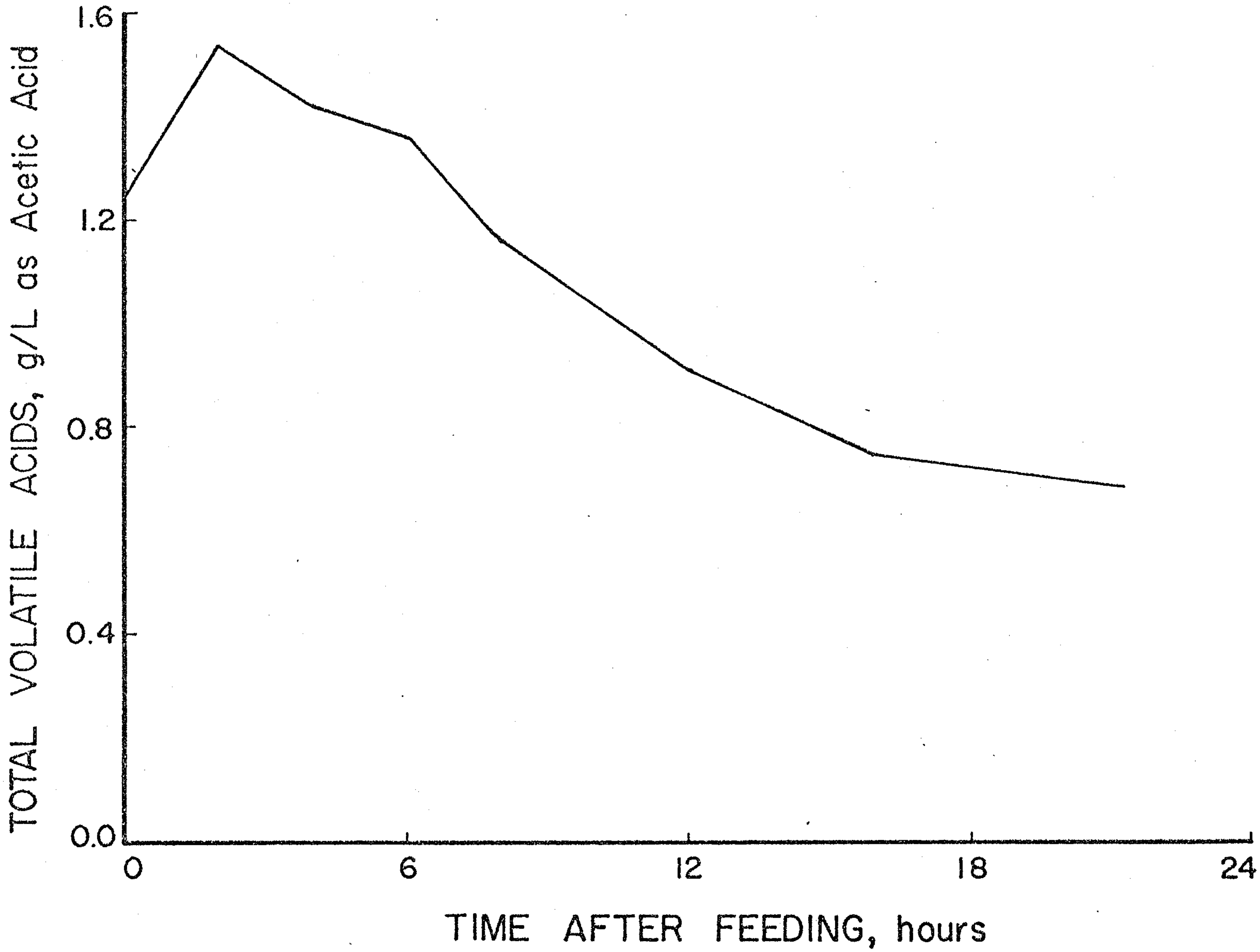

Figure 3.5. Changes in Total Volatile Acids with Time After Feeding (T = 55°C, HRT = 12 days, Loading Rate = 5.2 kg VS/m^3·day)

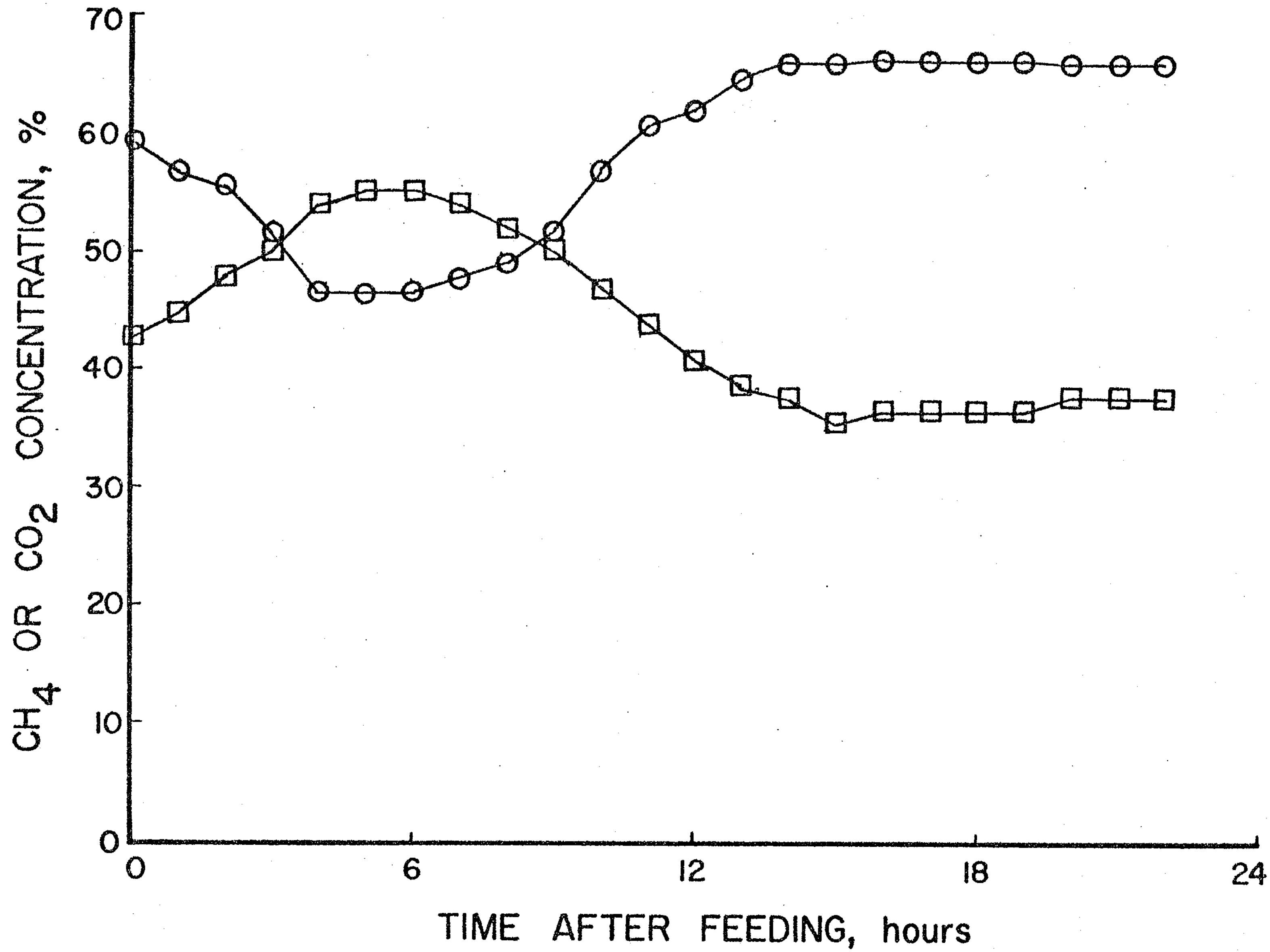

Figure 3.6. Changes in CH_4 (○) and CO_2 (□) Concentration with Time After Feeding (T = 55°C, HRT = 12 days, Loading Rate = 5.2 kg VS/m^3·day).

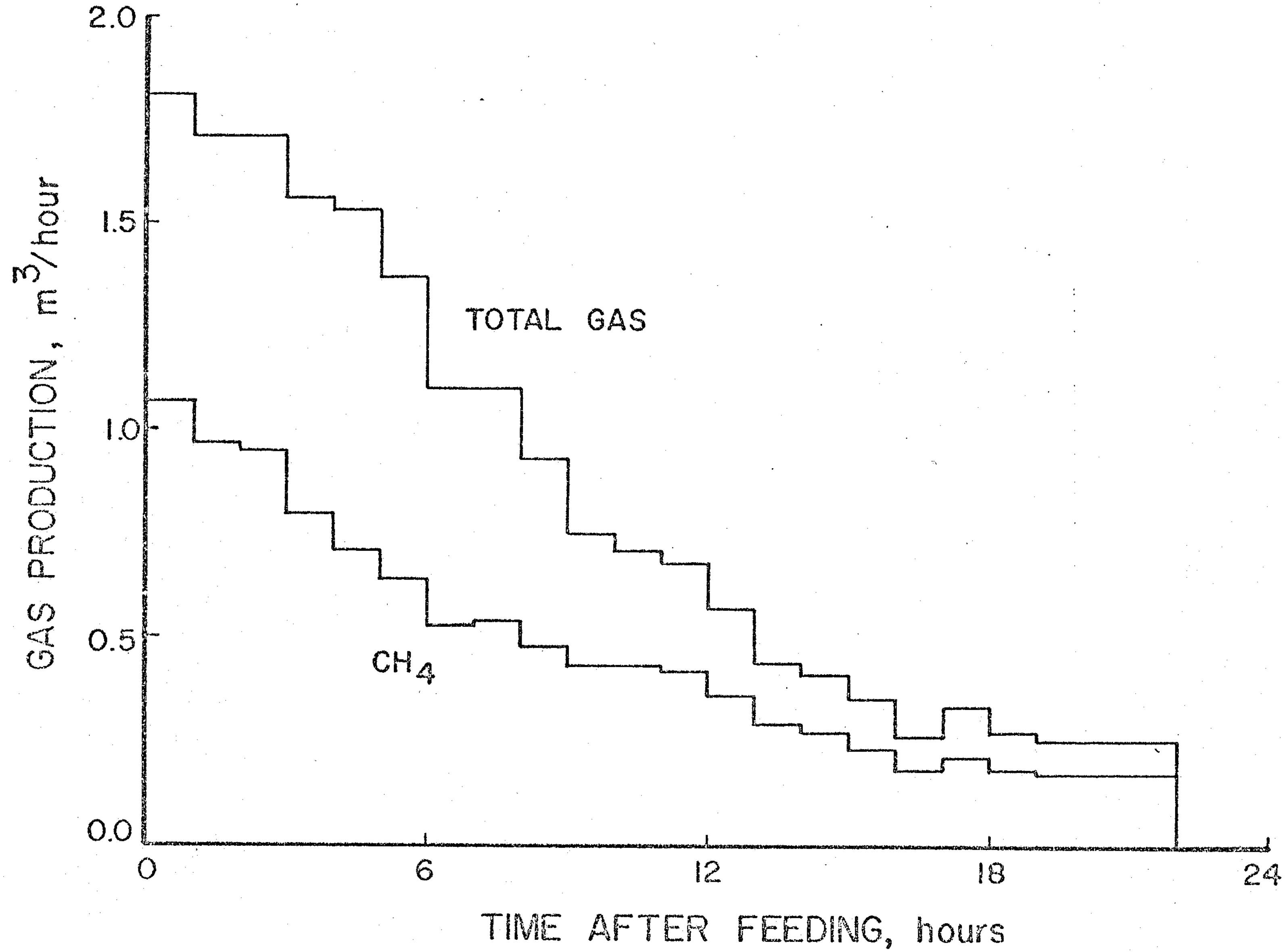

Figure 3.7. Changes in CH_4 and Total Gas Production with Time After Feeding (T = 55°C, HRT = 12 days, Loading Rate = 5.2 kg VS/m^3·day).

TABLE 3.6. EXPERIMENTAL AND PREDICTED METHANE PRODUCTION RATES[a] OF THE PILOT-SCALE FERMENTOR

Feeding x/day	Mixing h/day	Temperature °C	μ_m day^{-1}	HRT day	S_o g VS_f/L	K	Y_V, L CH_4/L·day Exp	Y_V, L CH_4/L·day Pred	Ratio Pred/Exp
1	24	55	0.586	12	61.8	0.60	1.59	1.64	1.03
1	24	55	0.586	6	68.7	0.65	2.73	3.18	1.16
1	24	55	0.586	4	59.5	0.60	3.28	3.60	1.10
1	24	55	0.586	7	82.6	0.80	3.47	3.61	0.95
1	24	50	0.521	6	80.2	0.80	3.85	3.08	0.88
1	24	50	0.521	6	59.8	0.60	2.59	2.62	1.05
1	2	50	0.521	6	61.4	0.60	2.60	2.69	1.07
1	24	55	0.586	5	85.0	0.85	4.23	3.85	0.98
22	24	55	0.586	5	84.0	0.85	4.65	3.81	0.88
22	24	55	0.586	5	82.3	0.80	4.70	3.86	0.87
22	24	55	0.586	4.5	76.0	0.70	4.30	3.89	0.96

[a]Assumes B_o = 0.35 L CH_4/g VS_f

ratio of the predicted to experimental γ_V was 0.99 with a standard deviation of ±0.10. This predictive capacity is very good, considering that K and μ_m were independently obtained, and is more than adequate for design applications.

3.4 SUMMARY

This Section summarizes the start-up and steady-state operation of the pilot-scale, thermophilic, anaerobic fermentor. The fermentor was operated at: temperatures of 45, 50 and 55°C; hydraulic retention times ranging from 12 to 4 days; mixed continuously or 2 hr/day; and fed 1 or 22 times/day. No difference in CH_4 production rate was observed when the fermentor was mixed 2 hr/day versus continuously. The CH_4 production rate was about 10% higher when the fermentor was fed 22 times/day as compared with once/day. The highest CH_4 production rate achieved by the fermentor was 4.7 L CH_4/L fermentor·day. This was the highest rate reported in the literature and about four times higher than other pilot- or full-scale systems fermenting livestock manure.

SECTION 4.0

ENERGY REQUIREMENTS FOR ANAEROBIC FERMENTATION SYSTEMS

Y. R. Chen and A. G. Hashimoto

4.1 INTRODUCTION

This section discusses the power and energy requirements for mixing, pumping, and heating the influent slurry and fermentor liquor. This discussion is necessary in order to maximize the net energy production of anaerobic fermentation systems.

4.2 ENERGY AND POWER REQUIREMENTS

4.2.1 Heating Requirement

The total heat required to maintain the fermentor liquor at a desired temperature can be expressed as follows:

$$Q_T = Q_f + Q_w + Q_g + Q_i - Q_r \tag{4.1}$$

where: Q_T = total fermentor heat requirement, J/day;

Q_f = heat loss through fermentor walls, floor and top, J/day;

Q_w = heat loss due to evaporation, J/day;

Q_g = heat loss due to the gas leaving fermentor, J/day;

Q_i = heat required to raise the influent slurry to the desired fermentor temperature, J/day; and

Q_r = heat of reaction from methane fermentation, J/day.

The heat loss through the fermentor walls (Q_f) is the sum of the heat loss through the top, side walls, and bottom of the fermentor, which can be calculated from the overall heat transfer coefficients of top, side walls and bottom of the fermentor.

The heat loss due to evaporated water (Q_w) is the sum of the sensible heat loss of the steam and the heat of evaporation of water. The sensible heat loss with the dry biogas leaving the fermentor (Q_g) is the sum of the sensible heat in CH_4 and CO_2. Ashare et al. (1978) have discussed in detail the calculation of Q_g and Q_w.

The heat required to raise the influent slurry to the fermentor operating temperature can be calculated from:

$$Q_i = W\, C_p\, (t - ts) \tag{4.2}$$

where t is the fermentation temperature (°C), ts is the influent slurry temperature (°C), W is the total weight of slurry to be added to the fermentor per day, and C_p is the specific heat of the influent slurry. The specific heat of the influent slurry depends on its total solids concentration. We have calculated the specific heat of beef cattle manure slurry to be:

$$C_p = 4.17 [1 - 0.00812 (TS)] \tag{4.3}$$

where C_p is in KJ/Kg·°C and TS is the total solids concentration in %.

Pirt (1978) suggested that in an anaerobic process, 3% of the available heat is liberated in the reaction. However, using our experimental data from 6 days HRT and a volatile solids loading rate of 15 kg/m^3·day, we found that 102 MJ/m^3·fermentor·day of heat energy was available from the influent fed to the 5.1 m^3 fermentor. The fermentor, however, produced 122 moles of CH_4/m^3·fermentor·day, which contained a heat energy of 107.7 MJ/m^3·fermentor·day. Since the heat energy in the CH_4 is essentially equal to the substrate heat energy, we concluded that the heat of reaction (Q_r) was negligible.

In the following calculations, fermentors with total working volume up to 785 m^3 were assumed to have working tank height to diameter ratios of 1.0, thus limiting tank height to 10 m. Tanks larger than 785 m^3 were designed with a maximum tank height of 10 m and sufficient diameter to accomodate the volume. The maximum tank diameter was assumed to be 80 m, resulting in a maximum tank volume of 5027 m^3. Systems requiring volumes greater than 5027 m^3 were designed with multiple tanks. The top, sides, and bottom of each fermentor were assumed to be insulated with materials having an overall heat transfer coefficient of 2.04 KJ/h·m^2·°C.

Table 4.1 gives the thermal energy requirements for fermentation systems operating at 55°C, 5 days HRT and 80 g VS/L influent concentration along with their gross methane energy production. Using B_o = 0.35 L CH_4/g VS for beef cattle manure, maximum specific growth rate (μ_m) of 0.586 day^{-1} (for 55°C) and kinetic parameter (K) of 0.8, the volumetric methane production rate (γ_v) of 3.96 m^3 CH_4/m^3·fermentor·day is obtained (Hashimoto et al., 1980).

Table 4.1 shows that for plant sizes ranging from 1 to 1,000 Mg TS/day, Q_f increases from 0.186 to 36.2 GJ/day; $Q_w + Q_g$ increases from 0.089 to 89.0 GJ/day; and Q_i increases from 1.910 to 1,910 GJ/day. The total heat energy requirement, however, decreases from 39.7% to 37.0% of the gross methane energy production, assuming a boiler efficiency of 70% and an ambient temperature of 10°C. Of the total heating requirement, 87.4% to 93.9% is for heating the influent, while heat loss through the fermentor walls accounts for 1.8% to 8.5% of the total heat requirement. The percent of heat required to compensate for the surface heat loss varies inversely with the size of the fermentor.

The net thermal energy production, i.e., the amount of CH_4 energy production minus heat energy requirement, ranges from 4.73 GJ/day for the 1 Mg TS/day plant to 4,950 GJ/day for the 1,000 Mg TS/day plant.

4.2.2 Pumping Power and Energy Requirements

The rheological properties of the slurry being pumped and mixed have a direct influence on the power requirements. Livestock waste slurries and fermentor liquor generally display non-Newtonian, pseudoplastic behavior. We previously used a power-law formula to describe the relationship between shear stress (τ)

TABLE 4.1. HEATING ENERGY REQUIREMENTS AND NET THERMAL ENERGY PRODUCTION FOR FERMENTORS OPERATING AT 55°C[a]

PARAMETER	PLANT SIZE (Mg TS/day) 1	10	100	1000
Each Fermentor Volume (m^3)	53.2	532	2660	13300
Number of Tanks	1	1	2	4
Gross Thermal Energy Production (GJ/day)	7.85	78.5	785	7850
Fermentor Surface Heat Loss (GJ/day)	0.186	0.862	5.19	36.2
Heat Loss Through Gas Line (GJ/day)	0.089	0.890	8.90	89.0
Heating Influent (GJ/day)	1.910	19.10	191.0	1910
Total Heat Loss (GJ/day)	2.185	20.85	205.1	2035
Heat Required (GJ/day)	3.12	29.79	293.0	2907
Net Thermal Energy Production (GJ/day)	4.73	48.8	493	4950

[a]Influent TS = 94 g/L; influent VS = 80 g/L; % CH_4 = 50; influent slurry temperature 10°C
Ambient temperature 10°C, Overall heat transfer coefficient = 2.04 $KJ/h \cdot m^2 \cdot °C$
Y_V = 3.96 m^3 CH_4/m^3 fermentor·day, HRT = 5 days.

and shear rate ($\dot{\gamma}$) (Chen and Hashimoto, 1976, 1979):

$$\tau = K\dot{\gamma}^n \qquad (4.4)$$

where K is rheological consistency index in $Pa \cdot s^n$ and n is rheological behavior index. The K and n of beef cattle manure slurries at different total solids concentration have been reported earlier (Chen and Hashimoto, 1979).

The method of calculating the pumping power requirement for a pseudoplastic slurry was described previously (Chen and Hashimoto, 1976). We have found that the onset of turbulence in pumping livestock waste slurry was delayed until the Generalized Reynolds number (N_{Re}') for livestock waste slurry was over 3,100. The Generalized Reynolds number is defined by:

$$N_{Re}' = \frac{\rho DV}{K} \left(\frac{8V}{D}\right)^{(1-n)} \left[\frac{4n}{3n + 1}\right]^n \qquad (4.5)$$

where ρ = slurry density, kg/m^3;

D = pipe diameter, m;

V = slurry flow speed, m/sec/

To prevent solid particles from settling in the pipe and to have better heat transfer characteristics when a heat exchanger is used to recover effluent heat, the piping should be designed to maintain turbulent flow ($N_{Re} > 4{,}300$). In our calculation, however, the pipe size is chosen so that the N_{Re}' is close to but does not exceed 5,000, and the pipe size is no smaller than 0.0191 m ID.

Table 4.2 lists the pipe diameter, pumping rate, number of pumps and the total influent and effluent volume to be pumped. The plants were assumed to operate at 5 days HRT and 80 g VS/L influent concentration (9.4% TS concentration). The effective pumping length was assumed to be 300 m, which does not include the pressure head due to the liquid height of the above-ground tank. The same pump used to pump the influent was also used to pump the effluent. Effluent pumping was not necessary for the 1000 Mg TS/d plant because there was sufficient head to use gravity flow.

The rheological properties were assumed to be: K = 0.61 $Pa \cdot s^n$ and n = 0.54 for the 10% TS influent; and K = 0.33 $Pa \cdot s^n$ and n = 0.5 for the 5% TS effluent.

Table 4.2 shows that the power required to pump the influent is 0.70 kW for the 1 Mg TS/day plant and 95.4 kW for the 1,000 Mg TS/day plant. The power requirement per unit volume of slurry pumped decreases from 65.8 W/m^3 for the 1 Mg TS/day plant to 8.97 W/m^3 for the 1,000 Mg TS/day plant. Because of the low viscosity of the fermentor liquor and use of gravity flow in the large plant, the power required to pump the effluent ranged from 3.55 to 0 kW for plants ranging from 1 to 1,000 Mg TS/day. The total energy required to pump the influent and effluent increased from 0.0345 GJ/day for the 1 Mg TS/day plant to 3.44 GJ/day for the 1,000 Mg TS/day plant.

Table 4.3 shows that if the pumping time for the influent slurry and the

TABLE 4.2. POWER AND ENERGY REQUIREMENTS FOR PUMPING EFFLUENT AND PROCESS SLURRIES[a]

PARAMETER	PLANT SIZE (Mg TS/day) 1	10	100	1000
Volume of Slurry to be Pumped per Hour (m^3/hr)	1.064	10.64	106.6	1064
Pipe Diameter (m)	0.0191	0.0381	0.1016	0.2286
Generalized Reynolds Number for Influent Pumping	495	2740	2783	4234
Number of Pumps	1	1	2	4
Influent Power (kW)[b]	0.70	4.31	12.90	95.4
Power/Volume (W/m^3)	65.8	40.5	12.12	8.97
Effluent Power (kW)[b]	0.26	3.55	1.61	0
Power/Volume (W/m^3)	24.2	33.4	1.513	0
Total Energy Required (GJ/day)	0.0345	0.283	0.524	3.44

[a] 10 hours pumping. Influent assumed: 10% TS, K = 0.61 Pa·s^n, n = 0.54; Effluent assumed: 5% TS, K = 0.33 Pa·s^n, n = 0.50. Effective length 300 m including the effective length due to suction, expansion, contraction of flow.

[b] Pump efficiency = 50% assumed.

TABLE 4.3. POWER AND ENERGY REQUIREMENTS FOR PUMPING EFFLUENT AND PROCESS SLURRIES[a]

PARAMETER	PLANT SIZE (Mg TS/day)			
	1	10	100	1000
Volume of Slurry to be Pumped per Hour (m^3/hr)	3.55	35.5	355	3550
Pipe Diameter (m)	0.0254	0.0635	0.1778	0.330
Generalized Reynolds Number for Influent Pumping	1446	4711	4260	4084
Number of Pumps	1	1	2	4
Influent Power (kW)[b]	2.10	14.21	19.21	33.0
Power/Volume (W/m^3)	197.4	133.6	18.05	3.10
Effluent Power (kW)[b]	1.20	8.92	0	0
Power/Volume (W/m^3)	11.28	83.8	0	0
Total Energy Required (GJ/day)	0.0358	0.250	0.415	2.854

[a] 3 hours pumping. Influent assumed: 10% TS, K = 0.61 Pa·s^n, n = 0.54; Effluent assumed: 5% TS, K = 0.33 Pa·s^n, n = 0.50. Effective length 300 m including the effective length due to suction, expansion, contraction of flow.

[b] Pump efficiency = 50% assumed.

effluent is shortened to 3 hr/day, the pipe size and power requirement will increase. Pumping the influent requires 2.10 kW for the 1 Mg TS/day plant and 33.0 kW for the 1,000 Mg TS/day plant. However, the total energy consumption remains about the same for the 1 and 10 Mg TS/day plants, and there is a 17% to 20% reduction in energy consumption for the 100 and 1,000 Mg TS/day plants, respectively, because much larger pipes are used for 3 hr pumping compared to pumping 10 hr per day.

4.2.3 Mixing Power Requirement

The pilot-scale fermentor liquor (at about 5% TS) and the influent slurry (at about 12% TS) were agitated by mixers equipped with dual, 3-blade marine propellers. Adequate agitation was achieved at rotational speeds of 140 rpm for the fermentor liquor and 316 rpm for the influent slurry.

The net power consumption was estimated from the plots of power number (N_p) and Reynolds number (N_{Re}) for mixing beef cattle manure (Chen and Hashimoto, 1979). The net power consumption for mixing the fermentor liquor was 86.7 W for one propeller and 156.2 W for dual propellers, using a factor of 1.8 for dual propellers (Bates et al., 1966). This gives a net power consumption per volume of 28.8 W/m^3.

The net power consumption for mixing the influent slurry was estimated using the same procedure, and was calculated to be 152 W/m^3 for single propeller and 213 W/m^3 for dual propellers. A factor of 1.4 for dual propellers is used because the separation of these two propellers is only one propeller diameter (Bates et al., 1966).

To maintain the same quality of mixing in large scale fermentation systems, the power consumption per unit volume should be preserved (Johnstone and Thring, 1957).

Table 4.4 shows the power and energy requirement for mixing fermentor liquor and influent slurry for different plant sizes. With continuous mixing, the fermentor mixing energy requirement increased from 0.222 to 222.0 GJ/day for plant sizes ranging from 1 to 1,000 Mg TS/day. For mixing the influent slurry, the power and energy requirement are 48% higher than those for the fermentor liquor.

4.3 DISCUSSION

4.3.1 Comparing Energy Requirements

Table 4.5 summarizes the energy requirements for systems fermenting beef cattle manure operating at 55°C, 5 days HRT, and 80 g VS/L influent concentration. The energy requirements for CO_2 scrubbing and CH_4 compression are also listed in Table 4.5. Ashare et al. (1978) concluded that the water scrubbing of CO_2 is the simplest and cheapest way to clean the biogas. The power required was estimated to be 5.88 W/m^3/day of the biogas flow rate. The net power required to compress the methane gas from 101.3 kPa (1 atmosphere) to 861 kPa (8.5 atmosphere) was used. Assuming an ideal gas and adiabatic process, the total power required to compress the CH_4 is 4.94 W/m^3/day (Perry and Chilton, 1973) with a compressor efficiency of 70%.

Table 4.5 shows that the heating required to maintain the fermentor at 55°C

TABLE 4.4. POWER AND ENERGY REQUIREMENTS FOR PROPELLER MIXING OF FERMENTOR LIQUOR[a] AND PROCESS SLURRY[b]

PARAMETER	PLANT SIZE (Mg TS/day) 1	10	100	1000
Fermentor Liquor				
Volume (m^3)	53.2	532	5320	53200
Power Required (kW)[c]	2.567	25.67	256.7	2567
Energy Required (GJ/day)[d]	0.222	2.220	22.20	222.0
Influent Slurry				
Volume (m^3)	10.64	106.4	1064	10640
Power Required (kW)[c]	3.80	38.0	380	3800
Energy Required (GJ/day)[d]	0.329	3.29	32.9	329
Total Energy Required (GJ/day)[d]	0.551	5.51	55.1	551

[a]Fermentor Liquor: 5% TS, K = 0.33 $Pa \cdot s^n$ and n = 0.5.

[b]Influent Slurry: 10% TS, K = 0.61 $Pa \cdot s^n$, n = 0.54.

[c]Motor Efficiency 80%.

[d]24 hours mixing.

TABLE 4.5. SUMMARY OF ENERGY PRODUCTION AND REQUIREMENT FOR ANAEROBIC SYSTEMS FERMENTING BEEF CATTLE MANURE AT 55°C[a]

PARAMETER	PLANT SIZE (Mg TS/day)			
	1	10	100	1000
Gross Methane Energy Production (GJ/day)	7.85	78.5	785	7850
Heating Energy Required[b] (GJ/day)	3.12	29.79	293.0	2907
Heating Energy Required[b] w/50% Effluent Heat Recovery (GJ/day)	1.757	16.15	156.5	1543
Pumping Energy Required[c] (GJ/day)	0.0345	0.283	0.524	3.44
Mixing Energy Required[d] (GJ/day)	0.551	5.51	55.1	551
CO_2 Scrubbing (GJ/day)	0.2142	2.142	21.42	205
CH_4 Compression (GJ/day)	0.0900	0.900	9.00	90.0

[a] Influent concentration 80 g VS/L, HRT = 5 days, γ_V = 3.96 m^3 CH_4/m^3·fermentor·day.

[b] Ambient and process slurry temperature, 10°C.

[c] 10 hours of pumping, 300 m effective length.

[d] 24 hours of mixing.

comprises the major portion of the total energy consumption, ranging from 39.7% to 37.0% at an assumed ambient temperature of 10°C. Of this heating energy requirement, 87.4% to 93.9% is used to heat the influent slurry. This indicates the absolute necessity of recovering the effluent heat energy for heating the influent. The heating energy requirement is reduced from 37.0% to 19.7% of the gross energy production for a 1,000 Mg TS/day plant if 50% of the effluent heat is recovered.

The next major energy consumption is mixing. It amounts to 7.3% of the total CH_4 energy production. Our laboratory and pilot studies showed that the CH_4 production did not vary with the duration of daily mixing of the fermentor liquor, indicating that the minimum mixing requirement for fermentation systems may be based on the slurry and effluent handling aspects. The energy consumption of mixing can be cut at least one half by reducing the mixing time.

The least energy is consumed in pumping. Pumping energy depends less on the plant size because larger pipes can be used for handling larger volumes of the slurry. Energy consumed per volume pumped decreases as pipe size increases. Three hours pumping has higher Reynolds number but requires larger pumps and pipes than 10 hours pumping and therefore involves higher capital cost (Tables 4.2 and 4.3).

The total energy consumption, including CO_2 scrubbing and CH_4 compression, but excluding heating energy, accounts for 11.3% of the gross CH_4 energy production for the 1 Mg TS/day plant and 10.8% for the 1,000 Mg TS/day plant.

4.3.2 Effect of Influent Concentration on Net Thermal Energy Production

Equation 2.1 and Figures 2.2 and 2.3 were used to calculate the gross CH_4 production from which the heating requirement was subtracted to determine the net thermal energy production. In calculating the CH_4 production at 55°C, $_m$ of 0.586 day^{-1} and K = 0.6, 0.65, 0.8, and 1.3 for S_0 = 60, 70, 80, and 100 g VS/L, respectively, were used. Figure 4.1 shows that at long HRT, higher influent concentration will produce more net thermal energy for the same plant size. However, at HRT less than 12 days, the net thermal energy production increases with S_0 only if S_0 is less than 80 g VS/L. As S_0 increases to 100 g VS/L, the net thermal energy production has a decreases sharply.

4.3.3 Net Thermal Energy Production of Mesophilic and Thermophilic Systems

Figure 4.2 compares the net thermal energy production from systems fermenting beef cattle manure at mesophilic (35°C) and thermophilic (55°C) temperatures with an influent concentration of 80 g VS/L. At 35°C and S_0 = 80 g VS/L, μ_m = 0.326 day^{-1} and K = 1.7 were used.

Figure 4.2 shows that, at long HRT, mesophilic systems may produce more net thermal energy than thermophilic systems, but at short HRT, thermophilic systems will produce more. The advantage of thermophilic over mesophilic systems will increase if the effluent heat is recovered for influent slurry heating.

4.4 SUMMARY

This section discussed the energy requirements for anaerobic fermentation

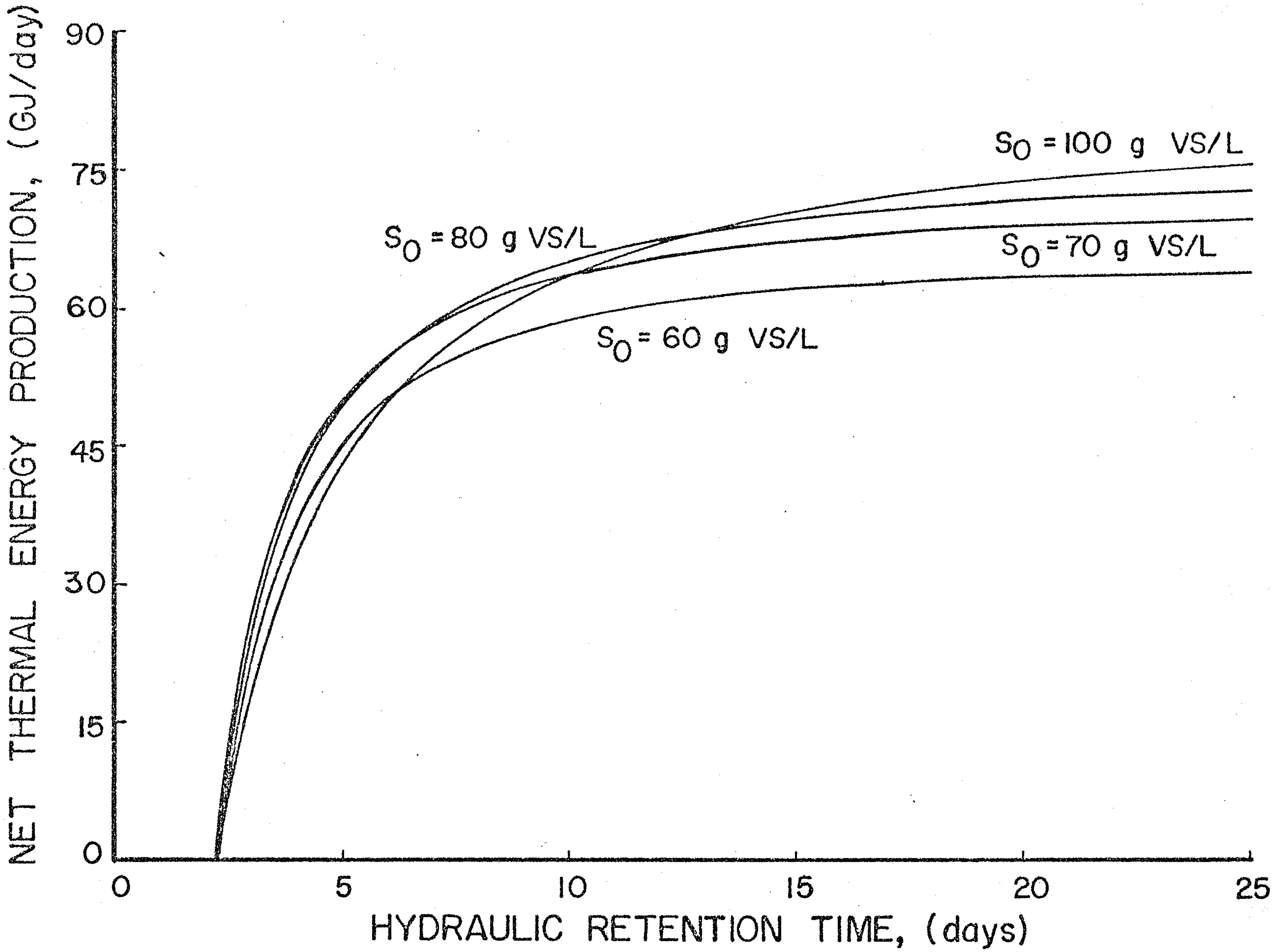

Figure 4.1. Comparing net thermal energy production from thermophilic anaerobic fermentation systems for different influent VS concentration (fermentation temperature 55°C, plant size 10 Mg TS/day, ambient temperature 10°C, and B_O = 0.35 L CH_4/g VS).

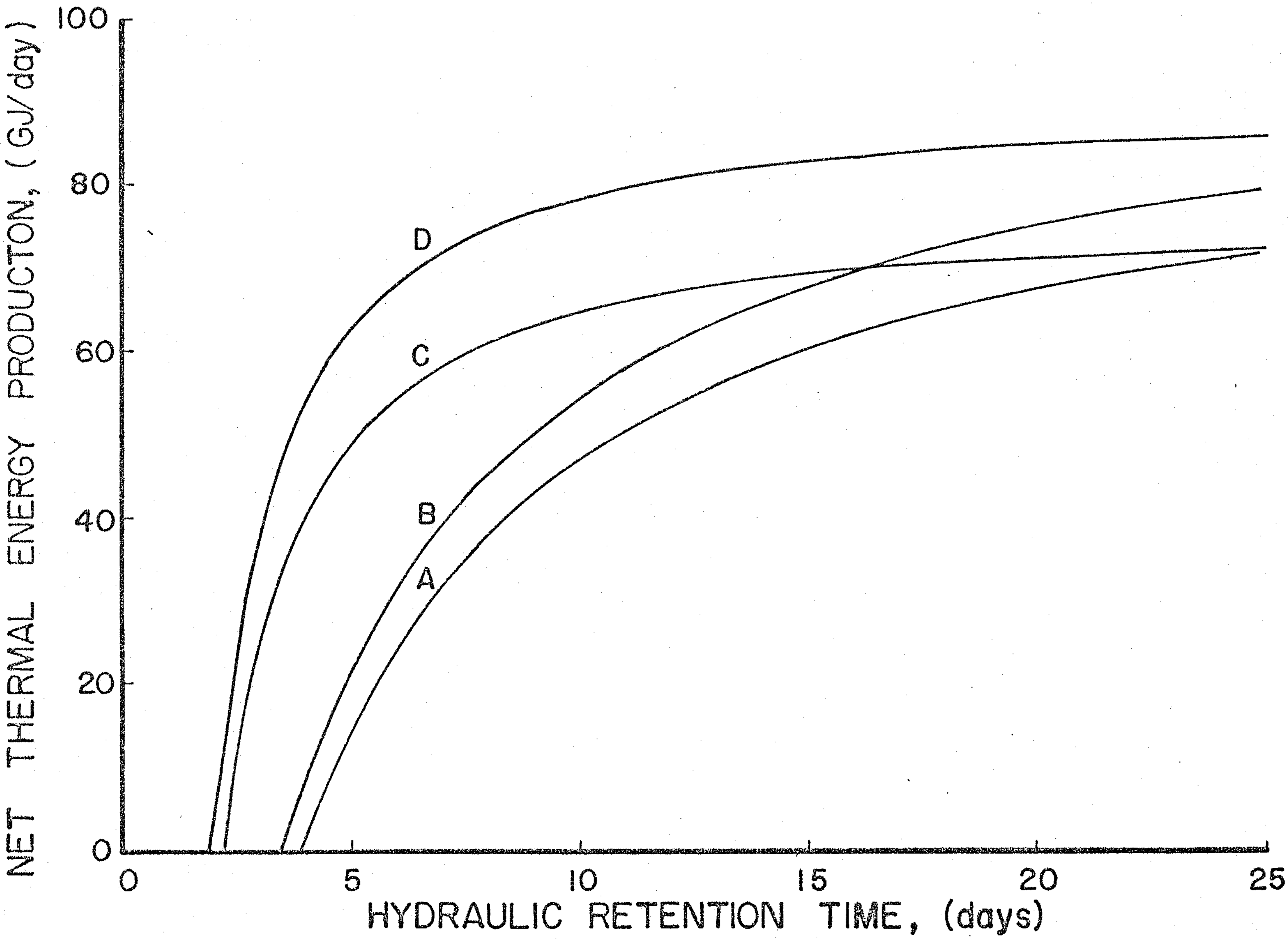

Figure 4.2. Comparing net thermal energy production from mesophilic (35°C) and thermophilic (55°C) anaerobic systems (plant size 10 Mg TS/day, ambient temperature 10°C, S_o = 80 g VS/L, and B_o = 0.35 L CH_4/g VS. A: 35°C without effluent heat recovery, B: 35°C with 50% heat recovery, C: 55°C without effluent heat recovery, D: 55°C with 50% heat effluent heat recovery).

systems. The major energy consumption for a thermophilic system is in maintaining the fermentor temperature. Of the total heating energy required, about 89 to 94% was for heating the influent slurry at an ambient temperature of 10°C. The need to recover the heat leaving with the effluent is apparent. With 50% effluent heat recovery, the heating energy requirement is reduced from 37.0% to 19.7% of the gross energy production for a 1,000 Mg TS/day plant.

The next major energy consumption was due to the mixing of the influent slurry and fermentor liquor. Mixing amounted to 7.3% of the gross methane energy production, assuming continuous mixing. The mixing energy can be reduced greatly if the mixing time is reduced.

The least energy was consumed in pumping. Pumping energy did not increase when the pumping time was shortened from 10 to 3 hours. Three hours of pumping has a higher Reynolds number but requires a larger pump and bigger pipes.

The total energy consumption excluding thermal energy consumption accounts 10.8 to 11.3% of the gross thermal energy production.

Using the kinetic constants given for different influent concentration and fermentation temperature, it was found that for HRT less than 12 days, the net thermal energy production increased with the influent concentration up to 80 g VS/L and began to drop at influent concentration greater than 80 g VS/L. Also, a longer HRT will produce more net thermal energy at the same influent concentration.

The net thermal energy production for 35° and 55°C were also compared for an influent concentration of 80 g VS/L. This comparison showed that a thermophilic (55°C) system will produce more thermal energy than a mesophilic (35°) system unless the fermenter is operated at a very long HRT.

SECTION 5.0

ECONOMIC OPTIMIZATION OF ANAEROBIC FERMENTOR DESIGNS

A. G. Hashimoto and Y. R. Chen

5.1 INTRODUCTION

It is apparent from this report and reports published elsewhere that producing methane (CH_4) from livestock manures is technically, and sometimes economically, feasible. The maximum amount of CH_4 per unit weight of substrate is obtained at long hydraulic retention times (HRT); however, long HRT require very large fermentor volumes and high capital costs. This section presents our approach to optimizing fermentor designs, based on maximizing the net energy production per unit fermentor cost.

5.2 OPTIMIZED DESIGNS

5.2.1 Capital Cost

Figure 5.1 shows the capital cost for various anaerobic fermentation systems plotted against the fermentor volume (Ashare et al., 1977; Burford et al., 1977; Coppinger et al., 1979; Fischer et al., 1978; Hashimoto and Chen, 1979; and Hayes et al., 1979). The capital cost included all equipment and facility costs for the fermentation system (including installation labor), except costs for effluent treatment or storage (e.g., centrifugation, filtration, lagooning) and biogas handling or use (e.g., CO_2 scrubbing and electrical generation).

Several important relationships should be noted from Figure 5.1. For fermentors larger than 100 m^3, there are two apparent cost-volume relationships (high and low capital costs), and both relationships show that capital cost increases with volume to the 0.7 power. Also, the high capital cost systems cost three times more than the low capital cost systems, and the capital costs for the plug-flow systems, reported by Hayes et al. (1979), resembles that for the same size conventional fermentor. The high capital cost systems represent "turn-key" systems that are designed, constructed and started-up by private contracting firms. The low capital cost systems represent farmer-contracted systems with partial construction labor provided by farm personnel.

5.2.2 Net Energy Production Per Unit Cost

The method we selected to optimize the design of anaerobic fermentation systems is to maximize the net thermal energy production per unit of capital cost. We calculated the gross energy production by using the kinetic equation and parameters presented in Section 2.0. The net thermal energy production was calculated by subtracting the heating requirements of the system from the gross energy production as described in Section 3.0.

Since the cost of the fermentation system is related to the fermentor volume to the 0.7 power (Figure 5.1), the net thermal energy production per unit of relative capital cost (NEPC) was calculated as follows:

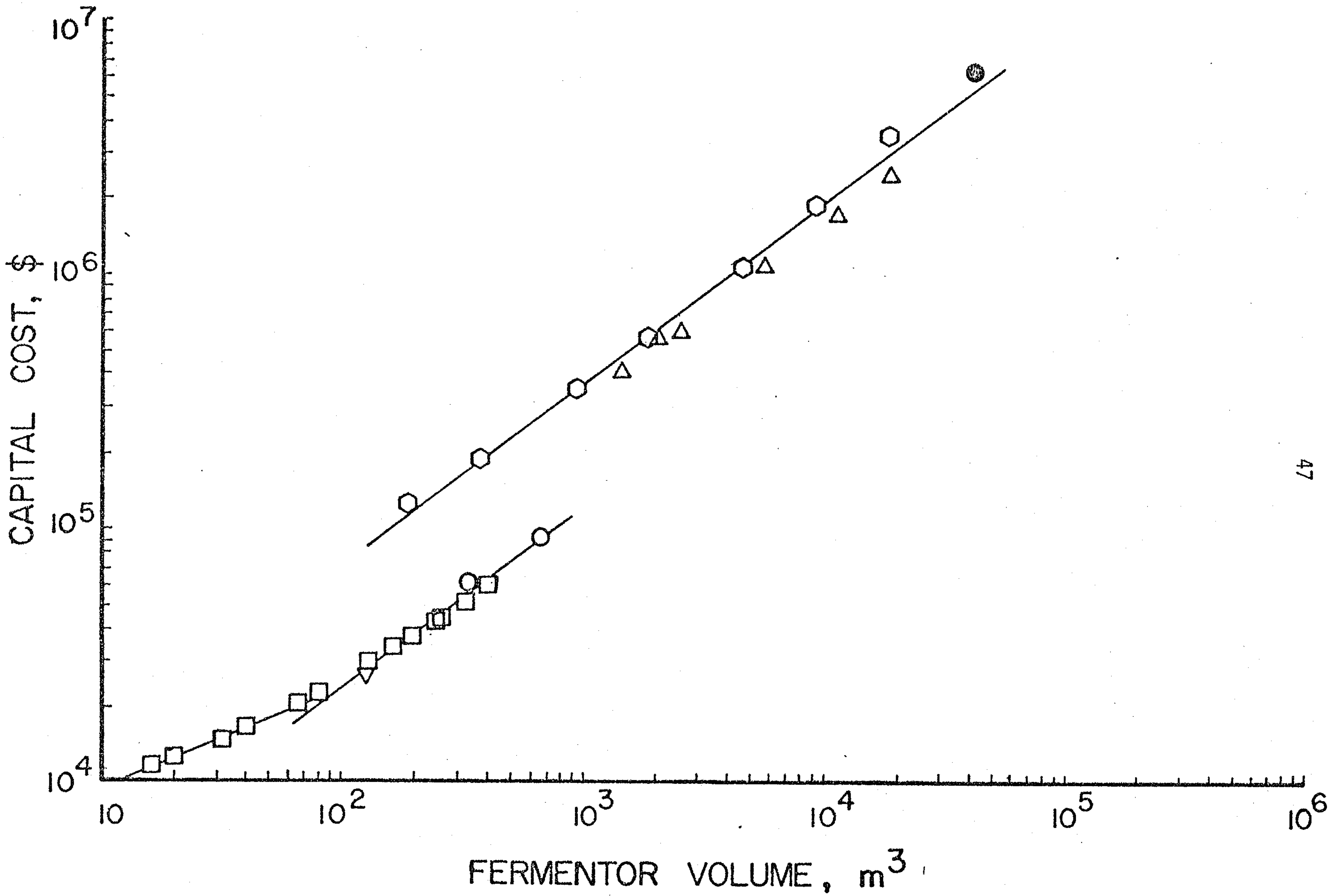

Figure 5.1. Effect of Fermentor Volume on Capital Cost (⬡, Ashare et al., 1977; ●, Burford et al., 1977; ○, Coppinger et al., 1979; ▽, Fischer et al., 1978; △, Hashimoto et al., 1979; and □, Hayes et al., 1979).

$$NEPC = \frac{\text{gross energy production - thermal energy requirement}}{(V/1000)^{0.7}} \quad (5.1)$$

Thus, the design conditions would be optimum when NEPC is maximum.

Figure 5.2 presents plots of NEPC versus HRT at different temperatures (35° and 55°C) and different influent volatile solids (VS) concentrations (50, 60, 70, 80, 90, and 100 g VS/L), assuming that the ambient and influent temperatures are 10°C and that 50% of the effluent heat is recovered. Figure 5.2 shows that the design conditions are optimum at 55°C, with HRT between 4 and 5 days and influent concentrations between 80 and 100 g VS/L. Under these conditions, the thermophilic system had about twice the NEPC of the mesophilic systems.

5.3 ECONOMICS

5.3.1 System Design

Using the optimum design conditions determined in the preceeding subsection, we made an economic assessment to determine the economic feasibility of using anaerobic fermentation systems in beef cattle enterprises. Beef cattle enterprises were selected because more information is available on the kinetic parameters of cattle manure fermentation systems than other livestock manures (Section 2.0), and because we have experience in operating a pilot-scale (5 m^3) thermophilic, anaerobic fermentor close to these optimum design conditions (Section 3.0). Based on the discussion in Section 3.0, we used the following kinetic parameters: $\mu_m = 0.586$ day^{-1} at 55°C and K = 1.0 at an influent concentration of 90 g VS/L. These kinetic parameters yield a CH_4 production rate of 4.15 L CH_4/L fermentor·day assuming a B_0 of 0.35 L CH_4/g VS. This CH_4 production rate is achievable since we have obtained a rate of 4.7 L CH_4/L fermentor·day in our pilot-scale fermentor. We assumed a CH_4 concentration of 55% in the biogas.

In the proposed fermentation system, manure from a confinement feedlot is scraped into a mixing tank where water is added to produce a slurry of 90 g VS/L (assuming VS = 85% of TS). The mixing tank is equipped with a mechanical mixer, degritting mechanism, and piping to heat the slurry to the desired temperature. The slurry is then pumped to the fermentor, which is mixed for 12 hr/day. The effluent is pumped through the piping in the mixing tank to recover about 50% of the heat. The effluent is then incorporated with other ration ingredients and fed back to the cattle. Supplemental heating of the influent is provided by a hot water heat-exchanger. Operation of the entire system is controlled by a microprocessor-controller.

We evaluated several options for biogas handling (with and without CO_2 removal) and use (heat or electricity) in this assessment, as follows:

Option	Energy Use	CO_2 Removed
A	On-site Heating	No
B	Sale as Natural Gas	Yes
C	Electricity	No
D	Electricity	Yes

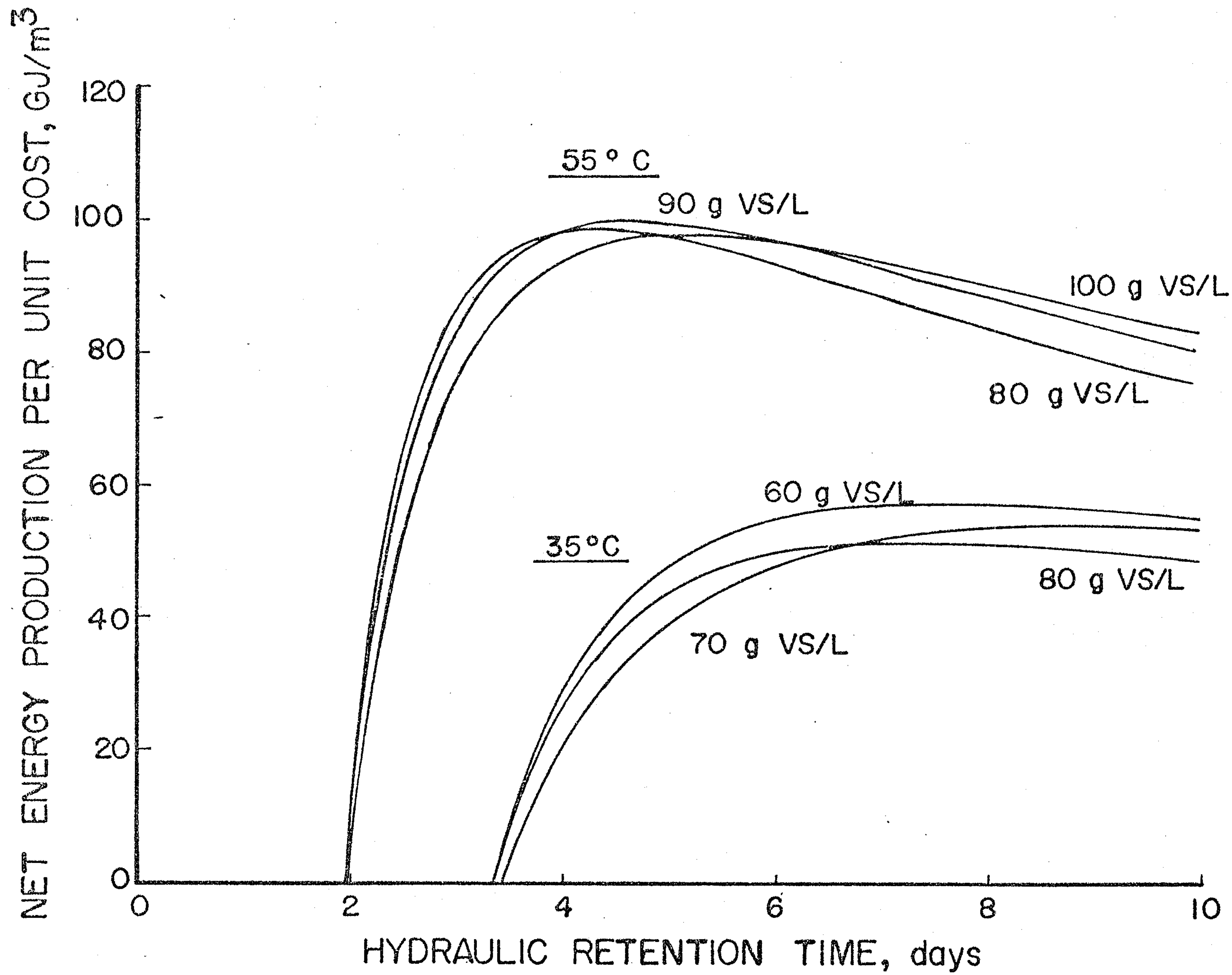

Figure 5.2. Effect of Hydraulic Retention Time, Temperature and Influent Concentration on the Net Energy Production Per Unit Cost.

For all of these options, H_2S and moisture are removed from the biogas by the iron sponge and glycol absorption processes, and the gas is compressed and stored at 860 kPa in tanks with a 1-day gas-production capacity. These processes are necessary to prevent corrosion and to allow some flexibility to modulate the variation in daily gas production. The CO_2 is removed by the water-stripping process.

Table 5.1 summarizes the fermentor volume, energy requirements, and net energy production for the various options and plant sizes. Procedures used to calculate the volume and number of fermentors and energy requirements are presented in Section 4.0.

Since the efficiency of an internal combustion engine to convert CH_4 into useful work increases as the engine size increases, the engine-generator efficiency was estimated by the following equation (assuming a 95% efficient generator and engine efficiencies published by Evans et al., 1973):

$$\text{Engine-generator efficiency (\%)} = 4.18 \ln (E) + 11.61 \qquad (5.2)$$

where E is the gross energy input to the engine (GJ/day) and the maximum efficiency is 38%. Equation 5.2 is only applicable for engines fueled with 100% CH_4; therefore, Equation 5.2 is applicable only for Option D. The engine-generator efficiencies for Option C were assumed to be 60% of that calculated from Equation 5.2, since the useful work output per unit energy input of a high compression engine, receiving gas containing 55% CH_4, is 60% of an engine receiving 100% CH_4 (Neyeloff and Gunkel, 1975).

The net CH_4 production, shown in Table 5.1, was calculated by subtracting the net heating requirement from the gross energy production. The net electrical production was calculated by subtracting the electrical energy requirements from the electrical energy produced by the engine-generator. Option D produced more net electrical energy than Option C because the higher engine-generator efficiency of Option D was sufficient to offset the increased energy for CO_2 removal.

We assumed that the waste heat from the engine would be used to heat the fermentor. This assumption is justified since the net heating requirement is only about 20% of the total energy production, and between 20 to 30% of the gross energy consumed by the engine can be recovered from the jacket cooling system. Since an additional 26 to 30% of the heat can be recovered from the engine exhaust, Options C and D can be a source of low temperature (75 to 85°C) process water, as well as electricity. We did not assume any use of the excess waste heat from the engine generator in our energy balance, nor did we assess any credit for the excess heat generated.

5.3.2 Capital Cost

Table 5.2 shows the installed equipment costs for major components of a 1860 m^3 anaerobic fermentation system, using the various biogas handling and use options identified above. The costs shown in Table 5.2 were used as the basis to estimate the total capital cost at various plant sizes. Total capital costs were estimated by using engineering and inspection fees, contingency, escalation and start-up costs of 14, 10, 18 and 10% of installed equipment costs, respectively. To estimate the total capital costs for different plant sizes, the scale-up factor of digester volume to the 0.7 power, as shown in

TABLE 5.1. ENERGY PRODUCTION AND REQUIREMENTS[a] FOR VARIOUS PLANT SIZES AND ENERGY USE OPTIONS

PARAMETER	PLANT SIZE, Mg TS/day 1	10	100
Working volume, m^3	47	472	4,720
Gross energy production, MJ/day	7,300	73,000	730,000
Energy requirements, MJ/day			
Net heating	1,550	14,100	136,000
Mixing	143	1,430	14,300
Pumping	21	129	635
Gas compression	152	1,520	15,200
Scrubbing pump	111	1,110	11,100
Electrical generator efficiency, %			
Option C	12	18	24
Option D	20	30	38
Net CH_4 production[b], MJ/day			
Options A & B	5,750	58,900	594,000
Net electrical production[b], MJ/day			
Option C	574	10,100	145,000
Option D	1,030	17,400	236,000

[a]Assumes 50% effluent heat recovery, 12 hr/day mixing of influent and fermentor, 10 hr/day pumping of influent and effluent, and ambient and influent temperature of 10°C.

[b]Gross production minus requirement (thermal or electrical energy).

TABLE 5.2. INSTALLED EQUIPMENT COSTS FOR MAJOR COMPONENTS OF A 1860 m^3 ANAEROBIC FERMENTOR

COMPONENT	COST (in $1000) A	B	C	D
Premix & degrit	50	50	50	50
Pump	30	30	30	30
Fermentor w/mixer	250	250	250	250
Heat exchanger	20	20	20	20
Piping	10	10	10	10
Microprocessor-controller	20	20	20	20
Gas cleaner	50	50	50	50
CO_2 scrubber	---	200	---	200
Compressors & storage tanks	100	50	100	50
Boiler	30	30	---	---
Engine-generator w/heat exchangers	---	---	100	100
TOTAL INSTALLED EQUIPMENT COST	560	710	630	780

TABLE 5.3. COSTS FOR PRODUCING METHANE AND ELECTRICITY AT VARIOUS PLANT SIZES

	PLANT SIZE, Mg TS/day		
PARAMETER	1	10	100
Capital costs, $1000			
Option A	65	327	1,640
Option B	83	417	2,090
Option C	73	365	1,830
Option D	91	454	2,270
Labor costs, $1000/yr	11	22	43
Fixed costs, $1000/yr			
Option A	16	79	394
Option B	20	100	501
Option C	17	87	439
Option D	22	109	546
Utility costs, $1000/yr			
Option A	1.7	17	164
Option B	2.2	22	220
Option C	0.1	1	11
Option D	0.1	1	11
Total annual costs, $1000/yr			
Option A	29	117	600
Option A'[a]	10	48	306
Option B	34	144	764
Option B'[a]	12	62	398
Option C	29	111	492
Option C'[a]	9	36	168
Option D	33	132	600
Option D'[a]	10	43	204

[a]Low cost option assumes labor and fixed costs to be 25% and 33%, respectively, of high cost option.

Figure 5.1, was used. Table 5.3 shows the total capital costs for the various options and plant sizes.

5.3.3 Annual Costs

Annual costs (labor, fixed and utility costs) were estimated for the various plant sizes (Table 5.3). Salaries for the plant operators were assumed to range from $11,000/yr for the 1 Mg TS/day plant up to $43,000/yr for the 100 Mg TS/day plant.

Fixed costs were calculated assuming an interest rate of 14% and a 20-year straight-line depreciation of the total capital cost. Taxes, insurance, and repair and maintenance were estimated to be 3, 1.5 and 3% of the installed equipment cost, respectively.

Utility costs were calculated based upon the energy requirements in excess of that produced. Utility rates were assumed to be $14/GJ (5¢/kWh) for electricity and $0.11/m^3 for make-up water. The utility costs shown for Options A and B reflect electricity and water charges. The only utility cost charged to Options C and D was for make-up water, since the engine-generator produced more electricity and waste heat than needed by the fermentation system.

Table 5.3 also shows the total annual costs for the low cost, farmer-contracted and operated systems. The capital costs for these systems were assumed to be one-third the cost of a comparably sized, "turn-key" plant (Figure 5.1). The labor costs were assumed to be one-quarter of the high cost system. This low level of labor can be justified because of the use of microprocessor-controllers to operate and monitor much of the routine operation.

5.3.4 Energy Production Costs

Energy production costs were calculated by dividing the total annual cost for each option by the annual net energy (CH_4 or electricity) produced by each option. The effect of plant size on CH_4 production costs for the high cost systems (Options A and B) and the low-cost systems (Options A' and B') are shown in Figure 5.3. The plant size at which CH_4 production costs equal the current natural gas prices of $3/GJ would be 75, greater than 300, 4.2 and 6.7 Mg TS/day for Options A, B, A' and B', respectively. Thus, although there is only about a three-fold difference in total annual cost between Option A and A', there is about an 18-fold difference in break-even plant size.

Figure 5.4 shows the effect of plant size on electricity production costs. The plant sizes at which the electricity production costs equal the current electricity rate of $14/GJ (5¢/kWh) would be 43, 22, 5.3 and 2.8 Mg TS/day for Options C, D, C' and D', respectively. These results show that there is an eight-fold difference in break-even plant size between the low-cost and high-cost systems, and that the increased electricity generation efficiency, caused by scrubbing CO_2 from the biogas (Options D and D'), more than compensates for the increased total annual costs.

Table 5.4 lists the energy production costs for the various options at different plant sizes (1, 10 and 100 Mg TS/day) and an effluent feed credit of $60/Mg effluent TS. The production costs with effluent feed credit were calculated by subtracting the feed credit from the total annual cost and by

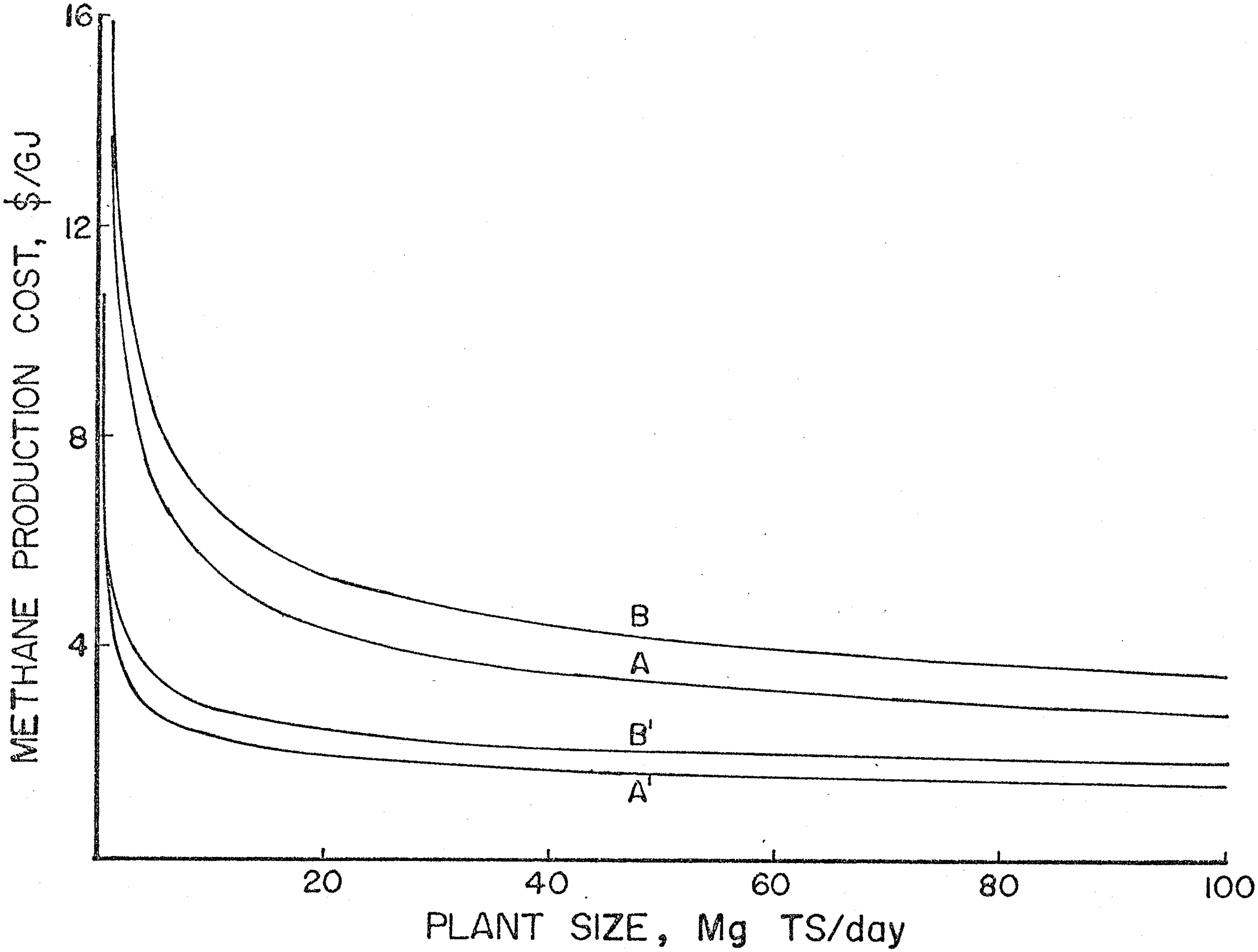

Figure 5.3. Effect of Plant Size on Methane Production Costs (Temperature = 55°C, HRT = 5 days, S_o = 90 g VS/L. A = High Cost, no CO_2 Removal; B = High Cost, CO_2 removal; A' = Low Cost, no CO_2 Removal; B' = Low Cost, CO_2 Removal).

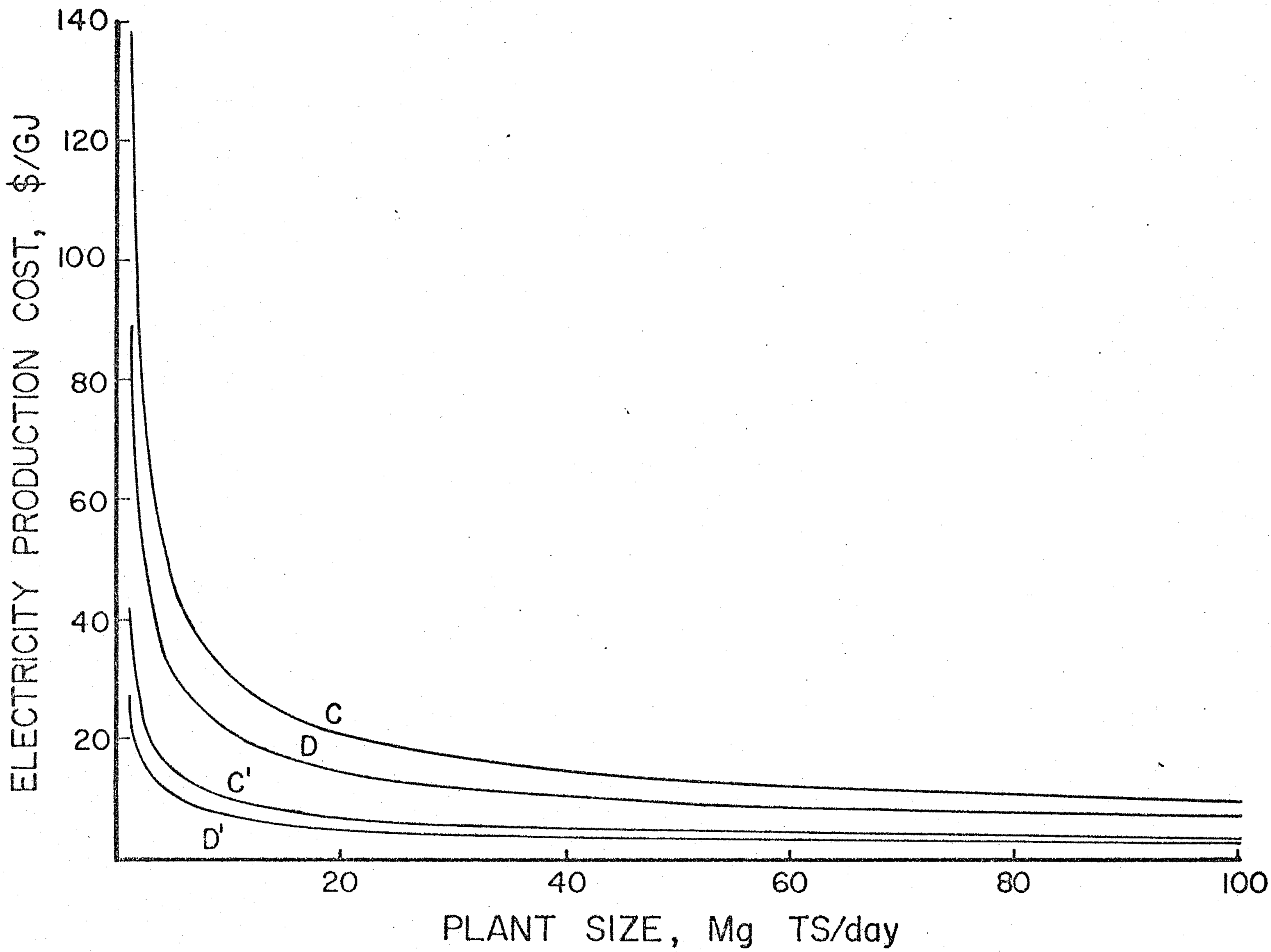

Figure 5.4. Effect of Plant Size on Electricity Production Cost (Temperature = 55°C, HRT = 5 days, S_0 = 90 g VS/L. C = High Cost, no CO_2 Removal; D = High Cost, CO_2 Removal; C' = Low Cost, no CO_2 Removal; D' = Low Cost, CO_2 Removal).

TABLE 5.4. ENERGY PRODUCTION COSTS[a] FOR VARIOUS PLANT SIZES, ENERGY PRODUCTION OPTIONS AND EFFLUENT FEED CREDITS

	PLANT SIZE, Mg TS/day					
	1		10		100	
	Feed Credit		Feed Credit		Feed Credit	
OPTION	$0/Mg	$60/Mg	$0/Mg	$60/Mg	$0/Mg	$60/Mg
A	13.71	8.01	5.46	-0.11	2.77	-2.76
A'	4.67	-1.04	2.26	-3.32	1.41	-4.11
B	15.88	10.23	6.67	1.14	3.49	-1.99
B'	5.57	-0.09	2.83	-2.70	1.82	-3.66
C	138.36	81.18	29.98	-2.46	9.31	-13.35
C'	41.87	-15.31	9.68	-22.76	3.17	-19.49
D	88.68	56.74	20.83	1.94	6.96	-6.94
D'	27.19	-4.75	6.76	-12.13	2.36	-11.54

[a]Costs expressed as $/GJ.

dividing the difference by the annual net energy production. The results in Table 5.4 indicate that the break-even plant sizes are decreased to between 3 and 8 Mg TS/day for the high cost systems when the effluent feed credit is used. For the low cost systems, the break-even plant size is less than 1 Mg TS/day when the effluent feed credit is used.

5.3.5 Implications of this Assessment

This economic assessment has shown that CH_4 can be economically generated at moderate plant sizes (3 to 6 Mg TS/day) when farmer-constructed and operated systems are used or when "turn-key" systems use an effluent feed credit of $60/Mg effluent TS. Assuming that 2.8 kg TS/day can be recovered from a confined steer, the break-even plant size discussed above would serve feedlots between 1,000 to 2,000 cattle. If the farmer-constructed and operated systems also fed the fermentor effluent, then anaerobic fermentation systems would be economically feasible for feedlots of less than 300 cattle.

In this assessment, we assumed that all of the energy produced would be used. Lipper et al. (1976) reported that the energy requirements for energy-intensive commercial feedlots in Kansas were 2.2 GJ/head·yr for natural gas and 0.32 GJ/head·yr for electricity. The net energy production from a plant receiving manure from a 1,000-head feedlot would be about 6.2 GJ/head·yr of CH_4 (Options A and B) or 0.9 (Option C) and 1.6 GJ/head·yr (Option D) of electricity. Thus, more energy would be produced by these systems than could be used by the livestock enterprise. Strategies must be developed to utilize this excess energy. Conscientious effort must be exercised to adjust energy use to production at the enterprise level, and there must be the opportunity to sell the surplus energy.

5.4 SUMMARY

We have presented our approach to optimizing anaerobic fermenter designs by selecting design criteria that maximize the net energy production per unit cost. Using this optimization technique, we estimate that a farmer-constructed and operated system would be economically feasible for beef feedlots between 1,000 to 2,000 head without an effluent feed credit, and about 300 head with an effluent feed credit of $60/Mg effluent TS. Commercial "turn-key" systems are only feasible for feedlots larger than 8,000 head when an effluent credit is not used, and for feedlots between 1,000 to 2,000 head when an effluent feed credit of $60/Mg effluent TS is used. Based on these results, we believe that the economics of anaerobic fermentation is sufficiently favorable for farm-scale application of this technology.

SECTION 6.0

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SECTION 7.0

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16. Abstract (Limit: 200 words) This report summarizes the research to convert livestock manure and crop residues into methane and a high protein feed ingredient by thermophilic anaerobic fermentation. The major biological and operational factors involved in methanogenesis were discussed, and a kinetic model that describes the fermentation process was presented. Substrate biodegradability, fermentation temperature, and influent substrate concentration were shown to have significant effects on CH_4 production rate. The kinetic model predicted methane production rates of existing pilot and full-scale fermentation systems to within 15%. The highest methane production rate achieved by the fermenter was 4.7 L CH_4/L fermenter day. This is the highest rate reported in the literature and about 4 times higher than other pilot or full-scale systems fermenting livestock manures. Assessment of the energy requirements for anaerobic fermentation systems showed that the major energy requirement for a thermophilic system was for maintaining the fermenter temperature. The next major energy consumption was due to the mixing of the influent slurry and fermenter liquor. An approach to optimizing anaerobic fermenter designs by selecting design criteria that maximize the net energy production per unit cost was presented. Based on the results, we believe that the economics of anaerobic fermentation is sufficiently favorable for farm-scale demonstration of this technology.

17. Document Analysis

a. Descriptors Anaerobic Conditions:Q1; Animal Foods; Biomass; Economics; Fermentation: T1,Q2; Manures:T2; Methane:T3; Recovery:Q3; Thermophilic Conditions: Agricultural Wastes; Alkanes; Cryogenic Fluids; Energy Sources; Fluids; Food; Hydrocarbons; Organic Compounds; Renewable Energy Sources; Solid Wastes; Wastes

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