Anaerobic sequencing batch reactor treatment of swine waste at 20° C

by

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I. INTRODUCTION

There has been a general trend in the swine industry away from small farms and toward the large confinement facilities. A large confinement facility enables a more efficient production of animals to satisfy national and global demands for pork. This movement away from small farms, along with an increasing awareness and concern for the environment, has put the spotlight on livestock confinement facilities in the United States.

A. Environmental Concerns

The environmental concerns of livestock waste management have been increasing since the 1960s. It has been claimed that livestock on U.S. farms release wastes to the environment equivalent to a human population of one billion people [14]. In Iowa, for example, the population equivalent of livestock wastes is about 100 million, some 36 times the state's human population [14].

Wastes from confinement facilities pose serious threats to the environment. Among the most severe threats are pollution of ground and surface waters. Perhaps the biggest "Achilles heel" to the swine industry is not the contamination of drinking water and streams, but the odor problem. A more severe, but lesser known problem is the toxicity of the gases emitted from the manure pits in swine facilities. If waste is allowed to sit in pits, anaerobic digestion begins, emitting

methane, carbon dioxide, ammonia, and hydrogen sulfide. The ammonia and hydrogen sulfide levels have to be carefully watched because toxic levels of these gases have been found in confinement facilities [23]. Without proper ventilation and disposal of the waste, these gases can cause injury or death to humans and animals.

B. Scope and Objectives

This study is a continuation of work done by Pidaparti [20]. He studied the anaerobic treatment of swine waste at 35° C and 25° C with an anaerobic sequencing batch reactor (ASBR). The ASBR was developed by Dague and co-workers at Iowa State University and is a patented process (Registration No. 07/701,045). This study was initiated by lowering the temperature of Pidaparti's reactors to 20° C. The reactors were to be run over a range of increasing organic loads until reactor failure. Various hydraulic retention times (HRT) were also evaluated.

There are four reasons for this study. First, the nature of the ASBR process allows solids to be retained in the reactor, thus, allowing stabilization of swine waste at lower temperatures. Secondly, the reactor permits ideal conditions for the generation of biogas (carbon dioxide and methane). The methane produced can be used to generate electricity or burned for its heating value. Third, the ASBR provides an

enclosed vessel for the waste that does not allow odor emissions. The stabilized sludge and supernatant that comes from the digester are odor free since the intermediate acids have been reduced to methane. Fourth, if the reactor was able to stabilize waste at 20° C the process may be able to be used without external heating of the reactor. The energy savings in this case could be substantial.

II. LITERATURE REVIEW

A. Introduction

The purpose of this section is to review anaerobic digestion with emphasis on microbiology, theory, and environmental factors. First, the microbiology and biochemistry of anaerobic processes will be reviewed. This will be followed by a review of parameters important to anaerobic digestion. An introduction to the anaerobic sequencing batch reactor (ASBR) will be next, followed by a qualitative and quantitative analysis of swine waste and its applicability for anaerobic treatment. The concluding section will be a review of studies done on the stabilization of swine waste.

B. Microbiology and Biochemistry

1. Introduction

The microbial environment in an anaerobic reactor is a complex world where different types of anaerobic bacteria live and thrive. Each type of bacteria depends on other types of bacteria to maintain a habitable environment. The types of microbial reactions which take place between species can be described in five different categories. These categories are: mutualism, commensalism, amensalism, competition, and prey predator (Table 1). All five interactions are part of a complex interplay that regulates carbon and electron flows and

Table 1. Microbial interaction prototypes for two-member mixtures [1]

Mutualism	defines interactions, where both members of the mixture derive some advantage from each other's presence in terms of increased growth rates or increased population sizes.
Commensalism	is the situation, where only one member of a community benefits from the presence of the second species, which itself does not derive any advantage or disadvantage from the Presence of the first species.
Amensalism	is an interaction, where growth of one population is restricted by the presence of a second, which itself is unaffected by the metabolism of the inhibited population.
Competition	defines the situation, where the growth rate and final populations are limited by a common dependence on an external growth factor.
Prey-predator	these relationships describe situations where one member of the mixed culture, the predator, gains directly on the expense of living biomass of the second member in the culture, the prey.

the population dynamics in different anaerobic ecosystems. Each of the interactions are for two-member matrixes but the different combinations make up the anaerobic ecosystem.

Anaerobic microorganisms are important because they can colonize anoxic habitats. They do this by various fermentation pathways, which allow substrate-level phosphorylation during the reduction of organic matter into more oxidized and reduced carbon components [1]. The reduced fermentation products are further oxidized with inorganic

electron acceptors other than oxygen. Anaerobic reactions cover a range of redox potentials from -300 to 400 mV. Redox potentials for anaerobic reactions are listed in Table 2. Higher redox potential reactions will occur first in the anaerobic environment because of the higher energy gain from these reactions.

2. Anaerobic reactor environment

Anaerobic digestion of complex matter without the presence of sulfate or nitrate occurs in three trophic groups. These groups are [1]:

1. Hydrolytic fermentation
2. Syntrophic acetogenic
3. Methanogenic fermentation and respiration

Care must be taken when breaking the digestion of complex matter into three categories. These reactions are not separate inside the reactor. Each step needs the others in order for the reactor to perform efficiently.

3. Trophic group I: hydrolytic fermentation

This group breaks down complex organic matter into simple, low molecular weight fermentation end products such as lactate, ethanol, acetate, formate, H_2, propionate, and butyrate (Figure 1). The process starts by the hydrolization of polymers and lipids outside the cells by cell enzymes. The enzymes release the basic structural units such as monosaccharides, amino acids, etc (Figure 2). These units are

Table 2. Important microbial reactions in methanogenic ecosystems [8 & 10]

Reaction	G, kJ/reaction
A. Fermentative reactions:	
(1) $C_6H_{12}O_6 + 3H_2O \rightarrow 3CH_4 + 3HCO_3^- + 3H^+$	-403.6
(2) $C_6H_{12}O_6 + 2H_2O \rightarrow$ 2 ethanol $+ 2HCO_3^- + 2H^+$	-225.4
(3) $C_6H_{12}O_6 \rightarrow$ 2 lactate $+ 2H^+$	-198.1
(4) $C_6H_{12}O_6 + 2H_2O \rightarrow$ butyrate $+ 2HCO_3^- + 2H^+ + 2H_2$	-254.4
(5) $C_6H_{12}O_6 \rightarrow$ 3 acetate $+ 3H^+$	-310.6
(6) 3lactate $\rightarrow$ 2propionate + acetate $+ HCO_3^- + H^+$	-164.8
B. Syntrophic acetogenic reactions:	
(7) lactate $+ 2H_2O \rightarrow$ acetate $+ 2H_2 + HCO_3^- + H^+$	-4.0
(8) ethanol $+ 2HCO_3^- \rightarrow$ acetate + 2 formate $+ H_2O + H^+$	+7.0
(9) ethanol $+ H_2O \rightarrow$ acetate $+ 2H_2 + H^+$	+9.6
(10) butyrate $+ 2H_2O \rightarrow$ 2 acetate $+ 2H_2 + H^+$	+48.1
(11) benzoate $+ 7H_2O \rightarrow$ 2 acetate $+ 3H_2 + HCO_3^- + 2H^+$	+53.0
(12) propionic $+ 3H_2O \rightarrow$ acetate $+ 3H_2 + HCO_3^- + H^+$	+76.1
C. Methanogenic reactions:	
(13) acetate $+ H_2O \rightarrow$ methane $+ HCO_3^-$	-31.0
(14) $4H_2 + HCO_3^- + H^+ \rightarrow$ methane $+ 3H_2O$	-135.6

then small enough to pass through the cell wall and be used as an energy source. The cells then produce the end products mentioned above. The quantity and availability of enzymes are important in this step. Therefore, proper temperature control, adequate mixing, and a large biomass are important [3]. A deficiency in these parameters may lead to inefficient operation. The bacteria compete for nutrients and

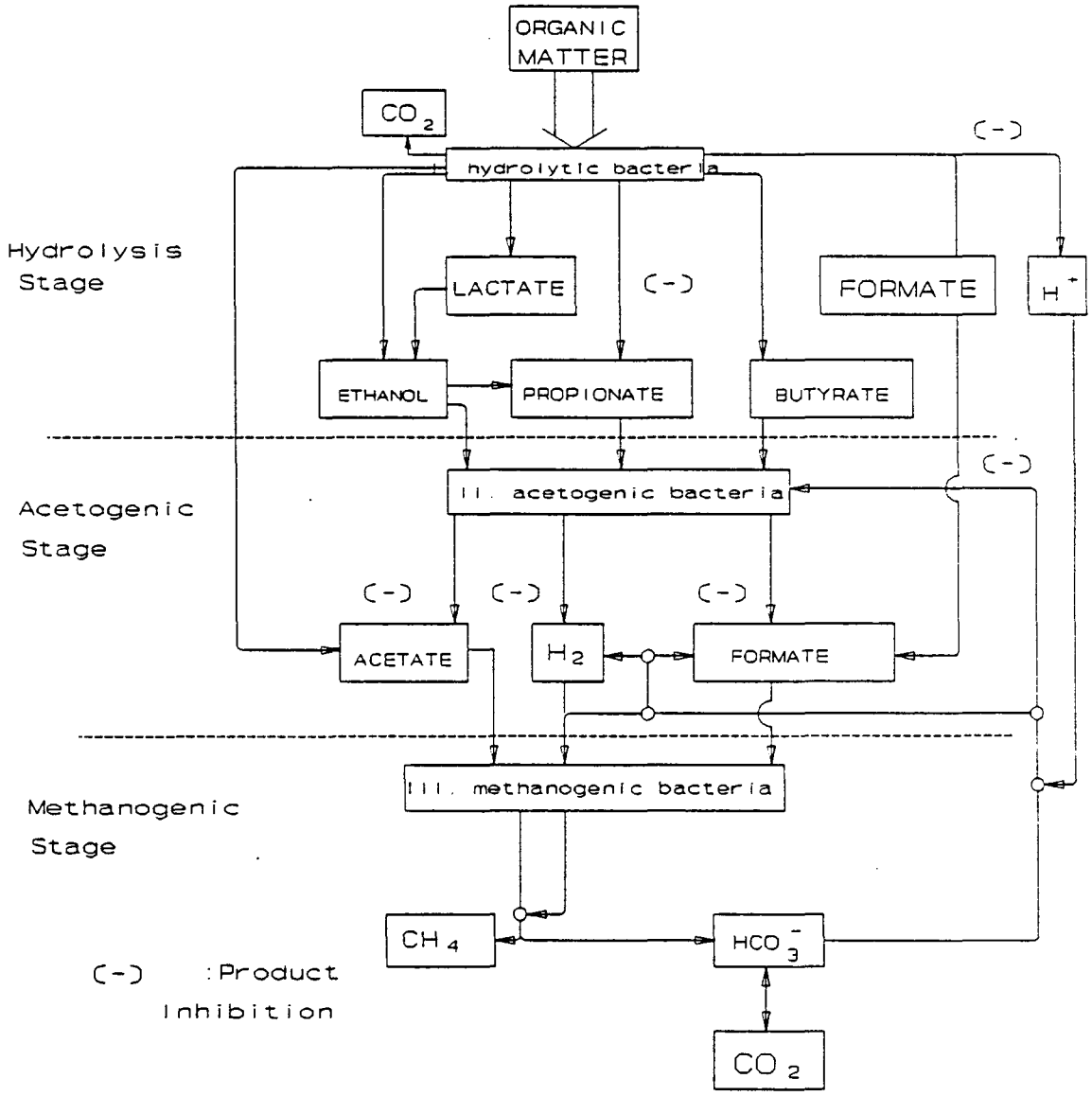

Figure 1. Carbon and electron flow diagram in anaerobic digestion [1]

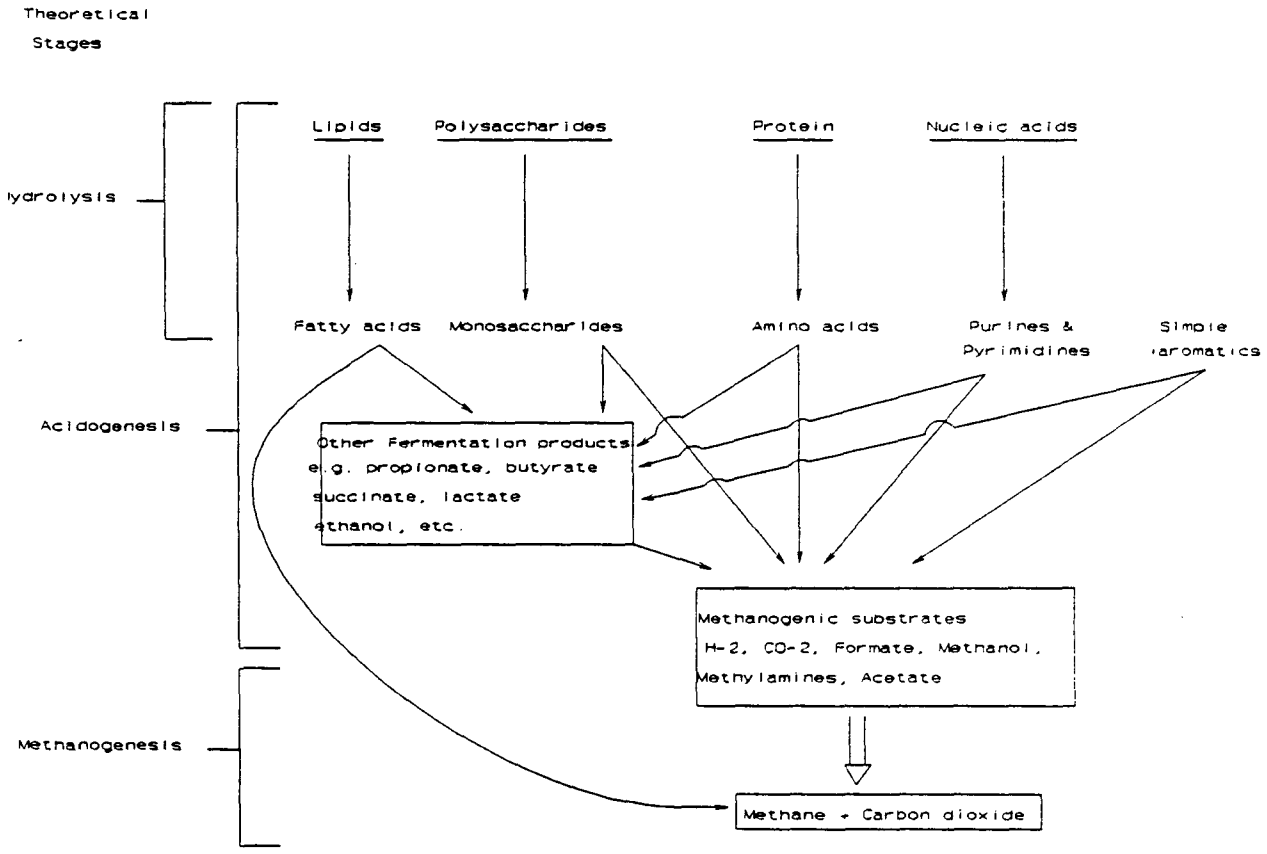

Figure 2. A diagram of the patterns of carbon flow in anaerobic digestion [2]

carbon sources, and are each affected by the several species of bacteria present in this step. The complex nature of the matter creates niches for different types or colonies of fermentative organisms. These organisms have mutualistic metabolic interactions which balance population size and growth. One example is where one type of bacteria removes and uses matter that is toxic to a second type of bacteria. The second type of bacteria, in turn, produces an end product usable by the first type. The growth of the hydrolytic fermentation group is inhibited by high concentrations of the end products. This group is especially sensitive to organic acids, anions, protons, and hydrogen. It depends on the other groups to remove these end products.

Energy metabolism involves substrate level phospholylations during glycolysis and the disposal of the generated reducing equivalents with the organic fermentation end products produced, including ethanol and lactate. (Table 2, reactions 2 and 3). Excess reducing equivalents discharge from fermenting cells as hydrogen or formate.

If a waste is very complex and hard to hydrolyze the hydrolytic fermentation step can be rate limiting. Also, not all organic matter can be hydrolyzed. Non-hydrolyzable matter is called non-biodegradable. Often, however, the speed of this reaction and the low pH at which it occurs can inhibit acid removal by the other groups causing the reactor to fail or to sour. This is an example of an amensalistic

interaction. No methane is produced in this step, therefore no stabilization takes place [4].

4. Trophic group II: syntrophic acetogenic bacteria (SAB)

The SAB group contains eubacterial species that oxidize hydrolytic fermentation end products like ethanol, propionate, butyrate, and benzonate to acetate (Figure 1). The reducing equivalents generated by oxidation are used to reduce protons to H_2, or CO_2 to formate (Table 2). The H_2-producing acetogenic bacteria oxidize alcohols to acetate and H_2. They also carry on β-oxidation of fatty acids. Even numbered carbon fatty acids are converted to acetate while odd numbered carbon fatty acids are converted to acetate, propionate, and H_2 [8]. These reactions are characterized by a positive free energy change. This group grows in the presence of H_2/formate (XH_2) consuming bacteria (syntrophy). They provide H_2/formate as an energy source to the methanogenic bacteria. This process is termed "interspecies electron transfer" or "XH_2 transfer". The high free energy of XH_2-consuming bacteria result in low levels of XH_2. Large acetogenic XH_2 production may lead to product accumulation and inhibition of SAB. The accumulation is controlled by the methanogenic specific growth rate.

5. Trophic group III

This is the final and most important stage. This group removes the acetate, formate, and H_2, and produces methane. Most of the waste stabilization takes place in this step.

There are thirty different known methanogenic species which can be classified into fourteen genera and five families. Table 3 shows the different pathways in the anaerobic process. The percentages are based on COD (chemical oxidation demand) of sewage. The table shows that the vast majority of methane is formed from acetic acid by the acetoclastic methanogens. Besides acetic acid, there are only a few other substrates which are usable by the methanogens.

6. Acetoclastic methanogens

This is the most important microbial group of the methanogens. The conversion of acetic acid to methane accounts for 72% [4] of the methane produced. This is remarkable since only four types of methanogens can use acetic acid as a substrate. This process also has the smallest free energy change making it the least desirable pathway. Another important result of the acetoclastic methanogens is pH control. The production of bicarbonate alkalinity (Table 2) helps maintain the pH in the reactor, thus providing a healthy environment for the whole consortium of microorganisms. The reaction which shows how acetic acid is converted into methane is called acetic acid cleavage (Table 4). The asterisked carbon is the methyl group of the acetic acid. This group is the source of the methane. The other carbon group is the carboxyl group which gives carbon dioxide. Carbon dioxide is important because it further combines with excess hydrogen in a chemical reaction resulting in methane.

Table 3. Characteristics of methanogenic species in pure culture [11]

Genera	Species	Morphology
Methanobacterium	*formicicum*	Long rod to filament
	bryantii	Long rod
	thermoauto-trophicum	Long rod to filament
Methanobrevibacter	*ruminantium*	Lancet-shaped cocci
	smithii	Lancet-shaped cocci
	arboriphilus	Short rods
Methanomicrobium	*mobile*	Short rods
Methanogenium	*cariaci*	Irregular, small cocci
	marisnigri	Irregular, small cocci
Methanospirillum	*hungatei*	Short to long Wavy spirillum
Methanosarcina	*barkeri*	Pseudosarcina
Methanococcus	*vannielii*	Irregular, small cocci
	voltae	Irregular, small cocci

Table 4. Major mechanism of methane formation [4]

I. Acetic Acid Cleavage:

$$C^{*}H_3COOH \rightarrow C^{*}H_4 + CO_2$$

II. Carbon Dioxide Reduction:

$$CO_2 + 8H \rightarrow CH_4 + 2H_2O$$

7. Other methanogens

The remaining methane is produced by other methanogens using substrates such as formic acid, methanol, or hydrogen. These reactions are important because they provide 28% of the methane based on COD. Table 5 gives most of the known methanogenic reactions along with free energies for some of the reactions. All of the reactions yield methane as the end product. However, the ratio of methane/carbon dioxide changes with the oxidation state of the substrate. Buswell and Mueller [9] developed a prediction equation for the amount of methane produced from a given substrate. Tests have shown that the actual yields are 95 to 100% for this theoretical equation:

$$C_nH_aO_b + \left(n-\frac{a}{4}-\frac{b}{2}\right)H_2O \rightarrow \left(\frac{n}{2}-\frac{a}{8}+\frac{b}{4}\right)CO_2 + \left(\frac{n}{2}+\frac{a}{8}+\frac{b}{4}\right)CH_4$$

8. Separation of stages

Several studies have been done on the separation of stage one and two from stage three in anaerobic treatment. The advantage is that optimization of the hydrolysis and SAB step can be achieved as well as optimization of the methanogenic step. At first glance, this seems like a good idea because the two stages have optimal environments at different pHs. The methods of separation of the stages include kinetic control, dialysis membranes, and poisoning or inhibiting methane formers [3]. The separation of the stages has its

Table 5. Energy yielding reactions used by methanogens [8 & 20]

Reaction	G° (kJ/reaction)	G° (kJ/CH_4)
$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O$	-138.8	-138.8
$4HCOOH \rightarrow 3CO_2 + CH_4 + 2H_2O$	-119.5	-119.5
$4CH_3OH \rightarrow 3CH_4 + CO_2 + 2H_2O$	-310.5	-310.5
$4CH_3NH_3 + 2H_2O \rightarrow 3CH_4 + CO_2 + 4NH_4^+$	-225.7	-75.2
$4CO + 2H_2O \rightarrow CH_4 + 2CO_2$	-185.6	-185.6
$CH_3COOH \rightarrow CH_4 + CO_2$	-27.6	-27.6

drawbacks since the methanogens are important to the SAB's. The methanogens remove the hydrogen and formate which is toxic to the SAB's. Also, the SABs. produce nutrients for the methanogens which may be lost during separation of stages. Stage separation is only considered feasible when the first stage is rate limiting or the waste has a high solids content. A high solids waste such as trash would be a good candidate for stage separation.

9. Conclusion

The anaerobic process is much more complicated than the three stage theory suggests. Figure 1 shows that there are many interactions between the different groups of bacteria and

that each group depends on another group for its survival. The whole environment can be described as mutualistic. The interactions between groups control the growth and population sizes of the reactor.

C. Parameters Important to Anaerobic Treatment

1. Introduction

There are certain factors that play a particularly important roll in the anaerobic treatment process. These factors can be broken into two categories, environmental, and operational factors. The environmental factors include temperature, pH, alkalinity, volatile acids, toxic metals, and ammonia. The operational factors include solids retention time (SRT), and organic loading rate. These are not all of the factors that affect the process, but these are the most important.

2. Environmental factors

a. Temperature Temperature greatly affects the stabilization of waste in a reactor. At higher temperatures reaction rates tend to increase thus increasing the rate at which material is stabilized. The temperature not only affects the rate at which material is oxidized but also the rate of synthesis, regeneration, and endogenous respiration [12]. There have been studies which have shown that reactions can take place as low as 10° C [31]. However, Van Velson et al

[13] showed that at 13° C using swine waste no treatment took place. This was determined by the fact that no gas was produced in the reactor at 13° C. Pfeffer et al. found that a reactor at 25° C can get the same removal as that at a temperature of 35° C but that it takes a longer SRT to build the necessary increase in microbial population. This lag is explained by the slower reaction rates which occur at 25° C as opposed to 35° C. Heukelekian et al [27] found the optimum mesophilic temperature to be 28° C.

b. pH, alkalinity, and volatile acids The effect of non-optimum pH on methane fermentation is the result of many reactions and equilibrium shifts in a reactor. The pH tolerance of an organism is usually considered a direct reflection of the pH-activity characteristics of that organism's enzymes [15]. Some general ways in which pH affects enzyme activity are (a) change of state of the enzyme ionizable groups such as carboxyl and amino groups, (b) alteration of the non-enzyme components of the system such as substrate ionization, and (c) denaturing of the enzyme protein structure [15]. These and other effects can occur with pH changes in anaerobic treatment.

The pH of the liquor undergoing anaerobic treatment is related to several acid-base equilibria. However, for the area of normal anaerobic treatment (between 6.5 and 8) the major chemical system controlling pH is the carbon dioxide-

bicarbonate system. This system is related to the hydrogen ion concentration via the following equation:

$$[H^+] = K_1 \frac{[H_2CO_3]}{[HCO_3^-]}$$

The carbonic acid concentration (H_2CO_3) is related to the percent carbon dioxide of the reactor gas. K_1 is the ionization constant that governs the reaction. This constant changes as the temperature of the reactor changes. The bicarbonate ion concentration (HCO_3^-) forms part of the total alkalinity [4]. This formula is also important in determining the total alkalinity of the reactor.

Volatile acids are the measure of the intermediate organic acids in a reactor which are formed in the hydrolysis step. These acids can cause a low pH which can adversely affect the methane formers which need a pH near neutral.

The total alkalinity represents the buffering capacity of the reactor against a pH change. A reactor with a low alkalinity is susceptible to rapid detrimental changes in pH, causing imbalances. On the other hand, a reactor with a higher alkalinity can withstand slight imbalances. The alkalinity reacts with the volatile acids produced, neutralizing them, which helps maintain a healthy environment for the methane formers. Total alkalinity is based on the bicarbonate alkalinity and the acetic acid concentration, and

is expressed by the following formula [4]:

$$BA = TA - (0.85)(0.833)TVA$$

where:

BA = bicarbonate alkalinity mg/l as $CaCO_3$;

TA = total alkalinity mg/l as $CaCO_3$;

TVA = total volatile acid concentration, mg/l as acetic acid.

The factor 0.85 comes from the fact that only 85 percent of volatile acid alkalinity is measured by titration of total alkalinity to pH of 4. The 0.833 is a conversion factor (50/60) to change total volatile acids, in mg/l as acetic acid to total volatile acid alkalinity, in mg/l as $CaCO_3$. The equation also assumes no other significant concentrations of materials which will produce alkalinity. These materials could be, but are not limited to, phosphates, silicates, ammonia, hydrogen sulfide, or other acid salts.

There has been much study as to the optimum pH at which the anaerobic process should take place. All three phases of anaerobic treatment require a different optimum pH. When all three phases are put together the optimum pH is dependent on the rate limiting step of the overall treatment. This rate limiting step is the methane formation step. Heukelekian et al. found that the pH of 7.0 was the best pH for anaerobic treatment [27]. He found that as the pH moves either way of 7 the number of organisms fermenting waste sharply declines.

Later work calls this into question. Dague [14] indicates that the optimum pH is not one value but a range of values and the optimum range was 6.8 to 7.2 with the limit of the range for operation without significant inhibition being 6.5 to 7.6 [8]. This coincides with McCarty who said that the pH range without inhibition is 6.6 to 7.6 [4]. Clark and Speece found that acetate fermentation can be carried out without inhibition at a pH range between 6 and 8 [15]. This may be due to selection of pH tolerant methanogens in the reactor.

The total alkalinity inside a reactor tends to increase as the load increases. This is true as long as none of the reactions are inhibited. This is due to the production of ammonia from the degradation of proteins in the reactor. The protein-containing wastes result in production of ammonia which reacts with carbon dioxide to form ammonium bicarbonate (NH_4HCO_3) [8].

c. Ammonia Ammonia has many functions in the reactor. High levels of ammonia can be toxic. Ammonia nitrogen also forms alkalinity in a reactor and can change the total alkalinity, pH, and volatile acids.

Ammonia is usually formed in anaerobic treatment from the degradation of wastes containing proteins or urea. It is present in the form of the ammonium ion (NH_4^+) or as dissolved ammonia gas (NH_3). These two forms are in equilibrium with each other as follows:

$$NH_4^+ \rightleftharpoons NH_3 + H^+$$

When the hydrogen ion concentration is high (pH 7.2 or lower) the equilibrium is shifted left so the equilibrium is shifted to the ammonium ion. Normally all the ammonia is measured as ammonia nitrogen which includes both ammonia and ammonium.

The concentration at which ammonia is toxic has been stated in many ways. McCarty said that if the concentration of ammonia nitrogen is between 1500 and 3000 mg/l, and the pH is between 7.4 and 7.6, the ammonia (NH_3) can become inhibitory. Lowering the pH at these ammonia concentrations causes a shift in the reaction toward the ammonium ion and thus these ammonia-nitrogen concentrations will not be toxic. Regardless of pH, total ammonia nitrogen concentrations in excess of 3000 mg/l is quite toxic [6]. Albertson found that ammonia nitrogen concentrations of 1200 to 1400 caused failure at loads between .2 to .44 lb volatile solids (VS)/day/ft^3 [16]. It should be noted that in Albertson's studies the pH was quite high (7.2 to 7.8), which would lower the concentration at which ammonia would be toxic. Ammonia toxicity can also be expressed in the form of free ammonia. McCarty and McKinney found free ammonia concentration to be toxic starting from 130 mg/l, as N, to 150 mg/l, as N [73].

d. Toxic metals, salts, and sulfides Besides ammonia, there are many materials which are toxic to anaerobic reactors. Almost any essential nutrient is toxic at high concentrations. The concentration at which substances become toxic varies. Some substances are toxic at a fraction of a mg/l. Others are not toxic until many thousands of mg/l. Generally, salts and other materials have a stimulatory effect at low concentrations and then a toxic effect at higher concentrations. Figure 3 illustrates the effect of salts and other materials on biological reactions. Some of the most common forms of toxicity stem from the alkali and alkaline-earth metal salts. These include sodium, calcium, potassium, and magnesium. These do not commonly occur in great enough concentrations in municipal waste waters to cause a problem. However, industrial wastes may contain concentrations high enough to inhibit or stop anaerobic treatment. Table 6 [6] shows the stimulatory and inhibitory concentrations of the alkali and alkaline earth cations.

Sulfide toxicity can occur from the introduction of sulfides or the biological production of sulfides from sulfates and sulfur-containing organic compounds. In industrial wastes sulfate salts usually represent the major precursors of sulfides. Sulfides exist in one of three forms. They may exist as gaseous hydrogen sulfide, soluble, or insoluble form. The form depends on the pH of the solution as well as the amount of cations available to form a harmless

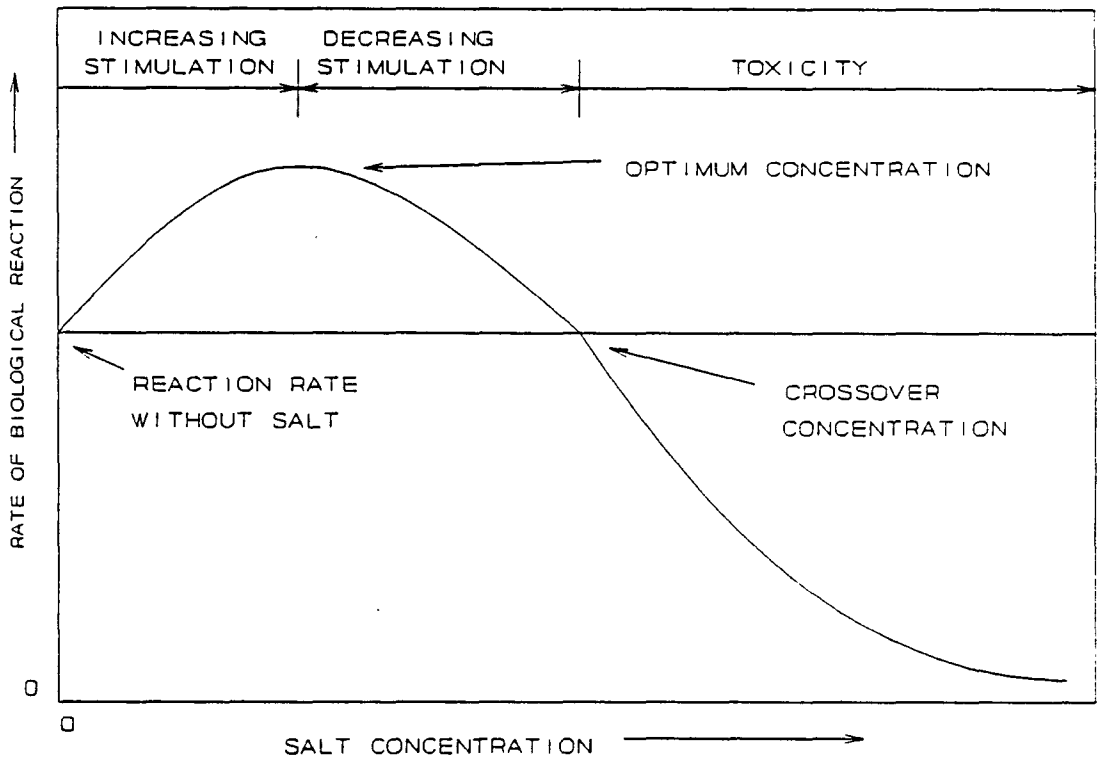

Figure 3. General effect of salts or other materials on biological reactions

Table 6. Stimulatory and inhibitory concentrations of alkali and alkaline-earth cations [6]

Cation	Concentrations in mg/L		
	Stimulatory	Moderately Inhibitory	Strongly Inhibitory
Sodium	100 - 200	3500 - 5500	8000
Potassium	200 - 400	2500 - 4500	12000
Calcium	100 - 200	2500 - 4500	8000
Magnesium	75 - 150	1000 - 1500	3000

precipitate. Sulfides become toxic at concentrations above 100 mg/l [6]. However, with some acclimation, sulfide concentrations up 200 mg/l have been tolerated in reactors.

Heavy metals are toxic at very low concentrations. Some of the most common heavy metals include copper, zinc, cadmium, lead, mercury, silver, and nickel. Heavy metals exist in two forms, insoluble and soluble. They inhibit the anaerobic process by inactivating a wide range of enzymes. They inactivate enzymes by reacting with their sulphydryl group. This inhibits or stops the growth of the organisms. The harmful metals are those in the insoluble form. The concentration of soluble heavy metals depends on the concentration of sulfide in the system. The heavy metals tend to form sulfide precipitates which renders them harmless. The equilibrium reaction between sulfides and heavy metals also depends on pH and temperature. In fact, most researchers determine toxicity of heavy metals based on sulfide concentrations because it is relatively easy to measure.

There have been studies done on the toxicity of heavy metals. Barth et al. did a report on the effect of heavy metals on the aerobic and anaerobic process [17]. Two very comprehensive studies were done on maximum concentrations and inhibitory concentrations of a variety of heavy metals by Mosey [18 and 19]. Mosey also studied the synergistic effects of heavy metals and presented new ways of determining heavy metal concentrations.

3. Operational parameters

a. Solids retention time Solids retention time is the average amount of time that the solids reside in the reaction vessel. Each different microbial group has some minimum retention time that needs to be maintained for that group to thrive. If the retention time is too low the bacteria group does not have time to reproduce and are "washed out". The minimum time needed in a system is affected by several different factors. These factors include temperature, availability of food, and toxic materials. In the anaerobic reactor the group requiring the longest retention time are the methanogens. At 35° C, some methanogens can regenerate in about ten days. Some species of methanogens require only four days. Pfeffer found that the SRT increases 50% for every 10° C of temperature decrease [12].

The solids retention time (SRT) is found by dividing the total biomass in the system by the biomass wasted daily.

$$SRT = \frac{Biomass\ inside\ system}{Biomass\ removed\ per\ day}$$

The use of suspended solids for this calculation is adequate in that the average retention time of the mixed suspended solids in the system is a close approximation to the real SRT. This is true as long as flocculation of the organisms is taking place. It has been shown that under certain conditions the above ratio does not adequately measure the SRT [74].

The minimum SRT can be expressed using Monod kinetics. McCarty [75] developed an equation for the growth of microorganisms as a function of time after organic waste has been added. The following is the relationship:

$$\frac{dM}{dt} = a\frac{(dF)}{dt} - bM$$

where:

dM/dt = growth of microorganisms in mass per unit time;

dF/dt = rate of BOD removal in mass per unit time;

M = Mass of microorganisms present;

a = growth yield constant;

b = microorganism decay rate in units/time.

The equation states that the rate of growth of microorganisms equals the rate of waste utilization for synthesis minus the endogenous decay rate which is a function of the mass of organisms. This means that the more food that is utilized the more organisms are produced and thus the more that die off.

The rate of substrate utilization was expressed by the following formula:

$$\frac{dF}{dt} = \frac{kMs}{K_s + s}$$

where:

k = maximum specific rate of waste utilization;

K_s = waste concentration at which dF/dt is one-half the maximum rate;

s = concentration of the waste surrounding the microorganism.

Combining the previous two equations, McCarty provided an expression for the growth rate of the microorganisms in terms of waste concentration:

$$\frac{\frac{dM}{dt}}{M} = \frac{aks}{K_s+s} - b$$

where:

$(dM/dt)/M$ = specific growth rate or fractional growth rate per unit time.

These equations explain the kinetics of the microbial reactions in anaerobic and aerobic systems. They show that the rate of substrate removal depends on the food available, the mass of microorganisms, the death rate of microorganisms, and the rate of waste utilization. The rate constants from the Monod functions are dependent on temperature and the other environmental factors mentioned above. The reciprocal of the

specific growth rate gives the biological solids retention time t_s. It can be defined as follows:

$$t_s = \frac{M}{\frac{\Delta M}{\Delta t}}$$

This equation is the derivation of the SRT equation given in the beginning of this section. The minimum SRT can also be derived from Monod kinetics. The minimum SRT is reached when $s = S$, and can be approximated by the following, considering b to be negligible:

$$(t_s)_{min} = \frac{1}{ak}\left(\frac{K_s + S}{S}\right)$$

The minimum t_s is a function of the fraction of organic waste converted to biological cells a; the maximum rate of waste utilization, k; and the raw waste concentration S. The value of a is much smaller in anaerobic processes than in aerobic processes giving the minimum retention a greater value [12].

b. Organic loading rate The organic loading rate is the rate at which feed enters the system. Monod kinetics shows that the amount and type of food plays an important role in the mass of microorganisms and the rate of their growth. The fraction of organic waste converted to biological cells is primarily a function of the energy yield from the metabolism of the substrate. Since the methane formers represent the rate-limiting step in anaerobic treatment, the minimum

retention time is determined by the energy yield in methane fermentation.

The amount of methane that can be produced from an organic load remains constant. This is often overlooked in research. Many researchers point out the rate of methane yield, but fail to look at the total methane produced per mass of organic load. The amount of methane and therefore the amount of biogas produced increases as the SRT increases. This is simply because the organisms have a longer contact time with the food. Figure 4 by Dague [14] shows how the total gas and methane rates increase with an increase in SRT.

Although Monod kinetics still prevail, the effects of mixing have to be mentioned. The food to microorganism ratio is not always consistent throughout the anaerobic reactor. A reactor with poor mixing may develop pockets of food where some parts of the reactor contain a lot of food and other parts very little food. This would cause some parts of the reactor to produce gas at higher rates than others. It is critical to get the food to the microbial population or the rate of treatment could be limited by inadequate mixing.

D. Fundamentals of the Anaerobic Sequencing Batch Reactor

1. Introduction

The anaerobic sequencing batch reactor is a patented system being developed at Iowa State University (Registration No. 07/701,045). This process was used by Pidaparti to treat

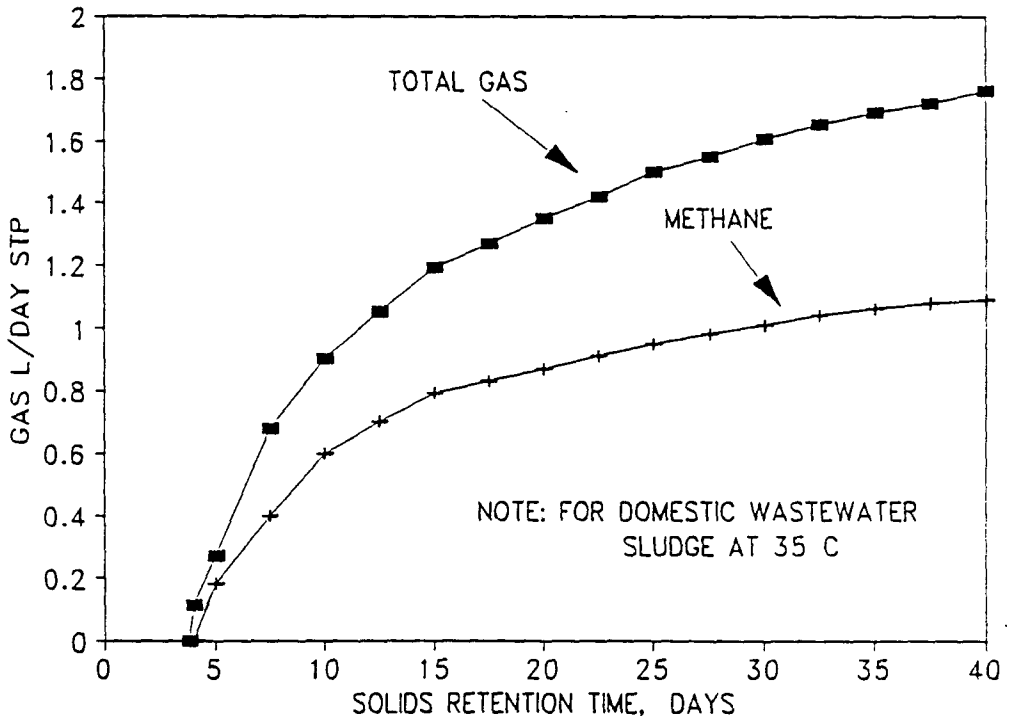

Figure 4. Effect of solid retention time on gas production [14]

swine waste at 35° C and at 25° C [20]. This process is also the process presented here to treat swine waste at 20° C.

2. Principles

Dague et al. [21], among others, realized long ago the importance of SRT in anaerobic methanogenic processes. The initial inspiration for the ASBR process came from work done by Dague et al. on the anaerobic activated sludge process [21]. This work was inspired by earlier work by Schroepfer et al. when they worked on meat packing waste using a process known as the "anaerobic contact process" [22]. Schroepfer's process was similar to the anaerobic activated sludge process, but employed a clarifier after the contact process to remove solids and reinduce them into the reactor. This improved the SRT and thus the performance of the reactor.

The ASBR can achieve solids capture and removal of organics in one vessel eliminating the need for a clarifier. The process operates on batch kinetics and consists of four phases (Figure 5). The first phase of the ASBR is the feed phase. The reactor is batch fed with an organic waste at a predetermined volume. The second phase is the react phase in which the reactor is mixed for a pre-determined period of time. The mixing allows the microorganisms to contact the food and stabilize it. According to Monod kinetics, the high food to microorganism (F/M) ratio to start the react cycle causes the gas production rate to be high. As the food starts

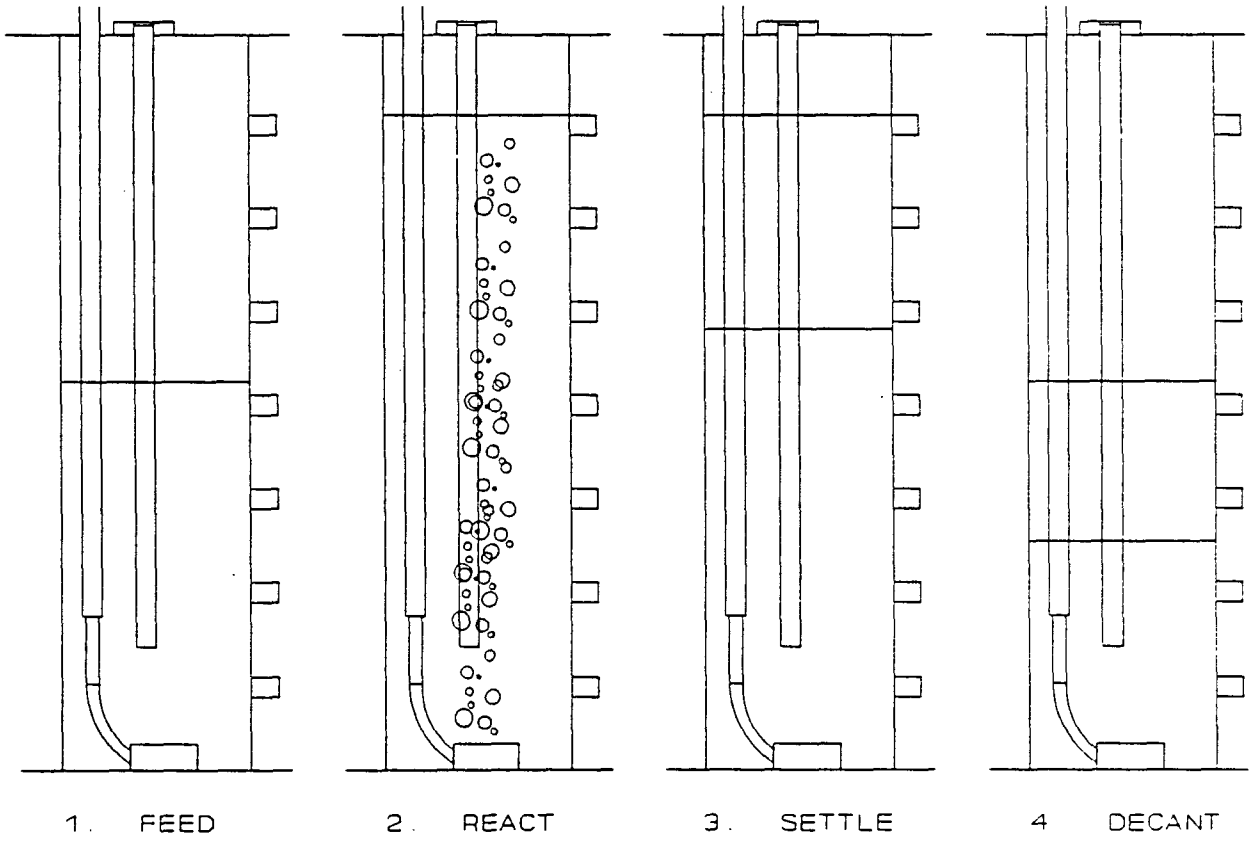

Figure 5. Principles of the anaerobic sequencing batch reactor

diminishing the gas production rate decreases. When the gas production rate is low, and the F/M ratio is low, the reactor is allowed to go to phase three which is settling. . The low F/M ratio which occurs during the settle phase allows the solids to settle much better than a high F/M ratio. It is important to note that the F/M in a continuously fed reactor is constant. After a period of settling, phase four takes place which is the decanting of supernatant. Only the clear supernatant is decanted leaving the high solids sludge blanket intact at the bottom of the reactor. The reactor is then fed again, completing the cyclical process.

The settling phase allows the solids to remain in the reactor. The solids continue to build up improving SRT and removals. The process works so well that many times solids have to be wasted to prevent too many solids in the reactor. Preliminary results in the lab have also found that this process promotes granulation which enhances the settling characteristics of the biomass.

The ASBR has shown a tendency to compensate for temperature. As the temperature is lowered in a reactor the microorganisms remove organic loads at a lower rate. However, the ultimate gas yield of an organic load remains the same at all temperatures. Since the ASBR has the ability to hold solids, as the temperature decreases the SRT tends to increase. This is mostly because the endogenous decay rate of the microorganisms is lower at the lower temperature. It

remains to be seen how low the temperature can be lowered in the ASBR and still get good removals. There will be a point at which the reactor cannot hold the solids long enough for removal to take place.

E. Swine Waste Characteristics

1. Introduction

Swine waste generated from confinement facilities represent a real problem to the facility owners and to the public. The most potent problem is odors. This problem has been compounded by the movement away from swine farms and toward the much larger confinement facilities. In order to solve the waste problem, a qualitative and quantitative analysis has to be run on the industry and its waste to determine design parameters. The following section is dedicated to characterizing swine waste generated by confinement facilities.

2. Quantitative aspects of swine waste

The quantity of waste is dependent on many factors. Not only is it dependent on the size of the hog and the metabolism, but also on environmental factors. The metabolism of the feed varies with the age of the animal. The hog weight gained per day is usually used to express the amount of swine waste generated. Hazen and Mangold [70], considering environmental conditions in the confinement feeding facility as well as feed rations, estimated the variations in the daily

manure production to be expected with the growth of the animals. They calculated the average daily manure production per hog to be one tenth the live weight of the animal.

Manure is not only influenced by the age of the hog and the feed, but also by environmental conditions. Changing environmental factors such as overcrowding, humidity, and temperature of the building can change the rate of manure production. The air and humidity play a factor because the higher these factors the more water is lost through exhaling air and evaporation of the manure. The number of pigs in a pen is a factor because pigs tend to get stressed when overcrowded, growing slower and producing less waste.

Taiganides [76] has presented a theoretical mass balance for the amount of waste produced per 100 lbs of live weight. The balance includes feed, water intake, heat and moisture loss to the environment, and weight gain. An average value of 4.7 lbs. of weight per 100 lb. live weight is suggested by Taiganides. Irgens and Day [77] reported values based on different weights of animals (Table 7).

3. Qualitative aspects of swine waste

Swine wastes contain the same types of organic matter as human feces but in different concentrations. Generally, the manure contains bits of undigested food, bacteria in the digestive track, digestive enzymes, and water. The chemical composition of swine waste is given in Table 8. There are other factors besides these which are more important to the

Table 7. Manure production from growing and finishing swine [77]

Animal Weight lb	Feces lb/d	Urine lb/d	Total Manure lb/d	Total Volume gallons
12 - 40	0.7	1.5	2.2	0.36
40 - 80	2.7	2.9	5.6	0.67
80 - 120	5.4	6.1	11.5	1.38
120 - 160	6.5	8.1	14.6	1.75
160 - 200	8.5	9.1	17.6	2.11

anaerobic treatment of swine waste. These parameters include: pH, alkalinity, volatile acids, ammonia, total solids, total volatile solids, COD, and BOD, among others. As the quantitative aspects of waste change with type of hog and environmental conditions so do the qualitative aspects. Donham et al. [23] did a study on these parameters by studying 23 different pits at confinement facilities of all different types. These types include farrowing, nursery-grower, grower-finisher, and finishing facilities. They found that the characteristics change based on the type of facility. They also found that farrowing buildings had a considerably higher sulfide concentration than the other types of facilities. The volatile solids and total solids were all much higher in buildings with larger sized pigs. Norman [78] summarized many qualitative studies done by researchers on swine waste.

Another qualitative characteristic of swine waste is the

Table 8. Chemical Composition of Swine Waste (dry weight basis, based on TS (total solids) [80]

Component	Concentration
Cellulose, % TS	11.4
Hemicellulose, % TS	16.5
Lignin, % TS	69.0
Protein, % TS	18.9
Lipids, % TS	13.7
Total Nitrogen, % TS	5.3
Ammonia Nitrogen, % TS	2.2
Ash, % TS	14.0
Calcium, % TS	3.2
Phosphorus, % TS	1.7
Potassium, % TS	0.9
Magnesium, % TS	0.8
Sulfur, % TS	0.3
Copper, ppm	249.0
Zinc, ppm	526.0
Iron, ppm	1940.0
Aluminum, ppm	544.0
Manganese, ppm	342.0
Cobalt, ppm	6.1
Molybdenum, ppm	0.3
Cadmium, ppm	1.0
Lead, ppm	12.1
Gross Energy Content, MJ/Kg	18.1

offensive odor. The odor is caused by many compounds and intermediate acids. Some of the more odorous compounds in swine waste include: hydrogen sulfide, ammonia, propionic acid, acetic acid, butyric acid, valeric acid, aromatic amines, and aromatic compounds such as phenols, and cresols.

4. Applicability of anaerobic treatment of swine waste

The applicability of the anaerobic treatment process for swine waste must be looked at before any study commences. The waste itself must be quantified and qualified as described above. These parameters are then used to determine if the right amount of nutrients are available. If nutrients are lacking some may have to be added to aid in treatment. Toxic levels of substances must be checked to determine if any substance exists at levels toxic or inhibitory to anaerobic treatment. If toxic substances exist than the waste may have to be diluted or treatment abandoned.

Dague [14] said that of all the inhibitory compounds to an anaerobic reactor, two would be of primary concern in the treatment of swine waste. These two compounds are ammonia and sulfides. It was also stated that the problem could be solved by dilution of the swine waste to dilute the ammonia and the addition of iron salts to precipitate the sulfides. Often in a swine operation, water is used to flush the pits thus reducing the ammonia concentrations to tolerable levels.

Donham et al. [23] look at the applicability of swine waste in an anaerobic environment. They found that the high

ammonia levels and low pH are the major common limiting factors in methane fermentation. This is of significance since methane fermentation is often the limiting factor in the anaerobic treatment process. This is especially true in swine waste since it appears that the environment is almost ideal for the syntrophic acetogenic group. Donham et al. also found that volatile acids were beyond toxic levels in eighteen (78%) of the pits tested. Their report and analysis also showed that swine waste has enough nitrogen, carbon, phosphorus, and other nutrients to support the anaerobic bacterial environment. It also appears that there is little concern for inhibition due to toxic metals or high concentrations of salts.

F. Treatment Systems Applied to Swine Waste

1. Introduction

There have been several systems tried on swine waste. Most of the systems have been continuous feed as opposed to batch fed systems. The two categories which can be delineated for treatment systems in the treatment of swine waste are the attached growth systems and the suspended growth systems. These represent two different approaches to treatment.

2. Attached growth systems

Attached growth systems are generally continuously fed anaerobic filters. The filters are either upflow or downflow and can be fully or partially packed with media. A partially

packed filter is usually called a hybrid because the unpacked portion is a suspended growth chamber. The theory is that the media provides a surface upon which the bacteria grow. The feed is introduced to the bacteria as it passes through the filter. The bacteria continue to build up on the filter until it weighs enough that it falls or "sloughs" off the media and is carried out in the effluent. Some effluent of the filter is usually recycled through the filter. Because of their physical make up, filters are only good for low solids waste. Therefore, the swine waste fed to these is usually screened to remove solids so the filter will not get clogged. Table 9 is a summary of all of the results on fixed bed reactors found by the researchers described in the following paragraphs.

In 1983 Oleszkiewicz [36] ran a comparison of suspended and attached growth systems. The attached growth system was called a biofilter and consisted of two meter high columns packed with plastic media (2 cm polypropylene balls). The filters operated in the upflow mode and were continuously fed. The reactors were operated at 23° C. The samples were collected from a farm of 10,000 swine that used tap water to flush the animal stands. The samples were sieved through a sieve with 1.5 mm openings. Removals of COD ranged from 70 to 94% with HRTs ranging from .57 to 14.3 days. The reactors were also tolerant to temperature changes within 18 to 26° C., and pH variations from 6 to 8.

Table 9. Summary of results using fixed film anaerobic reactors

Researcher	Year	Media	HRT	Removal	Temp.
Oleszkiewicz	1983	2 cm balls polypropylene	.57 - 14	70 - 94% (COD)[a]	23
Sorlini	1990	Wood Chips	4 - 5	60%(VS)[b]	30
		Expanded Clay	4 - 5	13.5%(VS)	30
		PVC	4 - 5	64.7%(VS)	30
Hill & Bolte	1988	Felt	1 - 3	30 - 61% (COD)	35
Ng & Chin	1987	PVC	2.4 - 6.3	84 - 96% (COD)	30
Ng & Chin	1988	Expanded Bed	.75 - 6	67 - 90%	30
		Activ. Carbon	.75 - 6	75 - 89%	30
Colleren et al.	1982	Granite Chips	3	69.2%(COD)	30
		Mussel Shells	3	71.4%(COD)	30
		Coral	3	70.2%(COD)	30
		Reeds	3	68.6%(COD)	30
Wilkie[d]	1986	Polypropylene	6	66%(COD)	25
		Cascade	3	53%(COD)	25
		Mini-rings	3	60%(COD)	35

[a]COD = Chemical Oxygen Demand.
[b]VS = Volatile Solids.
[c]These lab reactors were of the expanded bed type.
[d]This study was done on a pilot plant not a lab reactor.

In 1990 Sorlini et al. [37] did a study comparing different packing materials in upflow fixed-bed reactors. The swine waste was diluted and contained 0.5 to 3% total solids. The waste was obtained from a swine breeding facility. The reactors had a volume of 15 liters. The HRT was held at 4 to 5 days. The packing materials used for comparison purposes were wood chips, expanded clay, and PVC. The reactors were

run at a temperature of 30° C. The volatile solids removal was similar in the wood chip and PVC packed reactor (60, 64.7%). However, the expanded clay reactor only had a volatile solids removal of 13.5%. This may have been due to the high pH (8.5) in the reactor stemming from the type of clay used.

A 1988 study was done by Hill and Bolte [38] on synthetic fixed media filters treating screened swine waste liquids. The media used in the reactors was a felt such as commonly used in air conditioning - air handling units. The reactors themselves were 300 liters and operated at HRTs of 5,3,2, and 1 day. Swine waste was obtained from a swine finishing house which used a conventional slatted floor with under floor flushing. The waste was screened through a number 18 mesh sieve. The removals varied with HRT with the best removals occurring at a 5 day HRT and the worst at a one day HRT. Volatile solids removals ranged from 61 to 30%. The study also concluded that the media held up for six months with no apparent degradation and plugging was not a problem.

In 1987 Ng and Chin [39] performed a study on a random-packed anaerobic filter using piggery wastewater. The waste was obtained from a pig farm where the pig wastes were flushed from the pens and channeled into a ditch. The ditch had a significant hydraulic retention time and anaerobic treatment was taking place in the ditch. The ditch waste was screened through a 2 mm sieve to remove large particles. The filters

had an effective volume of 21 liters and were packed with PVC tubing 25 mm long and 12 mm in diameter. The HRT varied from 6.3 to 2.1 hours. The removals by the filters were better at longer HRTs. Volatile suspended solids removals ranged from 99% (6.3 hr HRT) to 90% (2.1 hr HRT). COD removals ranged from 96% (6.3 hr HRT) to 84% (2.1 hr HRT).

Another study by Ng and Chin [40] in 1988 was done on expanded-bed filters. The filters used sand and activated carbon as media. The swine waste was collected as the study above and put through a 2 mm sieve to remove solids. The waste was passed through a 1 mm sieve and collected on a 0.85 mm sieve. HRTs used in the study ranged from 6 days to 14 hours. Removal of COD in the activated carbon study ranged from 75 to 89%. The COD removal of the fourteen hour HRT activated carbon filter was only 40%. The volatile suspended solid removals of the activated carbon filters ranged from 69 to 86%. The fourteen hour HRT filter had a removal of 43%. The sand filter had COD removals from 67 to 90% with the fourteen hour HRT filter achieving a 27% removal. The sand filters had volatile suspended solid removals ranging from 66 to 93% with the fourteen hour HRT filter achieving a removal of 15%.

In 1982 Colleran et al. [41] did studies comparing different medias in an upflow anaerobic filter. The filters all had a hydraulic retention time of three days and were

operated at a temperature of 30° C. The four types of materials used in the filters were granite chips, mussel shells, coral, and reeds. The waste was obtained from a gravity settlement chamber which had a hydraulic retention time of 12 to 15 days. This chamber took care of most of the hydrolysis and yielded a low solids supernatant with a COD load of 50 to 80% of the total pre-chamber COD load. The COD removals were almost equal for all types of media ranging from 76 to 78.1%. The methane yields were even closer, ranging from 1.3 to 1.35 $m^3/m^3/d$.

Another study done by Wilkie and Colleran [42] in 1984 looked at the start up of upflow anaerobic filters using four types of media. The media used for the filters were clay, coral, mussel shell, and plastic pall rings. The reactors were operated at a six day HRT and an empty bed volume of 18 L. The study showed that the fastest start up was the clay reactor which also had the lowest matrix surface to volume ratio. The slowest start up was the coral matrix which had a surface to volume ratio four times greater than the clay.

Wilkie and Colleran [43] did a study in 1986 on a pilot scale upflow anaerobic filter using pig slurry supernatant. The pilot plant had a total volume of 3.5 m^3 and an active volume of 2.6 m^3. The reactor was filled with polypropylene cascade mini-rings. The reactor was operated at HRTs of six and three days. The operating temperature at six and three

days was 25° C. An operating temperature of 35° C was tried on the three day HRT also. The plant was installed in a 700 head piggery attached to the Agricultural College at Atherny, Co. Galway, Ireland. The pigs were fed a grain-rich finishing diet and were housed on a slatted floor with underground storage channels of up to three weeks holding capacity. The waste went to a gravity chamber where it was allowed to settle and the supernatant was taken from this chamber. COD removal for the 6 day HRT was 66% at 25° C. At this temperature, the removal at a HRT of 3 day was 53%. When the temperature was increased to 35° C the 3 day HRT reactor yielded a COD removal of 60%. The study also found that ammonia was at 3000 mg/l and showed no toxicity. This is contrary to what McCarty suggested when he reported absolute toxicity at 3000 mg/l [6]. However, these results parallel others which have shown that anaerobic swine reactors can tolerate higher than normal amounts of ammonia [24].

3. Suspended growth systems

Suspended growth systems are those systems where the microorganisms move around the reactor as a result of mixing. They generally are mixed by some kind of mechanical mixer although mixing can occur through gas recycle. Reactors of the past have been continuously stirred and fed. Of course, the ASBR is not continuously stirred and is batch fed. As mentioned before, batch kinetics and continuous feed kinetics

are different due to the driving force of the F/M ratio. Swine waste is more suitable to suspended growth systems than to fixed growth systems because suspended growth systems can handle very high solids. In suspended growth studies, often times swine waste scraped from the floor can be used without screening. Table 10 describes removals of various types of suspended growth systems at different temperatures.

When Oleszkiewicz [36] did his work with anaerobic filters and swine waste in 1983 he also compared two types of suspended growth systems. The two systems were continuously fed, one with recycle and the other without recycle. The reactor without recycle was called an ANFLOW reactor and was a conventional, mixed, flow-through reactor. The other reactor had recycle and was a gas mixed contact reactor with sludge recycle called an ANCONT reactor. The ANCONT reactor had a volume of 5.1 liters of which 35% was a centrally located gas mixed compartment. The remaining 65% of the volume was an upflow sludge blanket biofiltration zone. The ANFLOW reactor had a volume of 3 liters and was operated at a temperature of 23° C. The ANCONT reactor was also operated at 23° C. The waste was a screened waste from a farm of 10,000 pigs. The farm practiced tap water flushing of the animal stands. The waste was sieved through a 1.5 mm sieve. The HRT of the ANFLOW reactor was at 9 days at optimum conditions. The COD removal at this point was 63%. The ANCONT operated most

Table 10. Comparison of suspended growth studies done by various researchers

Researcher	Year	Type	HRT d	SRT d	Removals %(type)	Temp. °C
Oleszkiewicz[a]	1983	CSTR	9	9	63(COD)	23
		ANCONT	3.5	20	85(COD)	23
Hill & Bolte[b]	1985	CSTR	10	10	32.3(VS)	55
Hill & Bolte[c]	1985	CSTR	10	10	60.9(VS)	55
Hill & Bolte[d]	1986	CSTR	15	15	64.1(VS)	35
			10	10	65.6(VS)	35
			7	7	57.4(VS)	35
			5	5	53.4(VS)	35
Bolte et al.[e]	1986	SPAG	10		64.3(VS)	35
			5		50.7(VS)	35
			3		46.5(VS)	35
			2		36.0(VS)	35
			5		66.9(VS)	55
			3		60.8(VS)	55
			2		48.9(VS)	55
			1		40.6(VS)	55
Zhang et al.[f]	1990	CSTR	14	14	66.0(VS)	35
Fisher et al.[g]	1984	CSTR	15	15	35-60(VS)	35
Wun-Jern[h]	1988	ASBR	15	338	32.5(COD)	28
			10	152	32.4(COD)	28
			5	51	32.2(COD)	28
Summers et al.[i]	1980	CSTR	10	10	36.0(TS)	35
			7	7	36.0(TS)	35
			5	5	32.0(TS)	35
			3	3	8.0(TS)	35
Pidaparti[j]	1991	ASBR	6	13-107	74-87(VS)	35
			6	20-125	77-92(VS)	25

[a]Waste used was sieved through 1.5 mm sieve.
[b]Waste used was scraped waste from floor.
[c]Waste used was flushed-screened waste.
[d]Waste used was screened-flushed.
[e]Waste used was screened-flushed.
[f]Waste used was from manure scrapers under slotted floor.
[g]Waste used was from concrete feeding floor.
[h]Waste used was raw wastewater from commercial pig farm.
[i]Waste used was from an under slat-channel of a pig farm.
[j]Waste was collected from gutter of a finishing house.

efficiently at a HRT of 3.5 days and maintained a SRT of 20 days at this HRT due to sludge recycle. COD removals of the ANCONT reactor were more than 85% at optimum conditions. The advantages of sludge recycle is obvious in this study. The sludge recycle allows operation of the reactor at a HRT of 3.5 days instead of 9 days. This allows a significant decrease in the design volume of the reactor which represents a significant capital cost savings.

A study was done by Hill and Bolte [44] in 1985 to compare thermophilic treatment of surface scraped and flushed-screened swine waste. The study was done on two 190 l bench scale reactors operated at a ten day HRT and at a temperature of 55° C. The reactors were continuously mixed and fed. The scraped waste was collected from an open sloping concrete floor of a swine production facility. Due to the slope, the urine portion of the waste was lost. The flushed waste was obtained from a slatted floor and was screened before use. After the flushed-screened run the reactors were fed waste which was not screened. The scraped waste had a volatile solids destruction of 32.3% and a COD reduction of 29.9%. The flushed-screened waste had a volatile solids destruction of 60.9% and a COD reduction of 53.8%. When the flushed-screened portion of the experiment ended, the whole waste was tried in the reactors. The reactors failed when the whole waste was used as food. Both the ammonia and volatile acid levels

increased, making it impossible to tell which of these caused the failure. Some information found was that the flushed-screened waste had ammonia levels one-third that of the scraped waste. Also, the specific methane productivity of flushed-screened waste is approximately 1.2 times that of the scraped waste (.66 L CH_4/g VS destroyed vs. .56 L CH_4/g VS destroyed).

In 1986 Hill and Bolte [45] followed up on their earlier work by anaerobically treating screened-flushed waste at thermophilic temperatures. This time they tried different loads and different HRTs. The study used two 380 liter reactors made from stainless steel. The waste used for the study came from a conventional slatted floor-underfloor gutter flushing system. The flushed waste was reconcentrated using a 46 cm vibrating separator operated with a 60 mesh at a rate of 457 l/min. The reactors were operated at HRTs of 15, 10, 7 and 5 days and were operated as conventional, continuously fed reactors at 55° C. The strength of the waste remained the same but the loading rate on the reactor changed depending on the HRT. The volatile solid reductions in the reactors were 64.1, 65.6, 57.4, and 53.4% for HRTs of 15, 10, 7, and 5 days respectively. It was speculated that hydrolysis became rate limiting at the lower HRTs because the methanogens showed no inhibition. The highest load used in this study was at a five day HRT and was 12.66 g VS./L - Day.

Bolte et al., continued their work in 1986 [46] once again using anaerobic treatment of screened pig waste at thermophilic and mesophilic temperatures. In this study, they used a new reactor technology that combines attached growth and suspended growth systems. They used a light weight polyurethane foam and reticulated nylon pads. These support materials are suspended in the reactor liquor by mechanical agitation resulting in a "semi-fluidized bed" requiring low levels of energy input to achieve particle suspension. The media allows the particles to be attached but protects them from hydraulic shearing forces. This new technology is called a suspended particle-attached growth (SPAG) reactor. The reactor was mixed by recirculating its contents through a controlled temperature water bath. The study was done at 35° C and 55° C and employed continuously stirred and fed SPAG reactors. The four reactors had an empty bed volume of 5 liters. The mesophilic part of the study ran with HRTs of 10, 5, 3, and 2 days. The thermophilic portion had HRTs of 5, 3, 2, 1 days. The mesophilic reactor was operated with volatile solids removals of 64.3, 50.7, 46.5 and 36.0% for HRTs of 10, 5, 3, and 2 days, respectively. The thermophilic reactor was operated with volatile solids removals of 66.9, 60.8, 48.9, and 40.6% at HRTs of 5, 3, 2, and 1 day, respectively. The thermophilic reactors outperformed the mesophilic reactors. It was also found that the SPAG reactors have a much faster

start up than conventional CSTRs.

A study conducted in 1990 by Zhang et al. [47] looked at a field scale application of an anaerobic reactor on a swine farm. The farm had a capacity equivalent to a farrow-finishing unit marketing 3000 hogs per year. The waste consisted of manure from manure scrapers located under a partially slotted floor. An anaerobic lagoon followed the reactor. The reactor was 199 m^3 in volume and was operated at 35° C. The reactor was batch fed when the facility was flushed and was continuously stirred. The reactor was operated for twenty months at a retention time of fourteen days. Over the twenty months of operation the average VS load was 23,754 mg/l/d. The volatile solids removal was 66%. The study also looked at the use of the methane as an energy source.

Fischer et al., in 1984, studied how increasing volatile solids of swine waste affects anaerobic treatment. They used a 0.42 m^3 CSTR reactor operated at a 15 day HRT and a temperature of 35° C. The manure used in the experiment was collected from a concrete floor from finishing hogs. The manure was diluted with tap water to desired loadings. Volatile solids concentrations tried were 60.3, 68.1, 81.9, 87.9, 97.1, and 108 g/l. Removals obtained were 60.3, 53, 55.1, 35.7, 35.1, and 0% respectively. The zero result suggested reactor failure at a VS concentration of 108 g/l. The study showed that as the volatile solids load is increased

the removals tend to drop.

In 1989 Wun-Jern [48] did a study on piggery wastewater using a sequencing batch reactor in an anaerobic mode. This technology is similar to the ASBR developed earlier at Iowa State University. A sixteen liter reactor was used for the experiment. It was operated at HRTs of 15, 10, and 5 days. The reactor was fed four times a day. The volume of feed depended on the HRT. The reactor was run at room temperature (28^0 C). The sequencing reactor allowed large solids retention times with the low hydraulic retention times. Solid retention times were 338, 152, and 51 days for HRTs of 15, 10, and 5 days, respectively. The swine waste used in the study was obtained from a commercial pig farm on a weekly basis. It was drawn from a slump pit at the farm. Influent COD loading rates were .4, .7, and 1.77 g/l/d for HRTs of 15, 10, and 5 days, respectively. The COD removals for the above loads were 32.5, 32.4, and 32.2%, respectively.

A 1980 study by Summers and Bousfield [49] looked at a CSTR pilot plant at various HRTs. The reactor was continuously fed and consisted of a 100 liter stainless steel vessel heated to 35° C with an external water jacket. The reactor was operated at HRTs of 10, 7, 5, and 3 days. The swine waste used was taken from the under-slat channel of a piggery housing fattening pigs. The waste contained feces and urine, but contained no wash water. The volatile solids

loadings were 3.3, 3.7, 3.7, and 3.5 g/l/d for HRTs of 10, 7, 5, and 3 days, respectively. The removals for the respective HRTs and loadings were 36, 36, 32, and 8.6%. It was found that at the three day HRT the removal was not very good. The plant was also operated at a variety of temperatures ranging from 25° C to 44° C. The percent reduction in volatile solids in this range did not change appreciably with the highest removal at 25° C. This confirms studies by others that show reduction of pig waste at 25° C and 35° C is virtually the same if the solids retention time is high enough.

Pidaparti [20] did a study using an anaerobic sequencing batch reactor (ASBR) at Iowa State University in 1991. The study was done on a variety of volatile solids loads at temperatures of 35° C and 25° C. The reactors were twelve liters in volume and mixing was accomplished by gas recirculation. The reactors were operated at a six day HRT. The reactors were batch fed once per day and the solids were allowed to build up due to the solids capture nature of the reactor. The swine waste was collected from Iowa State University's Swine Nutrition Farm. The waste was collected from a finishing facility. At 35° C volatile solids loadings ranged from 1 to 5.38 g/L/day and removals ranged from 87 to 74%. At 25° C the loadings ranged from 1.1 to 5.67 g/L/day and removals ranged from 92 to 77%. The results at the two temperatures were very close, verifying once again that swine

waste treatment at 25° C and 35° C proceeds at almost equal rates as long as the SRT remains high enough at both temperatures.

III. EXPERIMENTAL SETUP

A. Reactor Configuration

These experiments made use of three, identical, twelve-liter reactors which were mixed by gas recirculation. Figure 6 shows the complete system. A typical reactor is shown in Figure 7.

The reactors were cylindrical and were constructed from 12 mm (0.5 in) thick Plexiglas. The reactors had an inside diameter of 127 mm (5 in), outside diameter of 152 mm (6 in), and a length of 912 mm (36 in). The reactors also had a flange on the top and bottom which had a diameter of 228.6 mm (9 in). The total reactor volume was 14 liters of which 12 liters was the working volume.

The top and bottom of the reactors were fitted with a plate with the same outside diameter as the flange of the reactor. The plates of each reactor were also made of Plexiglas which was 12 mm (0.5 in) thick. The plates were drilled with twelve holes equally spaced around the circumference. These holes were 10 mm (3/8 in) in diameter and matched the twelve holes of equal diameter drilled into the flanges of the reactor. The plates were attached to the reactor using twelve 10 mm (3/8 in) bolts each with a pitch of 30°. The bolts were secured with two nuts and two washers, one nut and washer on the top of the plate and the other under

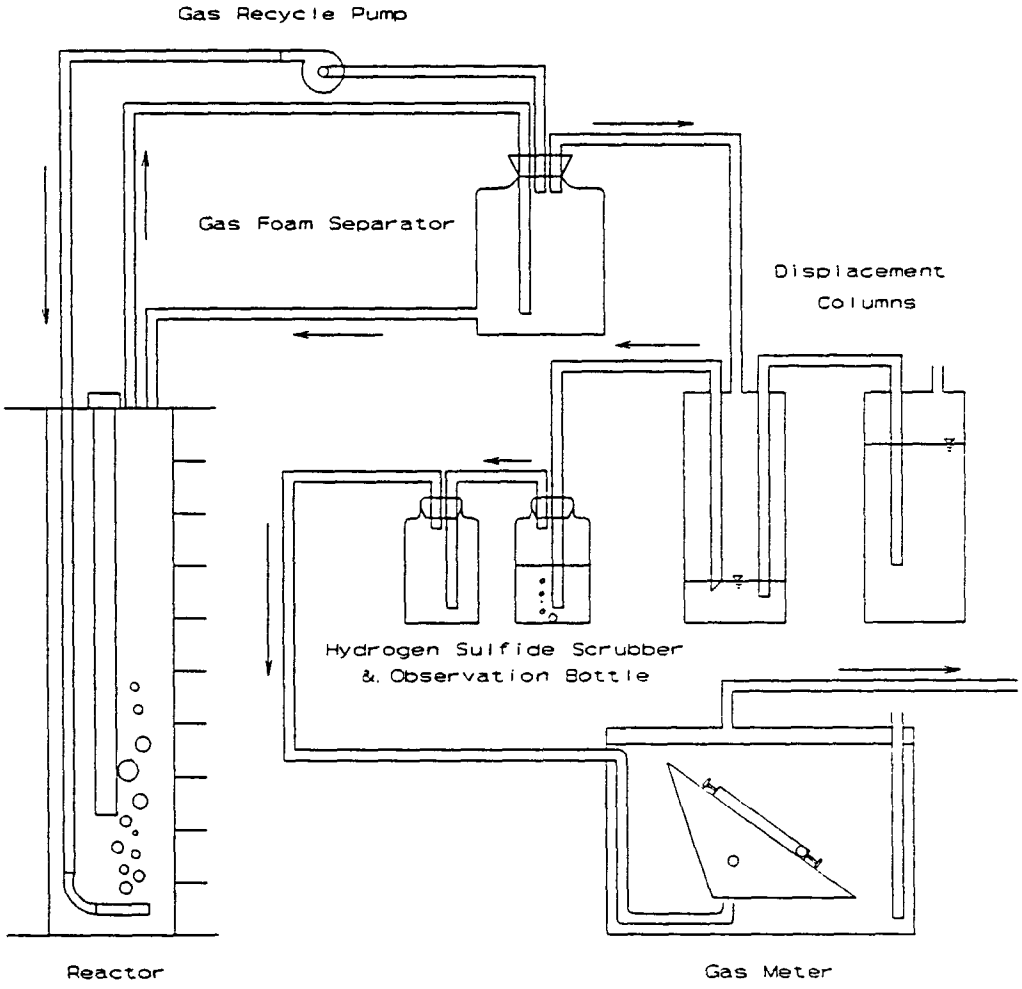

Figure 6. Schematic of reactor system setup

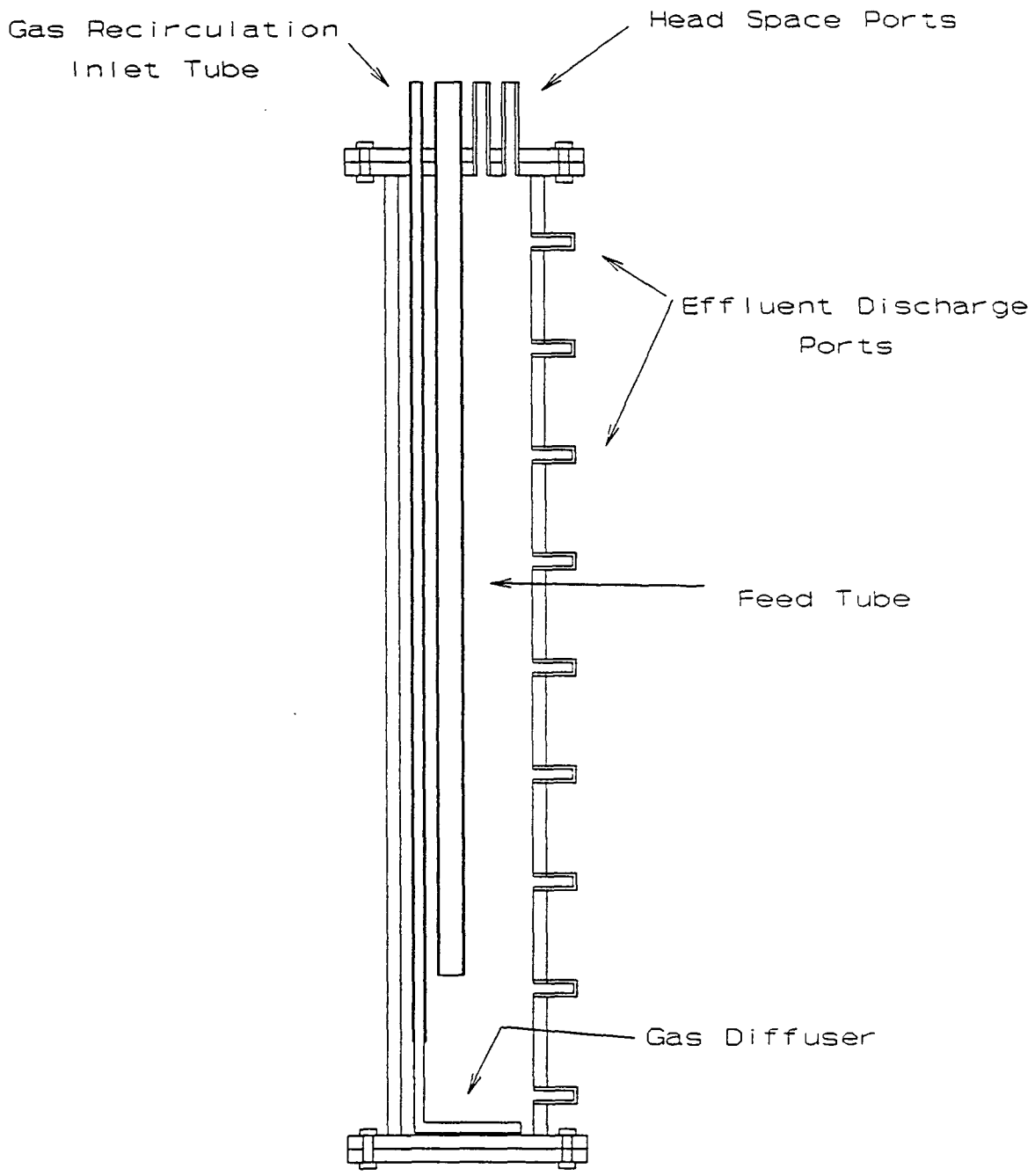

Figure 7. Schematic of a typical reactor

the flange of the reactor. Circular grooves were made on the bottom of the plate and on the top of the flange for placement of an O-ring. The O-ring was added to the top and bottom plate to insure a leak proof seal.

Each reactor had nine effluent decant ports for the withdrawal of reactor contents. The first one was located 50 mm (2 in) from the top of the reactor. The remaining eight were placed at an equal spacing of 102 mm (4 in) along the length of the reactor. Each port was 25 mm (1 in) long and had an outside diameter of 16 mm (5/8 in). The inside diameter of each port was 10 mm (3/8 in). The ports were fitted with a piece of Tygon tubing (R-3603) and clamped off during operation to prevent leaks.

The top and bottom plates also had ports to allow for the operation of the ASBR process. The bottom plate had one port in its center. The port was 16 mm (5/8 in) outside diameter, and 10 mm (3/8 in) inside diameter. The port had a length of 25 mm (1 in). The port was used to empty the reactor contents when the study was finished. The port was fitted with a piece of Tygon tubing (R-3603) and clamped during normal reactor operation. The top plate was fitted with four ports. Two ports were 25 mm (1 in) in length, with an outside diameter of 16 mm (5/8 in), and an inside diameter of 10 mm (3/8 in). Both ports entered the head space of the reactor. One of these ports was used for the gas and foam exit. The other port was used for foam return from the foam separation

apparatus. However, when foam was not present, this port served as an additional exit port for biogas. The third port was fitted with a stainless steel rod 12 mm (0.5 in) in diameter which reached to the bottom of the reactor. The rod was used for the return of recirculated gas used for mixing the reactor. At the start of the experiment the rod ended in a porous diffuser. From the middle to the end of the study the rod ended in a copper coil diffuser. This diffuser replaced the porous diffuser which malfunctioned. The tube was fitted to the top plate using a Swagelok fitting. A fourth port was added to the top plate to hold a tube for batch feeding. The feed tube was 25 mm (1 in) in diameter and was constructed of stainless steel. The tube extended into the reactor to a depth of 152 mm (6 in) above the bottom plate of the reactor. The fitting for the tube was similar to that of the gas recirculation tube, but was capped to prevent the reactor contents from coming up through the tube. All the ports in the top are shown in Figure 7.

B. Gas System

The gas, and any foam contained in the gas, exited the reactor through the top and was conveyed to gas-foam separation apparatus through Tygon tubing (R-3603). The tubing had an inside diameter of 3/8 in and a wall thickness of 1/16 in. The gas-foam separation apparatus (Figure 8) was needed to prevent foam from entering the system and clogging

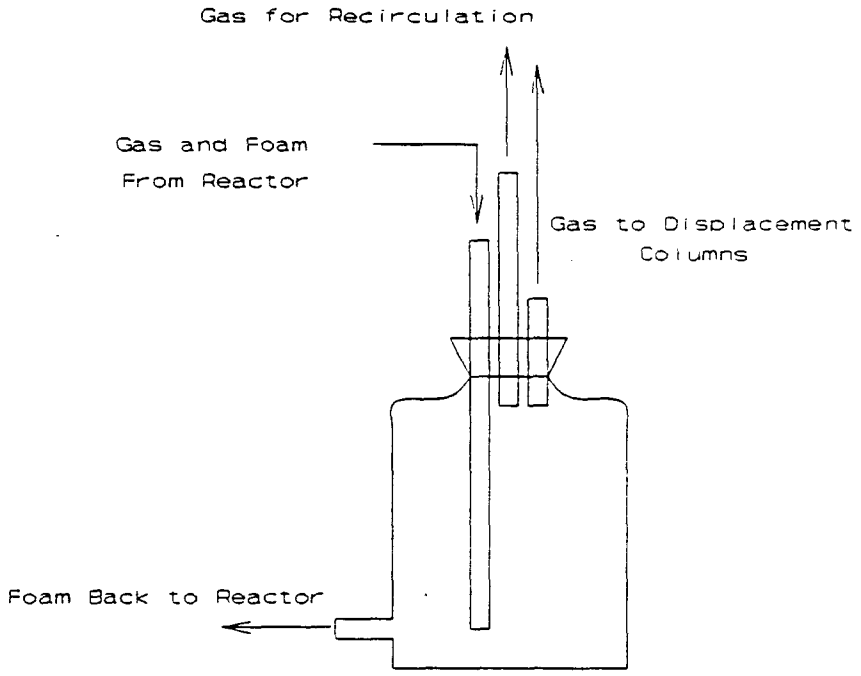

Figure 8. Gas-foam separator bottle

tubing and diffusers. The apparatus consisted of a four liter aspirator bottle with a discharge port located on the bottom. The top of the bottle was stoppered with a number 10 rubber stopper and sealed with silicon to prevent gas leaks. The stopper was drilled with three holes. The holes were filled with three 10 mm (3/8") pieces of glass tubing. Two of the tubes were short pieces about 50 mm (2 in) long. The other tube extended to within 25 mm (1 in) of the bottom of the aspirator bottle. The gas and foam entered the bottle through the long tube. The gas then separated from the foam and was carried through the two tubes at the top of the aspirator bottle. One tube on the top went to the pump for gas recirculation while the other went to the water lock. The port located on the bottom of the aspirator bottle was fitted with tubing and went to a gas effluent port located on the top plate of the reactor. This port was used to carry the foam out of the bottle and back into the reactor. Gas and foam were removed by the same size and type of tubing that brought it to the system.

The apparatus allowed the gas and foam to enter the bottle. Gas for recirculation was pumped from the top of the bottle. All the additional gas produced would leave the bottle through the other port on top. As the gas was leaving through the top of the bottle the foam would sink to the bottom of the bottle and be carried away via the port on the bottom of the aspirator bottle to the reactor.

The gas recirculation system (Figure 6) consisted of a 6600 rpm Masterflex peristaltic pump with speed controller. An 8 mm (0.31 in, size 18) inside diameter Masterflex neoprene hose was used inside the pump head. The pump was capable of pumping fluids at flow rates of 22.8 to 2280 mL per minute at a pressure of 20 psi. The stone diffuser was a bubble disk with a diameter of 51 mm (2 in). The diffusers were purchased at a pet store and are typically used in aquariums. Later in the experiment a course bubble diffuser was designed and installed to prevent diffuser clogging and inadequate mixing. It was made of 12 mm (0.5 in) copper tubing with an outside diameter was 102 mm (4 in). It was drilled with 8 equally spaced 1.6 mm (1/16 in) holes in the top of the tubing around the circumference of the diffuser. The holes were drilled at a 45° angle toward the center of the ring to ensure adequate mixing. A Chronotrol controller controlled the mixing times of the recirculation system. It was a four outlet 10 program timer that allowed flexible timing for controlling the gas recirculation system. All three gas recirculation pumps were connected to the timer and were operated this way.

The water lock (Figure 9) was used to maintain a positive constant head on the system as well as to prevent the addition of air to the reactor upon effluent withdrawal. The lock consisted of two Plexiglas tubes. The tubes had an outside

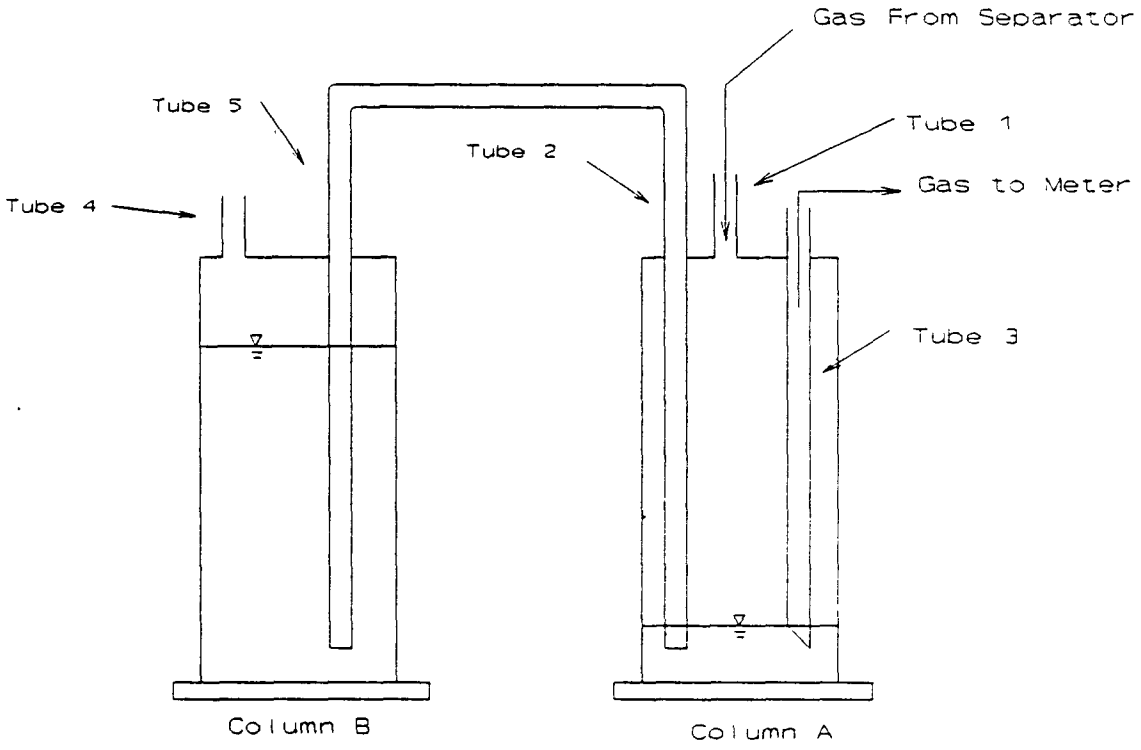

Figure 9. Displacement columns

diameter of 152 mm (6 in), an inside diameter of 127 mm (5 in) and a length of 292 mm (11.5 in). A 12 mm (0.5 in) thick circular plate was glued to the top of the tube. The plate had the same diameter as the tube. The bottom of the tube consisted of a 12 mm (0.5 in) thick square plate glued to the tube. The plate had a side length of 203 mm (8 in). The top plate of column A had three tubes. Each tube had an outside diameter of 16 mm (5/8 in) and an inside diameter of 10 mm (3/8 in). The shortest tube (#1) was 38 mm (1.5 in) was placed at the center of the column. The longest tube (#2) was 330 mm (13 in) long and placed 51 mm (2 in) from the center. The other tube (#3) was 254 mm (12 in) long and placed directly opposite #2 51 mm (2 in) from the center. The top plate of column B had two tubes. Each tube had the same inside and outside diameter as in column A. Tube 4 was 38 mm (1.5 in) long and placed 25 mm (1 in) from the side of the plate. Placed directly opposite 4 and also placed 35 mm (1 in) from the side of the top plate was tube 5. Tube 5 had a length of 330 mm (13 in).

Gas produced by the system entered column A through tube two through Tygon tubing, as described previously. The gas pushed the liquid in column A to column B via tube two in A and tube five in B. Tubes two and five are connected by a siphon. Tube four allows column B to remain at atmospheric pressure which allows the liquid from A to enter B. The liquid in A is pushed out until tube three is exposed to the

gas. Tube three goes to an observation bottle, a hydrogen sulfide scrubber, and eventually to the gas meter. This route has less head than in the water lock thus the gas exits through this tube allowing liquid in B to flow back to column A blocking tube three until enough gas enters A to expose tube three. The positive head applied is the difference in levels between column A and column B.

The gas was conveyed from the water lock to an observation bottle and H_2S scrubber (Figure 10). The observation bottle as well as the scrubber bottle were made from one liter glass bottles. Both bottles were fitted with rubber stoppers and sealed with silicon to prevent leaks. Both bottles had two holes drilled in the stopper. The holes were filled with two pieces of glass tubing. The tubing had an outside diameter of 10 mm (3/8 in). One tube extended to the bottom while the other extended 25 mm (1 in) below the stopper. The observation bottle was filled half full with water. The gas entered the observation bottle through the long tube and bubbled through the liquid. This apparatus was used to give a visual indication of gas being produced. The gas left the bottle through the short tube where it entered the scrubber. The scrubber bottle was identical to the observation bottle except it was filled with sponges soaked in ferric oxide. The ferric oxide reacted with any hydrogen sulfide in the gas. The hydrogen sulfide is removed from the

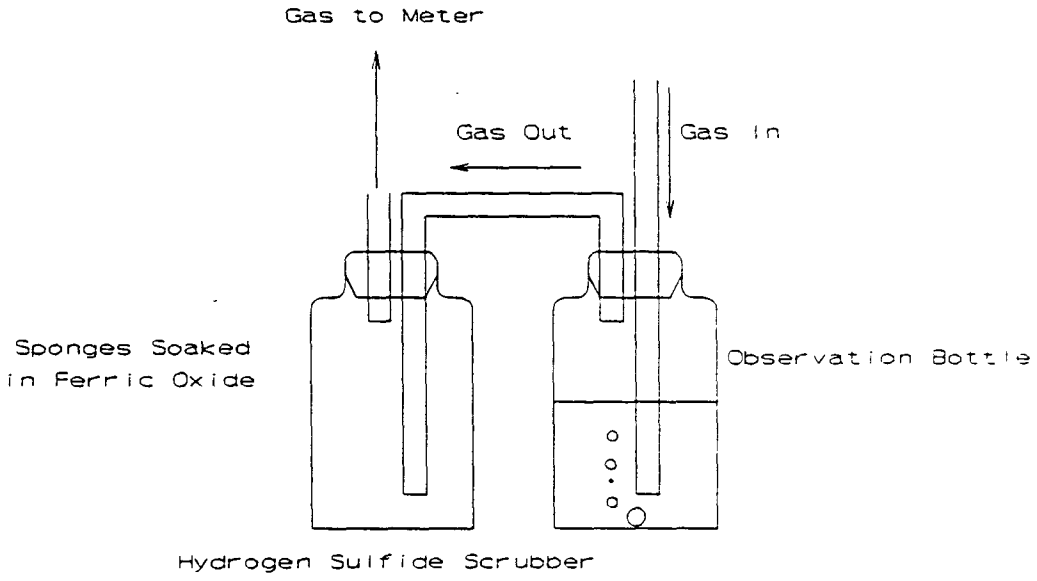

Figure 10. Hydrogen sulfide scrubber and observation bottle

gas to prevent any damage to the gas meters. The gas traveled the same way through this bottle as the observation bottle. It entered the long tube and was conveyed from the bottle by a short tube.

The gas left the scrubber bottle and went via Tygon tubing through a gas sampling port. The gas sampling port was made from a glass tube 63 mm (2.5 in) long with an outside diameter of 25 mm (0.5 in). The tube was fitted on both ends with a 10 mm (3/8 in) glass tube to allow connection to the system with Tygon tubing. The center of the tube contained a 16 mm (5/16 in) tube with a rubber septum. The septum allowed a needle to be repeatedly inserted without any leaking. The needle used was connected to a syringe and the gas sample collected was analyzed in a gas chromotographer.

The gas was measured with a wet tip gas meter produced from the Rebel Wet Tip Gas Meter Company (Figure 11). The meters work on the principle of buoyancy. Gas enters through the inlet pipe where it is collected under the tipping float. When enough gas is collected, the float tips due to the force exerted by the gas wanting to escape solution. When the tipping float lifts over half way, a spherical weight in a track moves horizontally along the float which completes the tip. By moving the adjustment screws the amount of gas needed to tip the meter can be increased or decreased. Each tip of the meter makes an electrical connection which counts on a digital counter. The meters are calibrated such that each tip

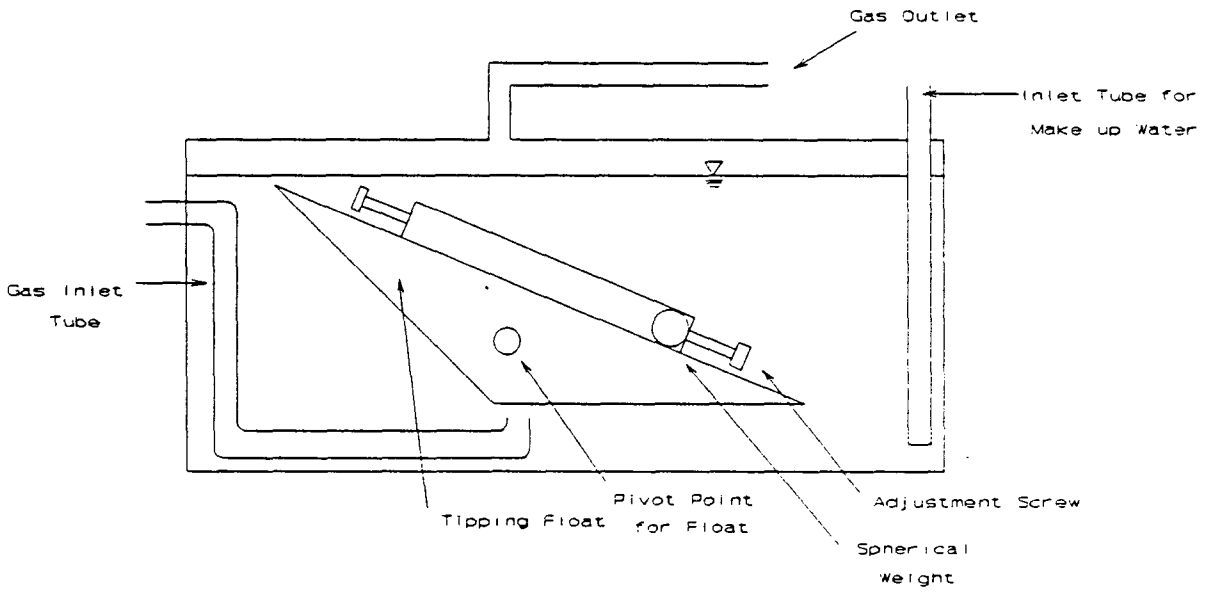

Figure 11. Tipping gas meter

constitutes .1 liters. After the gas tips the float it is released through the gas outlet and vented out of the building.

IV. EXPERIMENTAL PROCEDURES

A. Swine Waste Preparation

Swine waste was obtained from the Iowa State University Swine Nutrition Farm. The location from which the feed was obtained was changed on August 8, 1991. The reactors were given two months before data were taken to adapt to the new waste. The farm is a shower-in and shower-out facility which houses pigs in the weight range of 100 to 200 lbs. The feed was collected from the floor of the pens. The feed was put into two, five gallon buckets and transported to the lab to be chilled until needed. Waste was obtained from the farm about once every four weeks.

In the lab the waste was diluted by a factor of four with water and blended in a Corning industrial strength blender. Dilution was necessary due to the inhibitory nature of high concentrations of ammonia in swine waste. The blender was placed on a scale and the scale was tared. Waste in an amount of 1500 grams was placed in the blender and water was added to a volume of five liters. The waste was blended until it was of an even consistency. Enough waste was blended to fill two, twenty liter containers. The containers were stored in a refrigerator for direct use as feed to the reactors.

The strength and solids content of the waste was checked each time data were taken. The chemical oxygen demand(COD) test was the test used for strength. The Closed Reflux method

was used, as given in Standard Methods [79]. The amount of oxidizer used limits the test to a maximum detection level of 340 mg/l of COD. Due to this, the feed was diluted 250 times before running the test. Another test run regularly on the feed were the total solids series. The total solids test that was run is the one which is described in the laboratory methods section of this thesis. A characterization of the waste is given in Table 11.

Table 11. Characteristics of Swine Waste

Characteristic	Raw Manure	Blended Feed
Total Solids. %	18 - 26	7.7 - 11.2
Volatile Solids, %	15 - 24	6.5 - 8.6
TCOD, g/L	210 - 340	65 - 108
Five day BOD, g/L	73 - 80	22 - 24
TCOD/BOD_5 ratio	1.8 - 2.0	2.9 - 4.5
pH	4.7 - 5.8	6.2 - 6.4

B. Reactor Operation

The reactors were initially operated by Surya Pidaparti [20] from September 1990 to June 1991. His testing on the swine waste was done at a 6 day HRT(hydraulic retention time) and temperatures of 35 and 25° C. The temperature of the reactors was lowered to 20° C in June 1991. The lowered temperatures caused reactor problems so the HRT of reactor one

was raised to nine days. Reactor two was raised to twelve days, while reactor three remained at a six day HRT. The reactors were operated at the three HRTs for one year. The reactors were batch fed daily at volatile solid loadings ranging from 1 to 5 g/l/d. The reactors were allowed to achieve a quasi steady-state before samples were taken and analyzed at a given load. The feed rate was then increased to the next higher load. Quasi steady-state was defined by a constant and steady production of methane (+/- 5%). This state generally takes about three HRTs to achieve after changing to a new load. A data point was calculated using the samples from the reactor when it was in quasi steady-state. This consisted of a complete determination of all performance parameters for the reactors. Each data point consisted of the average of a set of three samples. The samples were taken and analyzed on three consecutive days.

The reactors were batch fed once per day. The amount of swine waste fed was based on the COD of the waste. Swine waste was fed until the loading requirement was met. Water was then added to make up the balance of the feed to meet the HRT requirements. The reactors were operated in a sequencing batch mode. The lengths of the cycles were as follows:

Feed Phase	0.25 Hours
React Phase	21.5 Hours
Settling Phase	2.0 Hours
Decant Phase	0.25 Hours

At all loads and during the one year operation various parameters were measured to determine the performance and the health of the reactors. Reactor test parameters and testing frequency are listed in Table 12.

Table 12. Testing parameters and Frequency

Test Parameter	Frequency
Gas production	daily
pH	3/wk
Gas composition	2/wk
Alkalinity	at data point
Volatile Fatty Acids	at data point
TCOD removal %	at data point
Solids removal	at data point

When the temperature was lowered, the HRT on all three reactors was six days and the TCOD load on the reactors was 3 g/l/d. After one month of operation the reactors started failing. The gas production and pH decreased while the total volatile acids increased. The HRT was increased on reactor one to nine days and on reactor two to twelve days. The HRT on reactor three remained at 6 days. Simultaneously, the load on all three reactors was lowered to a TCOD load of 1 g/l/d. The reactors were allowed to reach equilibrium. Table 13 is a history of when the TCOD loads on the reactors were raised.

Table 13. History of TCOD loads

TCOD Loads G/L/D			
	HRT		
Date	6 days	9 days	12 days
07/01/91	1.00	1.00	1.00
09/16/91	1.57	1.04	1.04
10/31/91	2.06	1.56	1.03
12/21/91	3.21	2.56	1.94
02/01/92	3.74	2.81	2.81
03/28/92	4.03	3.69	3.27
04/30/92	5.32		

C. Daily Reactor Procedures

The biogas production for the previous twenty four hours was measured by a tipping gas meter and recorded daily. The reading was taken close to the same time every day. This had to be done because the rate of gas production varies with time because of the continuously changing reaction rates due to the variation in the food to microorganism ratio. The volume of gas was corrected to standard temperature and pressure (760 mm HG & 273° K). The pressure was recorded daily from a mercury barometer. The barometer had an accuracy of +/- .1 mm. The temperature was recorded with a mercury thermometer capable of measuring temperatures between -20° C and 100° C(+/- 0.1° C). The following equation based on the ideal gas law was used to calculate the volume of gas at STP:

$$V_s = \frac{(V_{R_2} - V_{R_1})\,(P_R)\,(T_S)}{(P_S)\,(T_R + 273)}$$

where:

V_s = Volume of gas produced daily (liters);

V_{R1} = Volume of gas produced previous day (liters);

V_{R2} = Volume of gas produced current day (liters);

P_R = Daily pressure reading (mm Hg);

T_S = Temperature at STP (273° K);

P_S = Pressure at STP (760 mm Hg);

T_R = Daily temperature at gas meter (°C).

Besides being corrected for STP, the gas meters were calibrated with a known volume of gas at ambient temperature and pressure. The meters needed to be calibrated because the meter is under a water column and this head affects the gas volume.

The supernatant was decanted daily from the reactor. The amount of supernatant wasted depended on the HRT of the reactor. The volume of the reactor divided by the HRT in days is the volume of the daily supernatant according to the following equation:

$$\frac{V_R}{HRT} = V_S$$

where:

V_R = Volume of reactor (liters);

HRT = Hydraulic retention time (days);

V_S = Volume of supernatant wasted (liters/day).

The supernatant was withdrawn from the best port available.

There was usually some foam at the top of the reactor and a definite interface toward the bottom. The supernatant was withdrawn below the foam and above the interface. This usually meant the third or fourth port from the top.

If effluent samples were needed they were obtained at this point. Samples for pH, alkalinity, ammonia, and total volatile acids were obtained from the supernatant and put into a 200 ml beaker. The beaker was covered with parafilm to protect the integrity of the sample. Part of this was immediately tested for pH. The samples for TCOD and SS were taken from the total sample of supernatant. All supernatant was put into a two liter container and mixed. A composite sample of 200 ml was withdrawn and put into a 200 ml glass beaker. A composite sample was used due to the inconsistencies which may occur in these particular parameters at any one point in the supernatant.

The volume of feed was determined by the TCOD loading on the reactor and the strength of the feed. The strength of the feed varied every time it was collected. After the swine waste was added, tap water was added to make up the difference between the supernatant wasted and the swine waste fed. The make-up water also served to clean out the feed tube to insure all the feed reached the active liquid in the reactor.

The reactor gas recirculation mixing phase was started and allowed to run for thirty minutes. After mixing for thirty minutes to insure a complete mix, mixed liquor samples

were taken as needed. The samples were always collected from about the middle of the reactor at the fifth port from the top of the reactor. The sample was collected for use in TCOD and SS. Gas collection for composition analysis was collected from the sampling port. The gas was collected in a syringe and immediately went to the gas chromatograph for analysis.

D. Laboratory Analyses

1. pH

The pH was monitored about three times a week. The pH measurements were done using an Altex Instruments Model 4500 digital pH meter. The accuracy of the meter was +/- .01 pH units. The electrode used was a Markson electrode. Care was taken with the sample to absolutely minimize contact with the atmosphere which causes carbon dioxide to leave the sample and the pH to increase. This meant taking the sample from the reactor immediately to the pH meter and taking a reading. The pH meter was calibrated before every use with two buffer solutions one of pH 7.00 and the other at pH 4.01.

2. Total alkalinity

Total alkalinity was run at every data point and in between when necessary to analyze reactor health. Total alkalinity was measured using methods outlined in Standard Methods [79]. Almost all of the alkalinity in the reactor was of the bicarbonate form. The total alkalinity was calculated from the following equation in Standard Methods [79]:

$$Total\ Alkalinity (as\ mg/l\ CaCO_3) = \frac{50,000 \times N \times ml\ H_2SO_4}{(ml\ of\ sample)}$$

where:

N = normality of sulfuric acid (.1N).

3. Total volatile acids

The total volatile acids of the supernatant were run at every data point and at times to check for reactor imbalances. The distillation method was used, as outlined in Standard Methods [79]. The volatile acids were calculated with the following formula from Standard Methods [79]:

$$Total\ Volatile\ Acids\ (mg/L\ of\ acetic\ acid) = \frac{60,000 \times ml.\ NaOH \times}{mL.\ sample}$$

where:

N = Normality of sodium hydroxide;

ml. NaOH = Volume of NaOH used in titration (ml);

ml. Sample = Volume of sample used in analysis (ml).

4. Gas analyses

Gas analyses were done twice a week at every data point. The gas was analyzed for N_2, CO_2, and CH_4 using a gas chromatograph (GC) with a thermal conductivity detector. The operating conditions and parameters for the GC are represented in Table 14.

A standard was used to determine the various percentages of gas in the biogas. The standard consisted of 70% CH_4, 25% CO_2, and 5% N_2 (concentrations +/- 0.5%). The composition of

Table 14. Gas chromatograph operating conditions and parameters.

Gas Chromatograph	Hewlett Packard 5730A
Column	6 ft x 3 mm I.D. stainless steel
Packing	Poropak Q, 80/100 mesh
Temperature	Ambient
Carrier gas	Helium
Flow Rate	30 mL/min.
Column pressure	60 psig
Detector	Thermal conductivity
Temperature	200° C
Bridge Current	150 mA
Sensitivity	10 mA
Injection point temperature	100° C
Sample Size	0.9 mL

the standards was made to resemble the typical biogas composition. Four runs of standard gas were used for each GC analysis. The gas samples were obtained from gas sampling ports in the gas line. The ports contained a rubber septum to allow for many withdrawals while maintaining a leak proof seal. The samples were taken with a Hamilton Gas tight #1001TLL syringe. The syringe was purged with biogas three times before collecting a sample. The volume of the sample used was 0.9 ml. When a sample was drawn, more than 0.9 ml

were drawn in the syringe. Just before injection into the GC the excess gas was released so the sample volume was 0.9 ml. This was done to prevent detection of any air which might have accumulated in the tip of the syringe. Two samples were run for every reactor and the results averaged. Peak identification and integration was done using Maxima data station software.

5. Chemical oxygen demand

Chemical oxygen demand was performed for every sample in a data point. The test was performed on the feed, effluent, and the mixed liquor. It is the measure of the oxygen equivalent of the organic material in a sample which can be oxidized by the strong chemical oxidant, potassium dichromate, in an acid solution. The TCOD was measured using the Closed Reflux Titrimetric Method(method #508B) [79]. All samples were run in duplicate and the results averaged. The formula used to calculate the COD is as follows:

$$COD \text{ as } mg\ O_2/L = \frac{(A - B)\ x\ 8000\ x\ M}{(ml\ sample)}$$

where:

A= ml FAS titrant used for blank;

B= ml FAS used for sample;

M= molarity of FAS titrant.

6. Solids

Solids analyses were performed on the mixed liquor, effluent, and feed of the reactor. The entire solids series

was run including suspended, total, and volatile solids. Total solids were run according to Standard Methods [79]. The sample sizes were 50 ml for effluent and 10 ml for feed and mixed liquor. The total solids and total volatile solids were determined from the difference in weights and the volume of the sample by the following formula:

$$\textit{Total Volatile Solids (mg/L)} = \frac{B - C}{\textit{volume of sample (L)}}$$

$$\textit{Total Solids (mg/l)} = \frac{B - A}{\textit{volume of sample (L)}}$$

where:

B = Weight of dish after evaporation table (mg);

A = Tare weight of dish (mg);

C = Weight of dish after muffler oven (mg).

The suspended solids test was performed on the mixed liquor and effluent solids. Standard Methods [79] test 209C was used for total suspended solids and 209D for volatile suspended solids. The formula for total suspended and volatile suspended solids is based on the difference in weights and is as follows:

$$\textit{Total Suspended Solids (mg/l)} = \frac{B - A}{\textit{volume of sample (mg/l)}}$$

$$\textit{Volatile Suspended Solids} = \frac{B - C}{\textit{volume of sample (mg/l)}}$$

where:

A = Tare weight of filter and planchet (mg);

B = Weight of filter and planchet after 103^0 oven;

C = Weight of filter and planchet after muffler.

An alternate method was used for the mixed liquor total suspended and mixed liquor volatile suspended solids. A 100 ml sample of mixed liquor was centrifuged at 1000g. The sample was than run through a Whatman paper filter. A filtrate volume of 50 ml was collected and a total solids and total volatile solids test were done on it as stated above. The total solids on the filtrate were considered the total dissolved solids of the mixed liquor. The total volatile solids were considered the volatile dissolved solids. The total suspended solids and the volatile suspended solids were then determined using the following formula:

$$TSS = TS - TDS$$

$$VSS = TVS - VDS$$

where:

TSS = Total Suspended Solids (mg/l);

TS = Total Solids (mg/l);

TDS = Total Dissolved Solids (mg/l);

VSS = Volatile Suspended Solids (mg/l);

TVS = Total Volatile Solids (mg/l);

VDS = Volatile Dissolved Solids (mg/l).

This method was used because the solids were too thick to run by the conventional suspended solids test.

7. Ammonia

Ammonia analyses were performed occasionally to determine if there were any inhibitory effects on the reactor due to ammonia. The ammonia determinations were made by following Standard Method 417E, the ammonia-selective electrode method [79]. The method was modified by using 50 ml of sample instead of 100 ml, and the samples were not diluted.

The ammonia-selective electrode used was an Orion Research Model 95-12. The electrode was used with an Altex Instruments Model 4500 digital pH meter. A semi-log calibration plot of millivolts vs log(concentrations) was developed using standards at 1000, 100, and 10 mg/l ammonia. The NH_3-N concentration was found by placing the reading of the sample on the calibration curve and reading the ammonia nitrogen concentration.

V. RESULTS AND DISCUSSION

The following section is a summary of the results of the performance of the ASBR using swine waste at various hydraulic retention times and organic loads.

A. Influence of Organic Loading Rate

The organic loading rate on the reactors was defined by volatile solids (VS) load and total chemical oxidation demand (TCOD) load. The TCOD is often abbreviated COD instead of TCOD. Tables 15 and 16 list the various VS and COD loadings applied to the reactors. The performance of the reactor can be measured by the removal of VS and COD. The removal percentage is calculated by the following formula:

$$Removal(\%) = \frac{S_i - S_o \pm \Delta M}{S_i} \times 100$$

where:

S_i = Influent organics (COD or VS)/day;

S_o = Effluent organics (COD or VS)/day;

ΔM = Average daily change in mixed liquor COD or VS in the reactors.

The VS destruction at various HRTs are shown in Tables 17 through 19 and the COD removals are shown in Tables 20 through 22.

Table 15. Volatile solids loads applied to reactors

	HRT		
	6 days	9 days	12 days
Volatile Solids Load, G/L/D	1.40	0.93	0.93
	1.84	1.39	0.92
	2.39	1.91	1.45
	3.31	2.90	2.49
	4.19	3.84	2.97
	5.37		

Table 16. TCOD loads applied to reactors

	HRT		
	6 days	9 days	12 days
TCOD Load, G/L/D	1.57	1.04	1.04
	2.06	1.56	1.03
	3.21	2.56	1.94
	3.74	3.28	2.81
	4.03	3.69	3.27
	5.32		

Table 17. Volatile solids destruction at a 6 day HRT

VS Load G/L/Day	VS Influent G/Day	VS Effluent G/Day	M (VS) G/Day	VS Destroyed %
1.40	16.8	9.9	+.000	41.0
1.84	22.1	11.5	-.015	47.7
2.39	28.7	15.8	+.033	44.8
3.31	39.7	20.6	+.180	47.7
4.19	50.3	26.1	-.090	48.2
5.37	64.4	36.6	+.010	43.2

Table 18. Volatile solids destruction at a 9 day HRT

VS Load G/L/Day	VS Influent G/Day	VS Effluent G/Day	M (VS) G/Day	VS Destroyed %
0.93	11.2	5.6	+.000	49.9
1.39	16.7	4.6	-.014	71.9
1.91	22.9	11.5	-.038	50.1
2.90	34.8	19.2	+.230	44.2
3.84	46.1	20.8	-.060	55.0

Table 19. Volatile solids destruction at a 12 day HRT

VS Load G/L/Day	VS Influent G/Day	VS Effluent G/Day	M (VS) G/Day	VS Destroyed %
0.93	11.2	5.4	+.000	51.8
0.92	11.0	3.5	+.035	67.2
1.45	17.4	8.1	-.130	54.7
2.49	29.9	16.7	+.240	43.3
2.97	35.6	18.2	+.160	48.6

Table 20. Measured COD removals at a 6 day HRT

COD Load G/L/Day	COD Influent G/Day	COD Effluent G/Day	M (COD) G/Day	COD Destruction %
1.57	18.8	16.9	+.000	10.0
2.06	24.7	17.1	-.053	30.5
3.21	38.5	22.5	+.034	41.5
3.74	44.9	35.9	+.090	20.0
4.03	48.4	41.0	+.060	16.0
5.32	63.8	60.0	+.270	5.8

Table 21. Measured COD removals at a 9 day HRT

COD Load G/L/Day	COD Influent G/Day	COD Effluent G/Day	M (COD) G/Day	COD Destruction %
1.04	12.5	8.8	+.000	29.6
1.56	18.7	5.5	+.021	70.0
2.56	30.7	15.4	+.044	49.7
3.28	39.4	29.6	+.180	24.4
3.69	44.3	30.0	-.000	32.0

Table 22. Measured COD at removals at a 12 day HRT

COD Load G/L/Day	COD Influent G/Day	COD Effluent G/Day	M (COD) G/Day	COD Destroyed %
1.04	11.2	5.4	+.000	51.8
1.03	11.0	3.5	+.035	67.2
1.94	17.4	8.1	-.130	54.7
2.81	29.9	16.7	+.240	43.3
3.27	35.6	18.2	+.160	48.6

Figure 12 shows the VS destruction as a function of VS load. The plot also shows Pidaparti's [20] data. His data are relevant because the data were taken in the same reactors at an earlier time and higher temperature. This plot shows

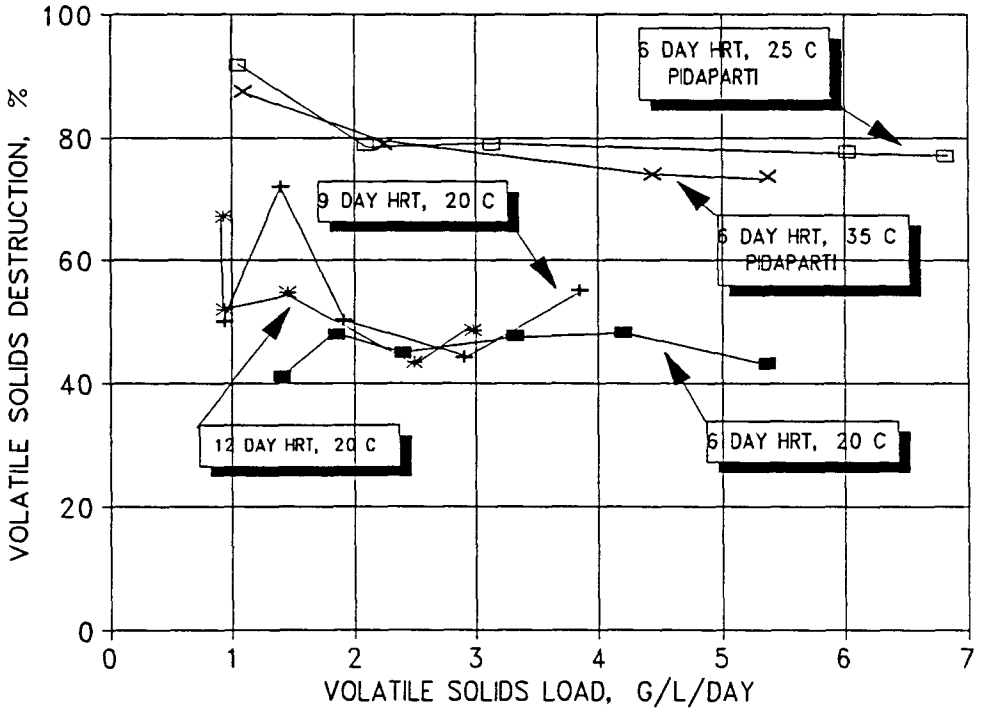

Figure 12. Volatile solids destruction at various HRTs

that as the VS load increased the VS destruction decreased. Initially, all three reactors had an increase in VS destruction with the first increase in VS load. This is due to the maturing of the reactor. The maturing process takes a long time, especially at 20° C. The first time the load was raised the reactors had not finished the microbial selection process. Once this maturing period was finished, the VS destruction did decrease with an increase in load. The 9 and 12 day HRT reactors seemed to display an increase in VS destruction at the final and highest load. This may have been due to error in analysis of VS. Another possible explanation is a change in the consistency of the feed. The swine waste tends to change depending on when it is collected and how long it is stored. The plot shows that at low loadings the 9 and 12 day HRT reactors tend to perform better than the 6 day reactor. At loadings of 2 to 3.5 g/l/day the performances at all HRTs tended to be about the same.

The longer the HRT the lower the load was at failure. A possible explanation is that something toxic to the consortium of microorganisms inhibited the microorganisms and caused the reactor to fail. In the study the load was set at a specific COD and the HRT was set at a specific time. This meant that every day the same amount of liquid was batch fed to the reactors. The amount depended on the reactors HRT. The daily feed was made up of swine wastes and tap water. Since the

reactor with a shorter HRT was fed more tap water than the reactor with the longer HRT, any toxic compounds would be more diluted. This would cause the toxicity to inhibit the reactor with a longer HRT at a lower load than the reactor with a shorter HRT. Therefore, the longer the HRT the faster the failure due to toxicity. Another explanation for the failure is that the "non volatile" solids choked out the microorganisms. In this case the HRT would function in the same way as for a toxic substance causing the solids to build up in the reactors with the longer HRTs.

Failure was characterized by a decrease in gas production and VS destruction. Also occurring at failure was a high concentration of volatile acids. When the load was increased on the reactors and the removals started to decrease, the supernatant of the reactor started to contain more solids. At low loads there was a clear and definite interface at the end of the settling phase. As the loads increased, this interface became more turbid. Most of this was due to gas bubbles adhering to particles causing them to float into the supernatant.

Pidaparti's [20] data showed almost identical removals at 25°C as at 35° C. Most studies show a substantial decrease in removals as the temperature is dropped. Even though the reactors had a lower removal at 20°, the VS destruction was around 50%.

There are two ways to determine the COD removal. One is to measure the COD directly with the COD test, and the other is to calculate the removals based on the methane production. One gram of COD removed, produces 0.35 liters of methane at STP. Measuring the COD is not a very easy task when working with swine wastes. The sample size for the test is 5 ml which means that one large solids particle can make a big difference in the COD. Also, due to the high COD of the waste, the sample had to be diluted by 50, 100, or 250 times. Every dilution introduces more error. To help minimize this error, samples were run in duplicate once a day for three days and averaged for a final value. Even with this adjustment for accuracy, the measured results were not very consistent. Therefore, both measured COD results and removals based on methane production are included. Measuring methane is much more precise than measuring COD, so these results tend to be more reliable.

Figure 13 shows the COD removals at the various HRTs as measured in the lab. It also shows Pidaparti's [20] data at 35° C and 25° C. As in the VS destruction data, the removals at 20° C are lower than at 35° C and 25° C. The graph clearly shows that the removals drop as the load is increased. The decline is sharp at all three HRTs. Figure 14 shows the COD removals based on methane production. As expected the removals declined as the load was increased. The decline in

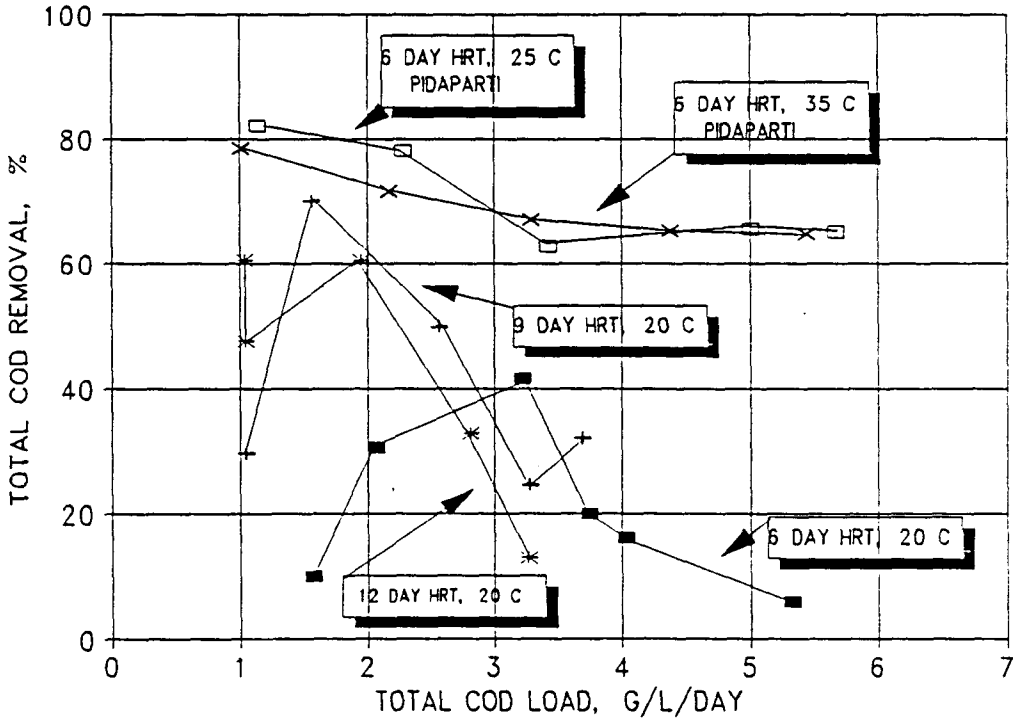

Figure 13. Measure TCOD removals at various HRTs

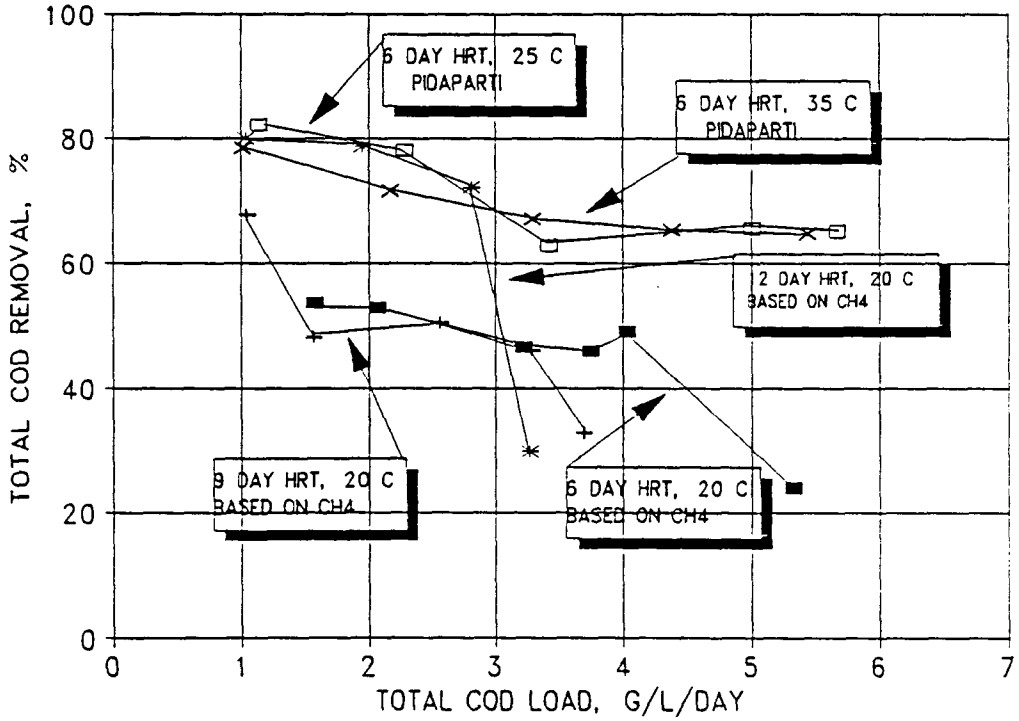

Figure 14. TCOD removals based on methane at various HRTs

the data from the methane equivalent COD value is less sharp than that shown from the measured COD values. The data in Figure 14 is more reliable than the data in Figure 13 because gas measurement is more accurate than the lab method. The 12 day HRT reactor was getting removals as high as Pidapartis at the low loads. However, when the reactor started failing the removal declined by about 40%. This large decline in COD before system failure occurred in all three reactors. It is interesting that although the removals at 20° C are less than at 35° C and 25° C the removals are still about 50%.

B. Total Gas and Methane Production

Gas production is the most important measurement in analyzing the performance of a reactor. Tables 23 through 25 show the quantity of biogas and methane produced for the reactors at the three different HRTs.

Figures 15 and 16 show total gas productions for the reactors at the three different HRTs. Pidaparti's data is also plotted on the graphs. Figure 15 shows the total gas based on COD, and Figure 16 shows the total gas based on VS. Figures 15 and 16 show total gas produced in liters per liter of reactor volume per day. The plots show that as the load is increased the amount and rate of gas produced increases. When the reactors failed it is immediately noticeable by a sudden decrease in the total gas produced and the rate of gas

Table 23. Biogas and methane production at a 6 day HRT

VS Load g/l/d	COD Load g/l/d	Total[a] Gas l/d	Total[b] Gas l/l/d	Methane l/d	Methane l/l/d	% Methane
1.40	1.57	5.23	0.44	3.53	0.29	67.5
1.84	2.06	6.42	0.54	4.59	0.38	71.5
2.39	3.21	9.86	0.82	6.27	0.52	63.6
3.31	3.74	11.63	0.97	7.20	0.60	61.9
4.19	4.03	13.60	1.13	8.29	0.69	61.0
5.37	5.32	10.53	0.88	5.37	0.45	51.0

[a]l/d = liters of gas produced by the reactor each day.

[b]l/l/d = liters of gas produced per liter of reactor volume per day.

Table 24. Biogas and methane production at a 9 day HRT

VS Load g/l/d	COD Load g/l/d	Total[a] Gas l/d	Total[b] Gas l/l/d	Methane l/d	Methane l/l/d	% Methane
0.93	1.04	4.67	0.39	2.97	0.25	63.6
1.39	1.56	4.82	0.40	3.15	0.26	65.4
1.91	2.56	8.66	0.72	5.43	0.45	62.7
2.90	3.28	10.78	0.90	6.34	0.53	58.8
3.84	3.69	9.00	0.75	5.08	0.42	56.4

[a]l/d = liters of gas produced by the reactor each day.

[b]l/l/d = liters of gas produced per liter of reactor volume per day.

Table 25. Biogas and methane production at a 12 day HRT

VS Load g/l/d	COD Load g/l/d	Total[a] Gas l/d	Total[b] Gas l/l/d	Methane l/d[1]	Methane l/l/d[2]	% Methane
0.93	1.04	4.67	0.44	2.97	0.29	63.6
1.39	1.56	4.82	3.15	3.15	0.38	65.4
1.91	2.56	8.66	0.82	5.43	0.52	62.7
2.90	3.28	10.78	1.13	6.34	0.69	58.8
3.84	3.69	9.00	0.88	5.08	0.45	56.4

[a]l/d = liters of gas produced by the reactor each day.

[b]l/l/d = liters of gas produced per liter of reactor volume per day.

production. From the graphs it can be seen that the gas production at a 12 day HRT followed Pidaparti's [20] data. The point at which the gas production quit following his production is the point at which the reactor started to fail. The 6 and 9 day HRTs gas production followed parallel paths, with the 9 day HRT performing slightly better than the 6 day HRT reactor, but failing first.

Figures 17 and 18 show the methane produced per liter of reactor volume per day. They are also plotted against the COD load and the VS load. All the figures indicate that there is a drop in gas production from 35° C to 20° C. The only exception is the gas production of the 12 day HRT reactor at TCOD loads less than 2.5 g/l/d.

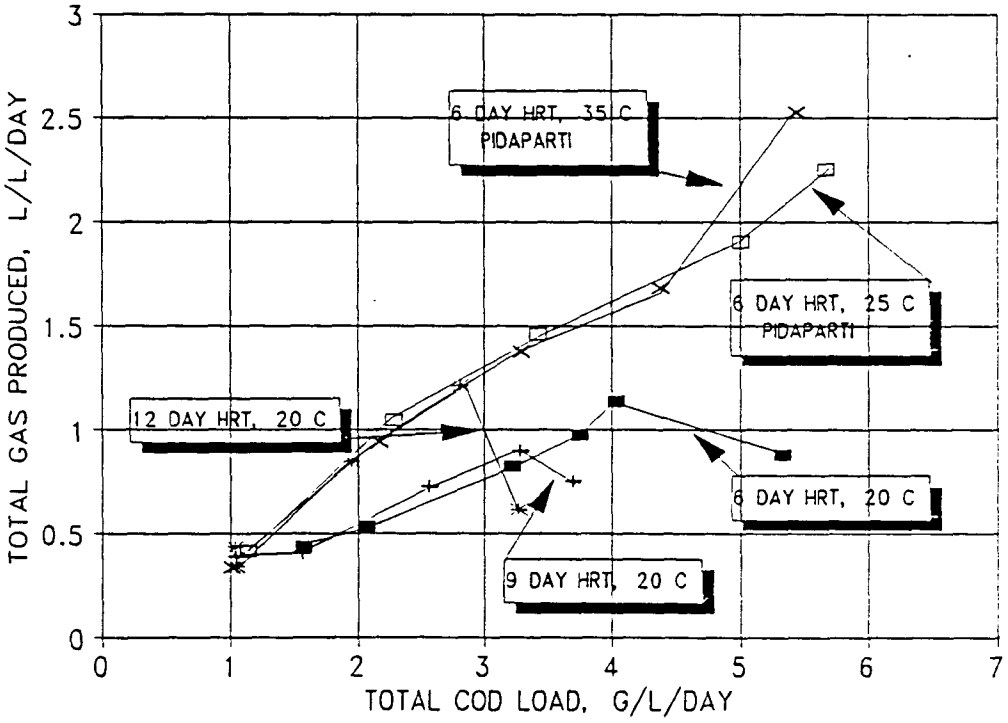

Figure 15. Total gas production in (L/L/D) at various COD loads and HRTs

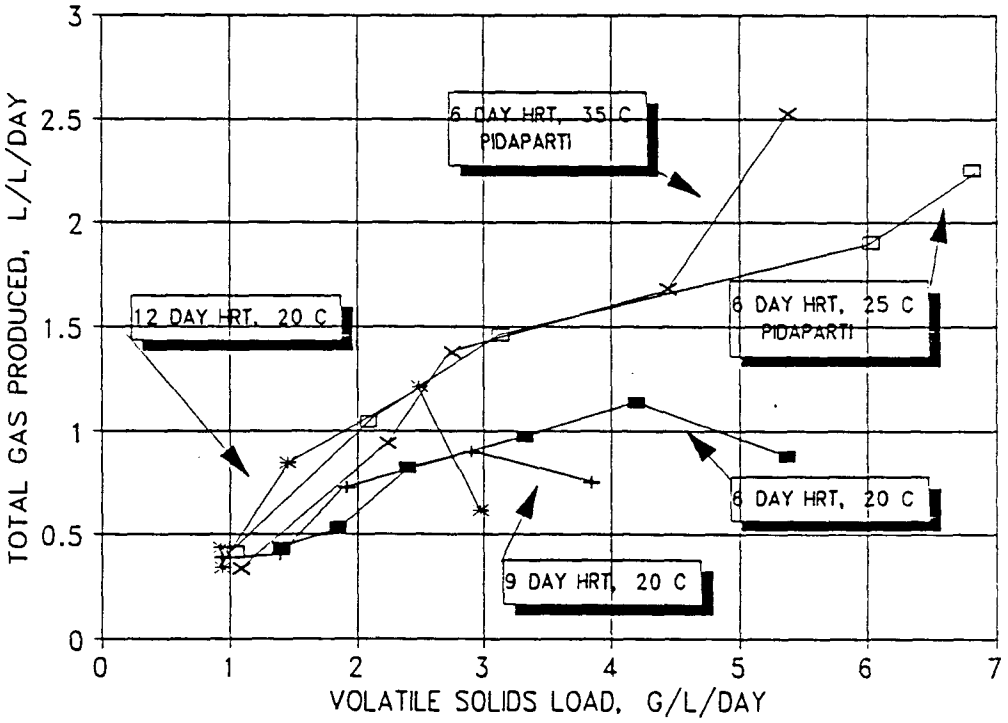

Figure 16. Total gas production in (L/L/Day) at various VS loads and HRTs

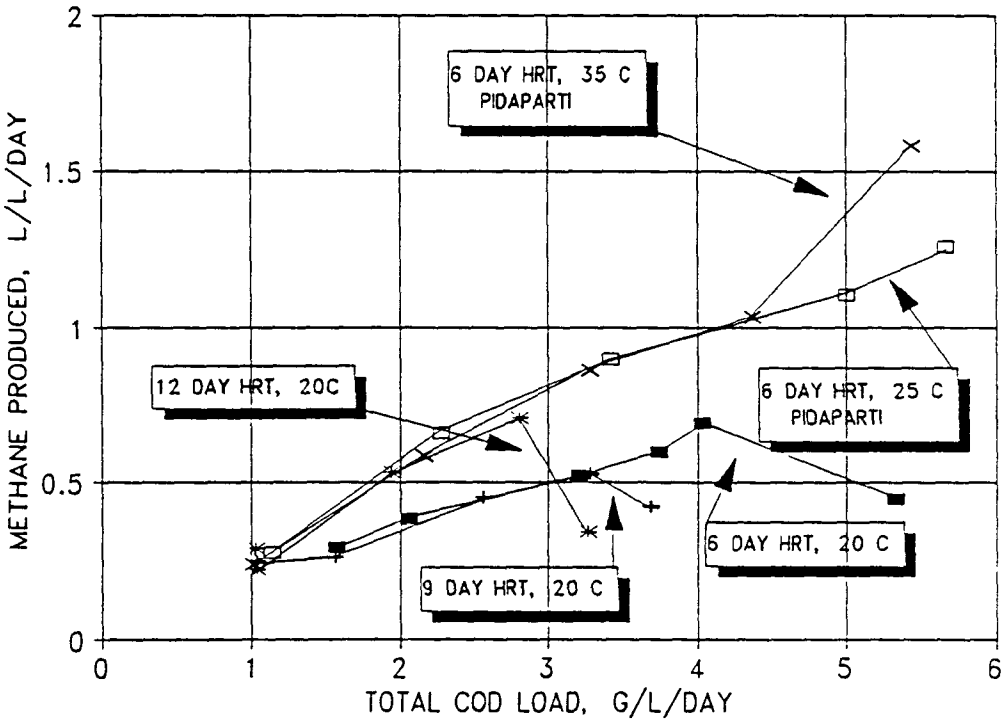

Figure 17. Methane production in (L/L/Day) at various COD loads and HRTs

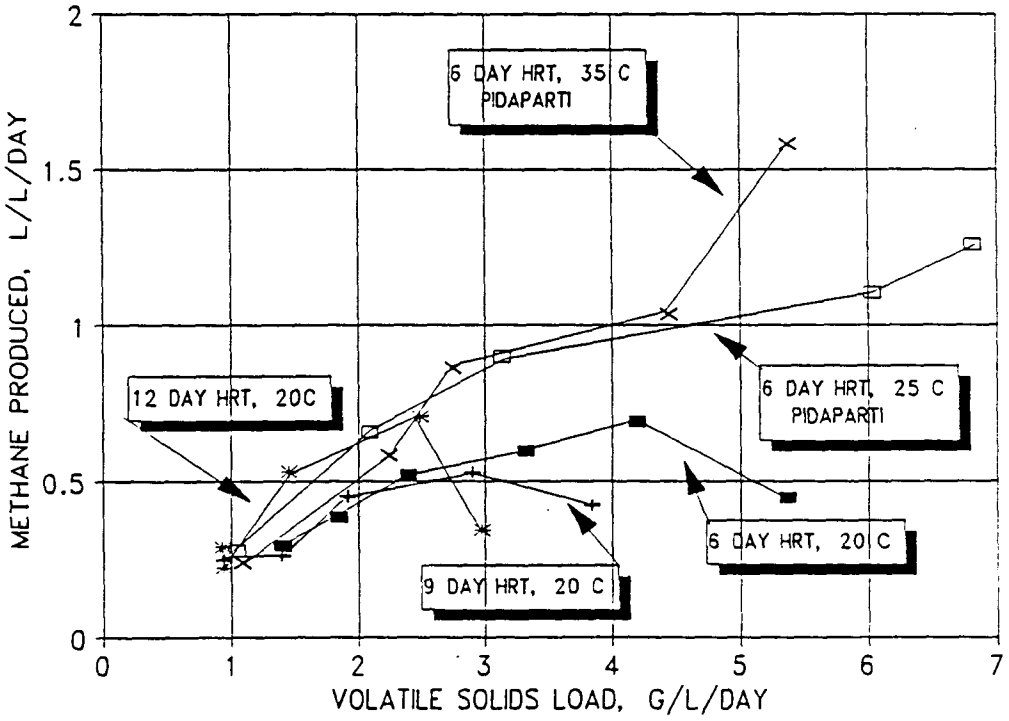

Figure 18. Methane production in (L/L/Day) at various VS loads and HRTs

C. Solids Retention Time

One of the theories about the ASBR is that it can compensate for temperature. If this is so, the solids retention time (SRT) inside the reactor should increase as the temperature decreases. As the temperature decreases the microorganisms tend to grow slower, but they also tend to die slower (lower endogenous decay). As long as the decrease in endogenous decay rate is the same or greater than the decreases in the growth rate, the system can compensate for lower temperatures. When the endogenous decay rate exceeds the rate of growth, the methanogens will be "washed out" of the system. The ability to hold solids is measured by the SRT, as described in the procedures section.

The SRT of the system is the average time a suspended solid spends in the system. Pfeffer [12] demonstrated previously that the minimum SRT for methanogenic regeneration at 35° is ten days and it increases 50% for every 10° C decrease in temperature. This means that at 20° C the minimum SRT needed to maintain a viable population is approximately 19 days. This assumes that the SRT is a measure of biomass. Actually, the SRT measures solids which is assumed to be representative of biomass. This assumption is not always correct making SRT a questionable means for measuring biomass. Tables 26 through 28 show the SRT of the reactors at the three

Table 26. Solids retention time at a 6 day HRT

VS Load g/l/d	COD Load g/l/d	Solids Retention Time (days)
1.40	1.57	16.24
1.84	2.06	20.07
2.39	3.21	12.77
3.31	3.74	18.04
4.19	4.03	12.85
5.37	5.32	8.52

Table 27. Solids retention time at a 9 day HRT

VS Load g/l/d	COD Load g/l/d	Solids Retention Time (days)
0.93	1.04	29.88
1.39	1.56	60.55
1.91	2.56	17.28
2.90	3.28	19.43
3.84	3.69	15.43

Table 28. Solids retention time at a 12 day HRT

VS Load g/l/d	COD Load g/l/d	Solids Retention Time (days)
0.93	1.04	33.20
0.92	1.03	63.55
1.45	1.94	17.80
2.49	2.81	19.16
2.97	3.27	23.82

HRTs.

Figures 19 and 20 are graphical representations of the SRTs plotted against VS and TCOD. Also shown on the graphs are Pidaparti's [20] data and the expected minimum SRT of nineteen days. It is plainly clear that the minimum SRT of nineteen days was not met after the load was increased. An explanation is that the SRT is not a measure of biomass. This has been discussed previously. This explanation is supported by the observation of the settling of the reactor. The settling was segregated due to larger particles or "granular" forming and settling first followed by the less settleable flocculent particles. This observation calls into question the assumption that solids is a measure of biomass. The problem of SRT measuring biomass has been addressed by many authors, most notably Dague [4].

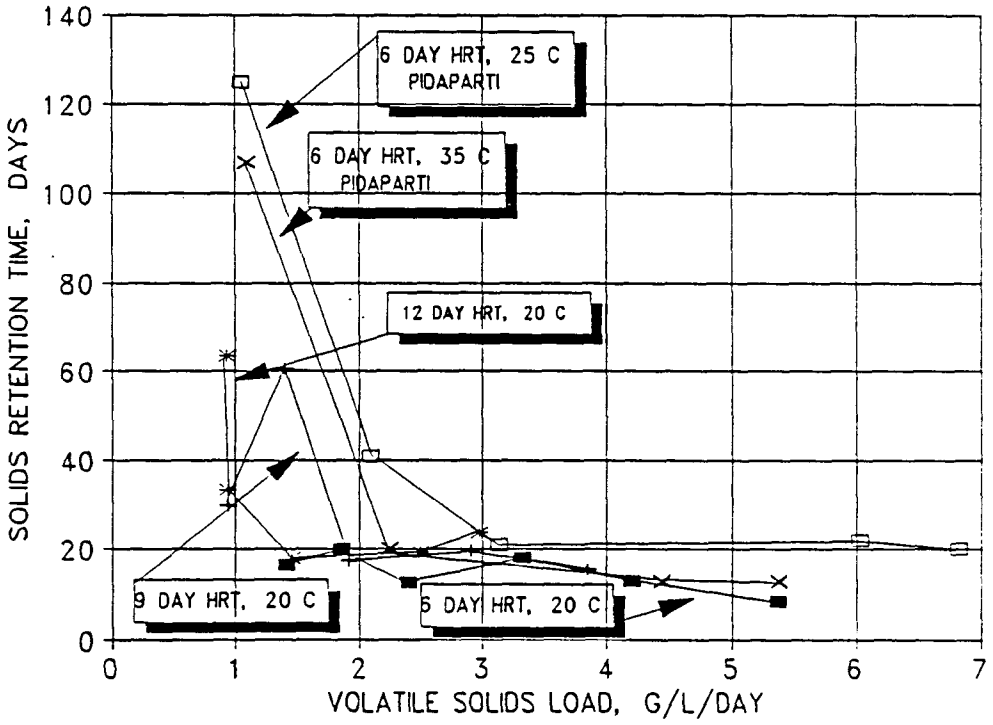

Figure 19. SRT at various VS loads and HRTs

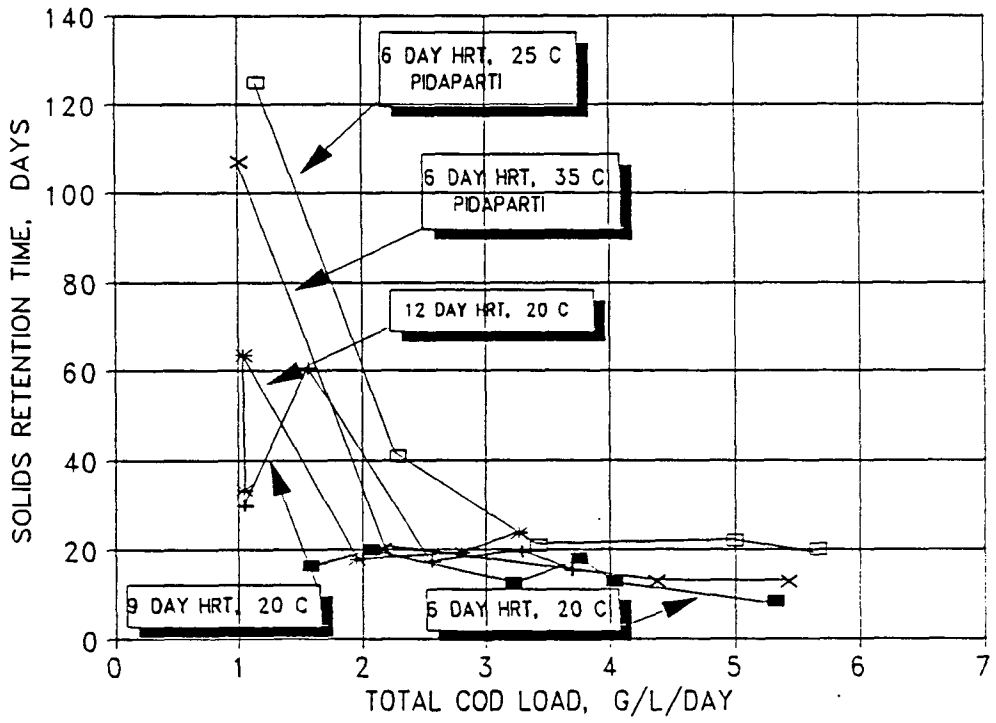

Figure 20. SRT at various COD loads and HRTs

It was expected that at the higher HRT the SRT would be longer because of the reduced hydraulic pressure. This turned out not to be the case. The 12 day HRT reactor did not have an appreciable higher SRT over the 6 day rector. This occurred because the effluent from the 6 day HRT reactor had less solids than the effluent from the 12 day HRT reactor, so even though twice as much effluent volume was being wasted from the 6 day HRT reactor, the SRT was about the same.

D. Mixed Liquor Suspended Solids

The mixed liquor suspended solids (MLSS) in the reactor is a good indication of how the system is holding solids. The MLSS can also be an indication of the amount of biomass that is in the reactor (see previous section). The MLSS was used to determine whether VS are being stored in the reactor instead of being destroyed. This factor is accounted for in percent removals by the variable ΔM. Tables 29 through 31 show how the mixed liquor suspended solids level changed in the reactor at various VS and COD loads.

Figures 21 and 22 show the mixed liquor suspended solids against VS load and COD load. The lines paralleled each other among the three HRTs. The highest MLSS was in the 12 day reactor followed by the 9 and the 6 day reactor. The MLSS gradually increased as the loads increased. This seems logical because no MLSS was wasted from the reactors at any

Table 29. Mixed liquor suspended solids at a 6 day HRT

VS Load g/l/d	COD Load g/l/d	Mixed Liquor Suspended Solids mg/l
1.40	1.57	17445
1.84	2.06	23487
2.39	3.21	20611
3.31	3.74	38321
4.19	4.03	36687
5.37	5.32	27930

Table 30. Mixed liquor suspended solids at a 9 day HRT

VS Load g/l/d	COD Load g/l/d	Mixed Liquor Suspended Solids mg/l
0.93	1.04	18107
1.39	1.56	26215
1.91	2.56	21236
2.90	3.28	40356
3.84	3.69	35100

Table 31. Mixed liquor suspended solids at a 12 day HRT

VS Load g/l/d	COD Load g/l/d	Mixed Liquor Suspended Solids mg/l
0.93	1.04	16350
0.92	1.03	25863
1.45	1.94	16051
2.49	2.81	34266
2.97	3.27	45835

time. At the higher loads the reactor performance started to decline which caused the effluent to increase in solids. This increase in effluent solids tended to reduce the MLSS due to washout of solids. Generally, in a full scale plant a solids management plan would be implemented to keep the MLSS at a healthy level. A high MLSS level can be unhealthy because as solids enter the system a majority of the VS get destroyed in the anaerobic process. This means that the increase in MLSS is usually attributable to an increase in the non-volatile solids. The non-volatile solids generally interfere with the performance of the reactor by "crowding out" the active microorganisms. This "crowding out" is the most probable reason for reactor failure.

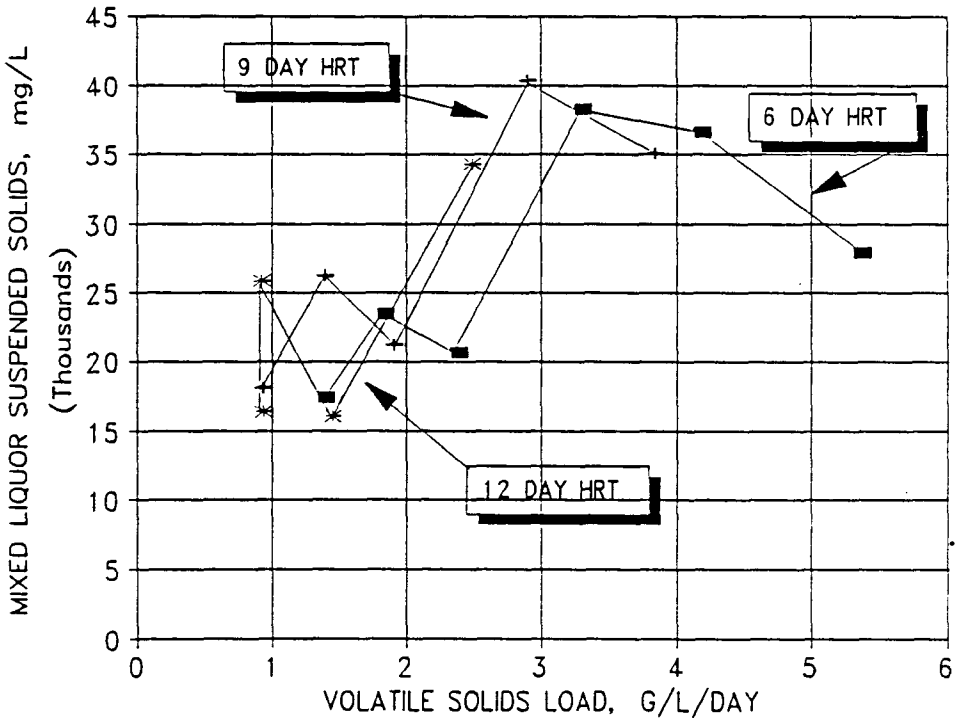

Figure 21. Mixed liquor suspended solids at various VS loads and HRTs

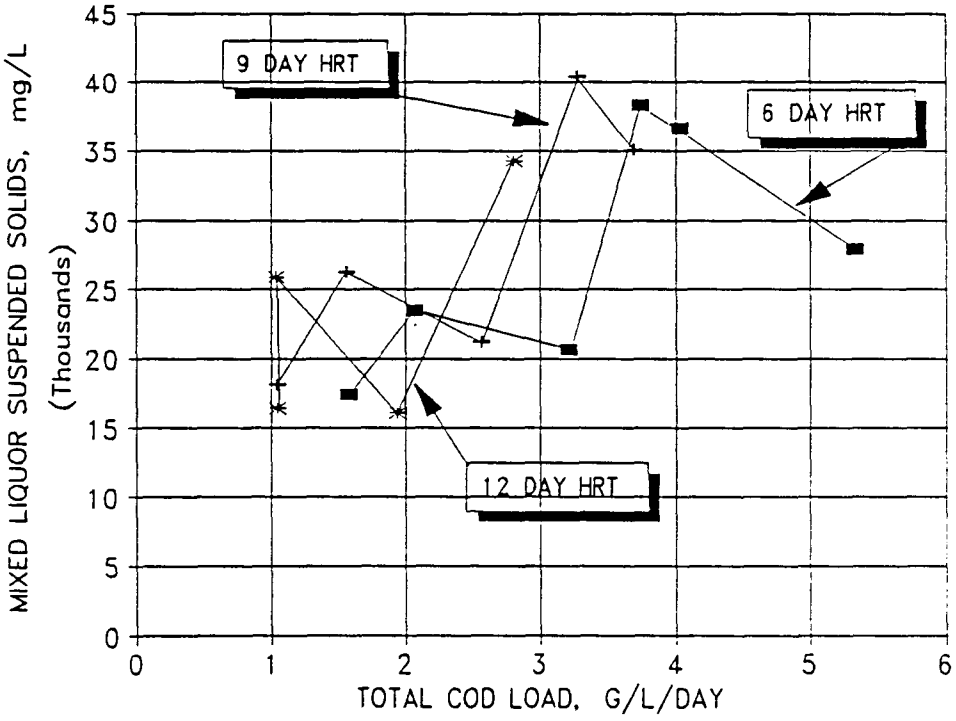

Figure 22. Mixed liquor suspended solids at various COD loads and HRTs

E. Ammonia

There was a concern at the start of this study that the high levels of ammonia in the waste could be a contributing factor to system failure. For this reason, the ammonia levels were watched as the load was increased. According to McCarty, ammonia (as NH_3 - N) becomes inhibitory at 1500 mg/l and toxic at 3000 mg/l [6]. Other research has found that anaerobic systems can be acclimated to concentrations of ammonia as high as 4000 mg/l, as long as the pH remains slightly above or below neutral. Tables 32 through 35 list ammonia concentrations over time, VS load, and COD load.

Figures 23 through 25 show a graphical representation of the ammonia against VS and COD loads as well as time. Figures 23 and 24 show that as the load increased the ammonia increased as well. The increase was linear with respect to the load. There is a stray point on the six day HRT which shows the ammonia leveling off. This point could be an erroneous data point. These two figures show that the lower HRT reactors have a diluting effect on toxic substances. The ammonia level never got above 800 mg/l as NH_3-N in any reactor. This was probably not the culprit that caused the reactors to fail. The pH of the reactors during the experiment ranged from 6.6 at low loads to 7.0 at higher loads.

Figure 25 shows the ammonia concentration over time. The graph shows that the ammonia levels in the reactor tracked

Table 32. Ammonia-nitrogen at a 6 day HRT

VS Load g/l/d	COD Load g/l/d	Ammonia Nitrogen mg/l as NH_3
1.40	1.57	83.8
1.84	2.06	273.5
2.39	3.21	412.5
3.31	3.74	463.5
4.19	4.03	463.8
5.37	5.32	457.3

Table 33. Ammonia-nitrogen at a 9 day HRT

VS Load g/l/d	COD Load g/l/d	Ammonia Nitrogen mg/l as NH_3
0.93	1.04	144.7
1.91	2.56	344.9
2.90	3.28	495.3
3.84	3.69	713.7

Table 34. Ammonia-nitrogen at a 12 day HRT

VS Load g/l/d	COD Load g/l/d	Ammonia Nitrogen mg/l as NH_3
0.93	1.04	128.0
1.45	1.94	307.1
2.49	2.81	505.5
2.97	3.27	751.6

Table 35. Ammonia-nitrogen as N, (mg/l)
day 0 is 06/11/91

Day	6 Day HRT	9 Day HRT	12 Day HRT
0	528.0	515.3	530.2
15	437.3	488.0	502.1
30	332.0	346.0	313.4
96	83.8	144.7	128.0
232	273.5	344.9	307.1
289	412.5	495.3	505.5
343	463.5	713.7	751.6
356	463.8	766.4	766.4
364	457.3	-----	-----

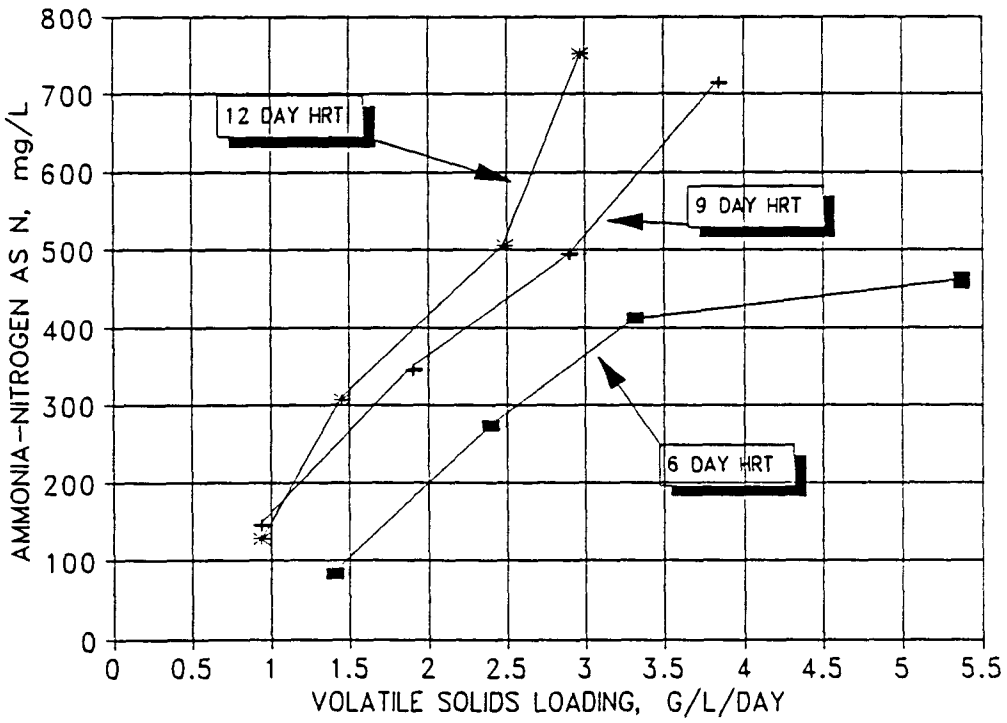

Figure 23. Ammonia at various VS loads and HRTs

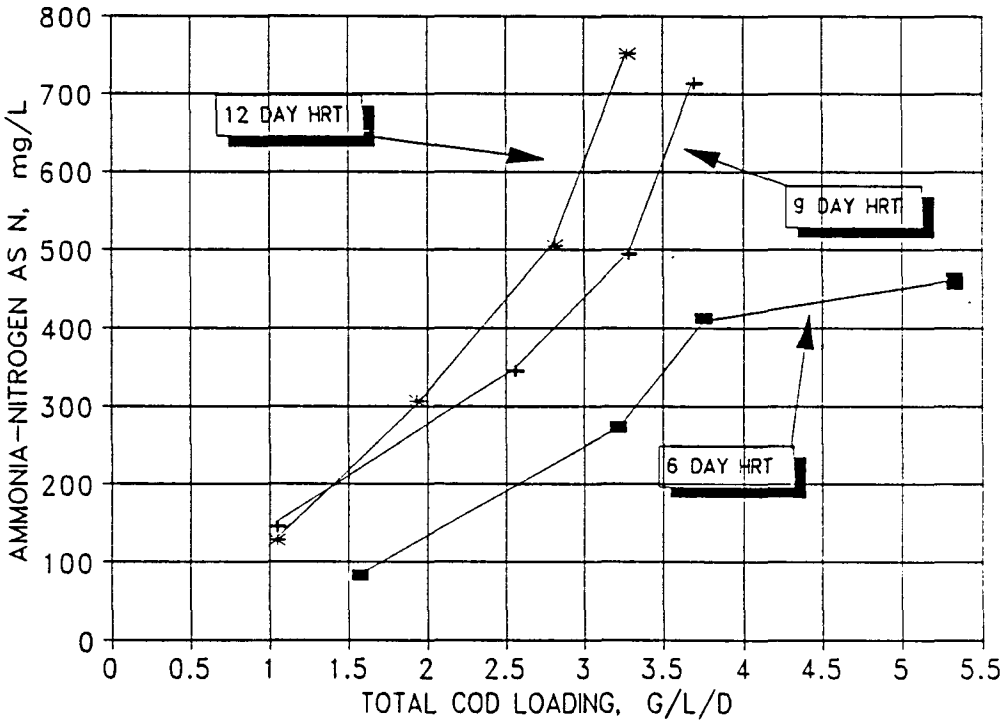

Figure 24. Ammonia at various COD loads and HRTs

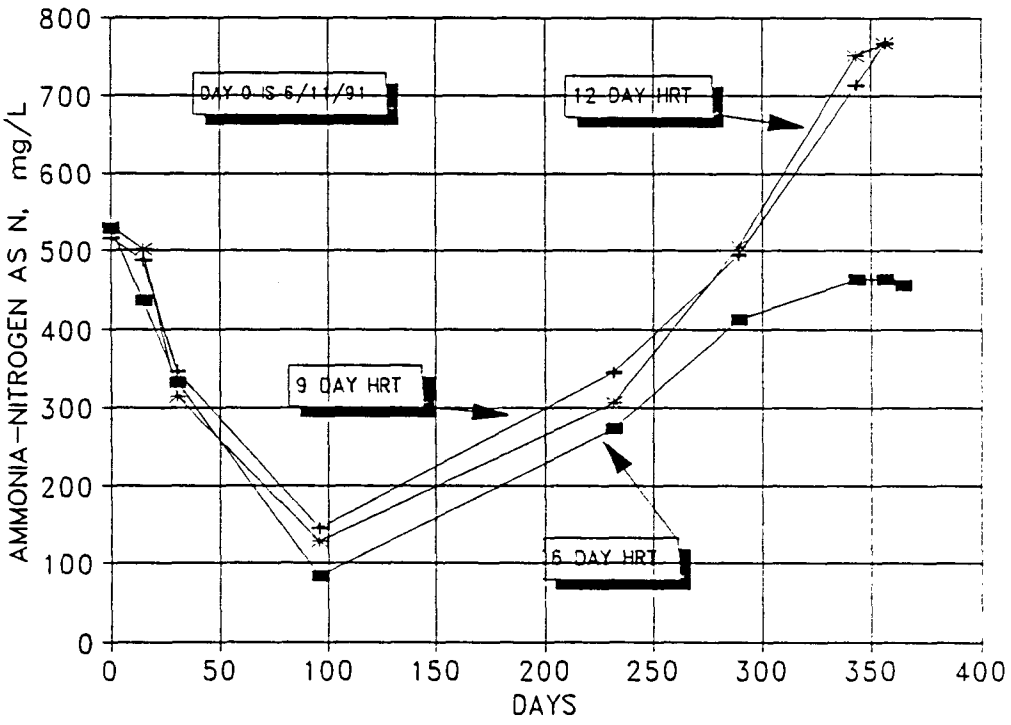

Figure 25. Ammonia at various HRTs over time

each other. The 6 day HRT had ammonia concentrations lower than the other two reactors. The ammonia concentrations in all three reactors started high, went to a minimum, and than increased over time. When the reactors were turned down to 20° C, the loading on the reactors was high. This caused the reactors to start to fail before they acclimated to the new, lower temperature. To prevent failure, the loads on the reactors were lowered. As the loads were lowered the concentration of ammonia decreased.

F. Alkalinity

The alkalinity in a reactor is a measure of the buffering capacity of the system against a change in pH. If acids start building up in the reactor, the acids react with the alkalinity and the pH is unaffected. A low alkalinity means that the system will be upset easily. There are different forms of alkalinity. The main form of alkalinity is bicarbonate. Bicarbonate alkalinity was added to the reactor at low loads to increase the alkalinity. The alkalinity was added in the form of sodium bicarbonate. As the load increased the alkalinity increased. This is due to the protein in the reactor being converted to ammonium bicarbonate. Tables 36 through 38 lists alkalinity over various VS and COD loads.

Figures 26 and 27 show the alkalinity as it varied over

Table 36. Alkalinity at a 6 day HRT

VS Load g/l/d	COD Load g/l/d	Alkalinity mg/l as $CaCO_3$
1.40	1.57	1954
1.84	2.06	1900
3.31	3.74	3175
4.19	4.03	3417
5.37	5.32	2850

Table 37. Alkalinity at a 9 day HRT

VS Load g/l/d	COD Load g/l/d	Alkalinity mg/l as $CaCO_3$
0.93	1.04	2292
1.39	1.56	2300
2.90	3.28	3583
3.84	3.69	3517

Table 38. Alkalinity at a 12 day HRT

VS Load g/l/d	COD Load g/l/d	Alkalinity mg/l as $CaCO_3$
0.93	1.04	2383
0.92	1.03	1800
2.49	2.81	4000
2.97	3.27	3117

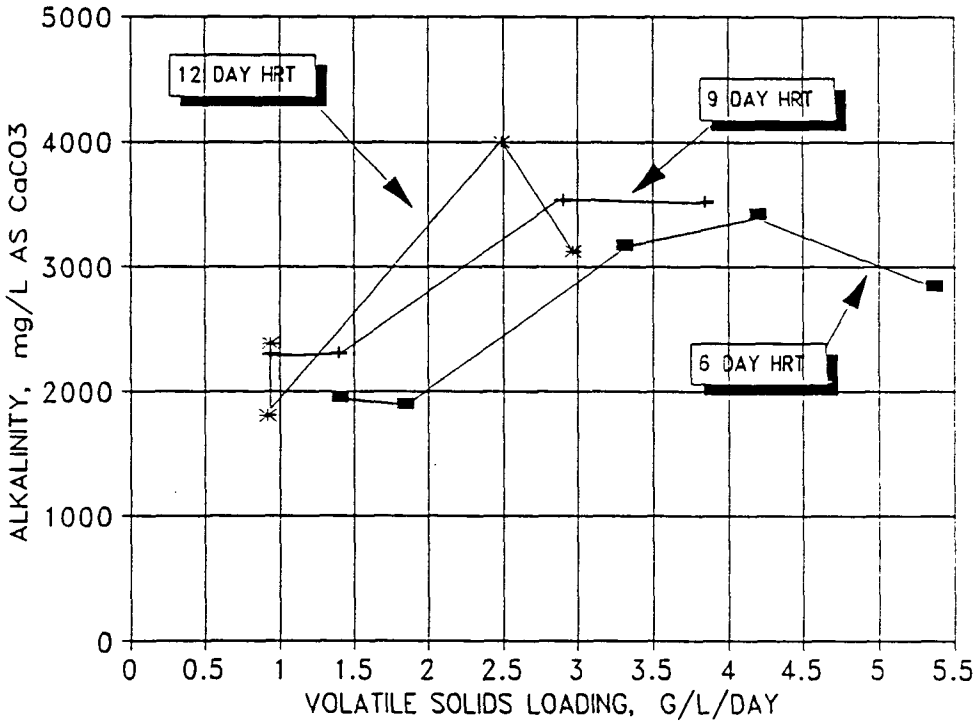

Figure 26. Alkalinity at various VS loads and HRTs

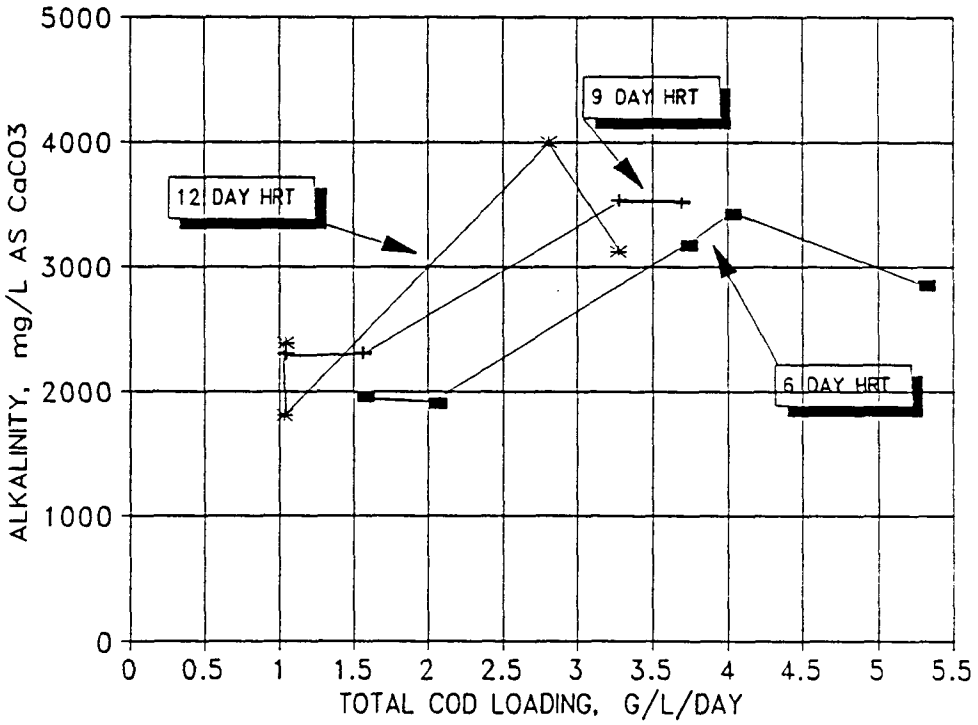

Figure 27. Alkalinity at various COD loads and HRTs

loadings and HRTs. As expected, the alkalinity increased as the load was increased. The alkalinity appeared to be independent of HRT.

Figure 28 shows how the alkalinity changed over time. It appears that the alkalinity is independent of time and dependent on organic loading. The data for Figure 28 has not been put into a table because the data is too extensive.

G. Stability of Effluent

Pidaparti [20] noticed that when the effluent was withdrawn from the reactors the effluent did not exhibit much of an odor. Additional experiments were run to determine how much more the effluent could be treated by transferring some of the supernatant to an additional beaker and measuring its volatile acids.

In this study, the effluent for the experiment for was withdrawn at reactor failure and had a very distinct hydrogen sulfide and intermediate acid odor. In order to quantify this, some tests were run where the effluent was taken from the reactor and brought up to 35° C to determine how much additional gas would be produced. The effluent was taken from the reactor with a HRT of 6 days. The effluent for these stability experiments was taken at the final load before reactor failure. The effluent was in a 35° incubator for forty days. The amount of additional gas produced was 3.96 ml of

methane per ml of effluent. This shows that if a full scale plant were designed at this load there would need to be some kind of additional treatment. It must be mentioned, however, that the reactor was operating in a stressed mode at reactor failure, and that under normal conditions the effluent would be much more stable.

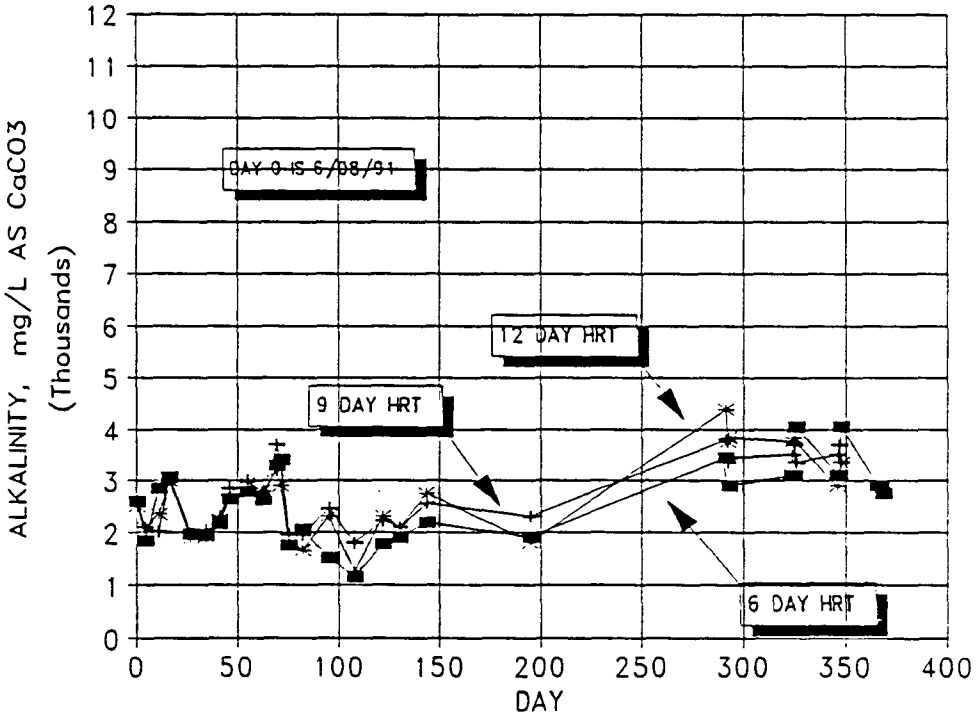

Figure 28. Alkalinity at various HRTs over time

VI. CONCLUSIONS

The following conclusions can be drawn from this study:

1. The anaerobic sequencing batch reactor is capable of sustaining volatile solids destructions over a variety of loads in the range of 40 to 60% at a temperature of 20° C. Other studies have not gotten CSTR reactors to work at this temperature.
2. The temperature compensation abilities of the ASBR are less going from 25° C to 20° C than going from 35° C to 25° C using swine waste as a substrate.
3. Reactor failure was probably due to "crowding out" of the methanogens by the build up of solids in the reactor.
4. Ammonia toxicity does not appear to be a problem at the loads tested.
5. At low loads, the effluent produces negligible amounts of odor. However, at higher loads, the effluent does give off more odor and additional treatment may be necessary.
6. Solid retention time is not always a good indication of the amount of biomass in a reactor.

VII. RECOMMENDATIONS FOR FUTURE RESEARCH

Based on the results and observations of this study, the following future research could be conducted on the treatment of swine waste with the ASBR:

1. This experiment was conducted at 20° C. The ASBR compensated for this lower temperature to a certain degree. The temperature of the reaction could be tried at a lower temperature such as 18° C or 15° C to determine at what temperature the ASBR stops becoming an effective treatment system.

2. It was observed that the introduction of a vacuum helped eliminate gas adhering to particles and floating them into the supernatant. The vacuum also improved the flocculating characteristics of the microorganisms. The introduction of a vacuum enhanced settling could be a big help in compensating for lower temperatures.

3. Since the reactor did better at a lower HRT, an even lower HRT might be investigated. Also, feeding two or three times a day instead of once in combination with the lower HRT would be an interesting investigation.

4. The reason for failure was probably due to microbial wash-out as the load was increased due to the high MLSS which may have "crowded out" the biomass. A study could be done where the MLSS is controlled to prevent the "crowding out" effect.

BIBLIOGRAPHY

1. Zeikus, G., and Johnson, E.A., *Mixed Cultures in Biotechnology*. McGraw-Hill: New York, 1991.

2. Holland, K.T., Knapp, J.S., and Shoesmith, J.G., *Anaerobic Bacteria*. Chapman and Hall: New York, 1987.

3. Parkin, G.F., and Owen, W.F., "Fundamentals of Anaerobic Digestion of Wastewater Sludges." *Journal of Environmental Engineering*, Vol 112:5, 1986, pp 867-920.

4. McCarty, P.L., "Anaerobic Waste Treatment Fundamentals: I. Chemistry and Microbiology." *Public Works*, No. 9, 1964, pp 107-112.

5. McCarty, P.L., "Anaerobic Waste Treatment Fundamentals: II. Environmental Requirements and Control." *Public Works*, No. 10, 1964, pp 123-126.

6. McCarty, P.L., "Anaerobic Waste Treatment Fundamentals: III. Toxic Materials and their Control." *Public Works*, No. 11, 1964, pp 91-94.

7. McCarty, P.L., "Anaerobic Waste Treatment Fundamentals: IV. Process Design." *Public Works*, No 12, 1964, pp 95 -99.

8. Seagren, E.A., Levine, A.D., Dague, R.R., "High pH effects in Anaerobic Treatment of Liquid Industrial Byproducts." *Proceedings of the 45th Industrial Waste Conference*, May, 1991, pp 377-385.

9. Buswell, A.M., and Mueller, H.F., "Mechanism of Methane Fermentation." *Industrial Engineering Chemistry*, Vol 44:3, 1952, pp 550-552.

10. Barker, H.A., "Biological Formation of Methane." *Industrial and Engineering Chemistry*, Vol 48:9, 1956, pp 1438-1442.

11. Starr, M.P., Stolp, H., Truper, H.G., Balows, A., and Schlegel, H.G., *The Prokaryotes - A Handbook on Habitats, Isolation, and Identification of Bacteria*. Chapter 76. Springer-Verlag: Berlin, Germany, 1981.

12. Pfeffer, J.T., Leiter, M., Worland, J.R., "Population Dynamics in Anaerobic Digestion." *Journal WPCF*, Vol 39, 1967, pp 1305-1322.

13. Van Velsen, A.F.M., Lettinga, D., "Anaerobic digestion of piggery waste. 3. Influence of temperature." *Netherlands Journal of Agricultural Science*, Vol 27, 1979, pp 255-267.

14. Dague, R.R., "Methane Gas Production From Swine Manure." Proceedings from the Waste Management Workshop, American Pork Congress, Des Moines, Iowa, March, 1974, pp 115-132.

15. Clark, R.H., and Speece, R.E., "The pH Tolerance of Anaerobic Digestion." *Advances in Water Pollution Research*. Volume 1. Pergamon Press: Oxford, England, 1971.

16. Albertson, O.E., "Ammonia Nitrogen and the Anaerobic Environment." *Journal WPCF*, Vol 33:9, 1961, pp 978-995.

17. Barth, E.F., Ettinger, M.B., Salotto, B.V., and McDermott, G.N., "Summary Report on the Effects of Heavy Metals on the Biological Treatment Process." *Journal WPCF*, Vol 37:1, 1965, pp 87-96.

18. Mosey, F.E., "Assessment of the Maximum Concentration of Heavy Metals in Crude Sewage which will not Inhibit the Anaerobic Digestion of Sludge." *Water Pollution Control*, Vol 75, 1976, pp 10-18.

19. Mosey, F.E., and Hughes, D.A., "The Toxicity of Heavy Metal Ions to Anaerobic Digestion." *Water Pollution Control*, Vol 74, 1975, pp 18-35.

20. Pidaparti, S.R., "Anaerobic Sequencing Batch Reactor Treatment of Swine Wastes at 35° C and 25° C." M.S. Thesis, Iowa State University, 1991.

21. Dague, R.R., McKinney, R.E., and Pfeffer, J.T., "Anaerobic Activated Sludge." *Journal WPCF*. Vol 38:2, 1966, pp 220-226.

22. Schroepfer, G.J., Fullen, W.J., Johnson, A.S., Ziemke, N.R., and Anderson, J.J., "The Anaerobic Contact Process as Applied to Packinghouse Wastes." *Sewage and Industrial Wastes*. Vol 27:4, 1955, pp 460-486.

23. Donham, K.J., Yeggy, J., and Dague, R.R., "Chemical and Physical Parameters of Liquid Manure from Swine Confinement Facilities: Health Implications for Workers, Swine and the Environment." *Agricultural Wastes*. Vol 14, 1985, pp 97-113.

24. Van Velsen, A.F.M., "Anaerobic digestion of piggery waste. 1. The Influence of detention time and manure concentration." *Netherlands Journal of Agricultural Science*. Vol 25, 1977, pp 151-169.

25. Daniels, L., "Biological methanogenesis: physiological and practical aspects." *Trends in Biotechnology*. Vol 2:4, 1984, pp 91-98.

26. Heukelekian, H., and Kaplovsky, A.J., "Effect of Change of Temperature on Thermophilic Digestion." *Sewage Works Journal*. Vol 20:5, 1948, pp 806-816.

27. Heukelekian, H., and Heinemann, B., "Studies on Methane -Producing Bacteria. I. Development of a Method for Enumeration." *Sewage Works Journal*. Vol 11:3, 1939, pp 426-435.

28. Heukelekian, H., and Heinemann, B., "Studies on Methane -Producing Bacteria. II. Enumeration in Digesting Sewage Solids." *Sewage Works Journal*. Vol 11:3, 1939, pp 436 -444.

29. Heukelekian, H., and Heinemann, B., "Studies on Methane -Producing Bacteria. III. Digestion of Sewage Solids by the Addition of Enriched Cultures of Methane-Producing Organisms." *Sewage Works Journal*. Vol 11:4, 1939, pp 571-586.

30. Heukelekian, H., and Heinemann, B., "Studies on Methane -Producing Bacteria. IV. Growth of Methane-Producing Organisms in Supernatant Sludge Liquors." *Sewage Works Journal*. Vol 11:6, 1939, pp 965-970.

31. Rudolphs, W., "Effect of Temperature on Sewage Sludge Digestion." *Industrial and Engineering Chemistry*. Vol 19:2, 1927, pp 241-243.

32. Pohland, F.G., and Bloodgood, D.E., "Laboratory Studies on Mesophilic and Thermophilic Anaerobic Sludge Digestion." *Journal WPCF*. Vol 35:1, 1963, pp 11-42.

33. McCarty, P.L., Jeris, J.S., Murdoch, W., "Individual Volatile Acids in Anaerobic Treatment." *Journal WPCF*. Vol 35:12, 1963, pp 1501-1516.

34. Wiegant, W.M., and Zeeman, G., "The Mechanism of Ammonia Inhibition in the Thermophilic Digestion of Livestock Wastes." *Agricultural Wastes*. Vol 16, 1986, pp 243-253.

35. Iannotti, E.L., and Fischer, J.R., "Effects of Ammonia, Volatile Acids, pH, and Sodium on Growth of Bacteria Isolated from a Swine Manure Digester." *Developments in Industrial Microbiology*. Volume 25, C.S. Nash, and L.A. Underkofler(eds.). Society for Industrial Microbiology, Arlington, Virginia, 1984.

36. Oleszkiewicz, J.A., "A Comparison of Anaerobic Treatments of Low Concentration Piggery Wastewaters." *Agricultural Wastes*. Vol 8, 1983, pp 215-231.

37. Sorlini, C., Ranalli, G., and Merlo, S., "Microbial Aspects of Anaerobic Digestion of Swine Slurry in Upflow Fixed-Bed Digesters with Different Packing Materials." *Biological Wastes*. Vol 31, 1990, pp 231-239.

38. Hill, D.T., and Bolte, J.P., "Authentic Fixed Media Reactor Performance Treating Screened Swine Waste Liquids." *Transactions of the ASAE*. Vol 31:5, 1988, pp 1525-1531.

39. Ng, W.J., and Chin, K.K., "Random-Packed Anaerobic Filter on Piggery Wastewater Treatment." *Biological Wastes*. Vol 20, 1987, pp 157-166.

40. Ng, W.J., and Chin, K.K., "Treatment of Piggery Wastewater by Expanded-Bed Anaerobic Filters." *Biological Wastes*. Vol 26, 1988, pp 215-228.

41. Colleran, E., Barry, M., Wilkie, A., and Newell, P.J., "Anaerobic Digestion of Agricultural Wastes Using the Upflow Anaerobic Filter Design." *Process Biochemistry*. Vol 17:2, 1982, pp 12-41.

42. Wilkie, A., and Colleran, E., "Start-Up of Anaerobic Filters Containing Different Support Materials Using Pig Slurry Supernatant." *Biotechnology Letters*. Vol 6:11, 1984, pp 735-740.

43. Wilkie, A., and Colleran, E., "Pilot-Scale Digestion of Pig Slurry Supernatant Using an Upflow Anaerobic Filter." *Environmental Technology Letters*. Vol 7, 1986, pp 65-76.

44. Hill, D.T., and Bolte, J.P., "Comparison of Surface -Scraped and Flushed-Screened Swine Waste as Methane Substrates for Thermophilic Digestion." *Transaction of the ASAE*. Vol 28:3, 1985, pp 870-874.

45. Hill, D.T., Bolte, J.P., and Prince, T.J., "Thermophilic Anaerobic Digestion of Screened-Flushed Swine Waste." *Transactions of the ASAE*. Vol 29:1, 1986, pp 229-234.

46. Bolte, J.P., Hill, D.T., and Wood, T.H., "Anaerobic Digestion of Screened Swine Waste Liquids in Suspended Particle-Attached Growth Reactors." *Transactions of the ASAE*. Vol 29:2, 1986, pp 543-549.

47. Zhang, R.H., North, J.R., and Day, D.L., "Operation of a Field Scale Anaerobic Digester on a Swine Farm." *Applied Engineering in Agricultural*. Vol 6:6, 1990, pp 771-775.

48. Wun-Jern, N., "A Sequencing Batch Anaerobic Reactor for Treating Piggery Wastewater." *Biological Wastes*. Vol 28, 1989, pp 39-51.

49. Summers, R., and Bousfield, S., "A Detailed Study of Piggery-Waste Anaerobic Digestion." *Agricultural Wastes*. Vol 2, 1980, pp 61-78.

50. Hobson, P.N., and Shaw, B.G., "The Anaerobic Digestion of Waste from an Intensive Pig Unit." *Water Research*. Vol 7, 1973, pp 437-449.

51. Fullen, W.J., "Anaerobic Digestion of Packing Plant Wastes." *Sewage and Industrial Wastes*. Vol 25:5, 1953, pp 576-585.

52. Heukelekian, H., "Thermophilic Digestion of Daily Charges of Fresh Solids and Activated Sludge." *Sewage Works Journal*. January, 1931, pp 3-23.

53. Cobb, S.A., and Hill, D.T., "A Comparative Analysis of Two Synthetic Media for Suspended Particle-Attached Growth Anaerobic Fermentation." *Transactions of the ASAE*. Vol 32:1, 1989, pp 223-231.

54. Flachowsky, G., and Hennig, A., "Composition and Digestibility of Untreated and Chemically Treated Animal Excreta for Ruminants - A Review." *Biological Wastes*. Vol 31, 1990, pp 17-36.

55. Kirchmann, H, and Witter, E., "Composition of Fresh, Aerobic and Anaerobic Farm Animal Dungs." *Bioresource Technology*. Vol 40, 1992, pp 137-142.

56. Hill, D.T., and Bolte, J.P., "Synthetic Fixed Media Reactor Performance Treating Screened Swine Waste Liquids." *Transaction of the ASAE*. Vol 31:5, 1988, pp 1525-1531.

57. Venkataraman, J., Kaul, S.N., and Satyanarayan, S., "Determination of Kinetic Constants for a Two-Stage Anaerobic Upflow Packed-Bed Reactor for Dairy Wastewater." *Bioresource Technology*. Vol 40, 1992, pp 253-261.

58. Cullimore, D.R., Maule, A., and Mansuy, N., "Ambient Temperature Methanogenesis from Pig Manure Waste Lagoons: Thermal Gradient Incubator Studies." *Agricultural Wastes*. Vol 12, 1985, pp 147-157.

59. Chen, Y.R., "Kinetic Analysis of Anaerobic Digestion of Pig Manure and its Design Implications." *Agricultural Wastes*. Vol 8, 1983, pp 63-81.

60. Durand, J.H., Fischer, J.R., Iannotti, E.L., and Miles, J.B., "Simulation and Optimization of Net Energy Production in a Swine Waste Digestion System." *Transactions of the ASAE*. Vol 30:3, 1987, pp 761-767.

61. Hobson, P.N., and Shaw, B.G., "The Bacterial Population of Piggery-Waste Anaerobic Digesters." *Water Research*. Vol 8, 1974, pp 507-516.

62. Coulter, J.B., Soneda, S., and Ettinger, M.B., "Anaerobic Contact Process for Sewage Disposal." *Sewage and Industrial Wastes*. Vol 29:4, 1957, 468-477.

63. Fernandes, L., McKyes, E., and Obidniak, L., "Performance of a continuous belt microscreening unit for solid liquid separation of swine waste." *Canadian Agricultural Engineering*. Vol 30, 1988, pp 151-155.

64. Noue, J., and Basseres, A., "Biotreatment of Anaerobically Digested Swine Manure with Microalgae." *Biological Wastes*. Vol 29,1989, pp 17-31.

65. Zeeman, G., Sutter, K., Vens, T., Koster, M., and Wellinger, A., "Psychrophilic Digestion of Dairy Cattle and Pig Manure: Start-up Procedures of Batch, Fed-batch and CSTR-type Digesters." *Biological Wastes*. Vol 26, 1988, pp 15-31.

66. Wrigley, T.J., Webb, K.M., and Venkitachalm, H., "A Laboratory Study of Struvite Precipitation after Anaerobic Digestion of Piggery Wastes." *Bioresource Technology*. Vol 41, 1992, pp 117-121.

67. Wong, M.H., "Anaerobic Digestion of Pig Manure Mixed with Sewage Sludge." *Biological Wastes*. Vol 31, 1990, pp 223-230.

68. Wilkie, A., and Colleran, E., "The Development of the Anaerobic Fixed-Bed Reactor and its Application to the Treatment of Agricultural and Industrial Wastes." *International Biosystems*. Volume III, D.L. Wise(ed.), CRC Press, Inc: Boca Raton, Florida, 1989, pp 183-226.

69. Fischer, J.R., Iannotti, E.L., and Porter, J.H., "Anaerobic Digestion of Swine Manure at Various Influent Solids Concentrations." *Agricultural Wastes*. Vol 11, 1984, pp 157-166.

70. Hazen, T.E., and Mangold, D.W., "Functional and Basic Requirements of Swine Housing." *Agricultural Engineering*. vol 41, 1960, pp 585-590.

71. Sykes, G., and Skinner, F.A., *Microbial Aspects of Pollution*. Academic Press: London, 1971.

72. Zehnder, A.J., *Biology of Anaerobic Microorganisms*. John Wiley & Sons: New York, 1988.

73. McCarty, P.L., and McKinney, R.E., "Salt Toxicity in Anaerobic Digestion." *Journal WPCF*. Vol 33:4, 1961, pp 399-415.

74. Dague, R.R., "Anaerobic Activated Sludge." Doctoral Dissertation, University of Kansas, Lawerence, Kansas, 1966.

75. McCarty, P.L., "Kinetics of Waste Assimilation in Anaerobic Treatment." *American Inst. of Biological Sciences*. Washington D.C., Vol 7, 1966, p 144.

76. Taiganides, E.P., "Characteristics and Treatment of Wastes from a Confinement Hog Production Unit." Doctoral Dissertation, Iowa State University, 1963.

77. Irgens, R.L., and Day, D.C., "Laboratory Studies on Aerobic Stabilization of Swine Wastes." *Farm Structures Engineering Report*. University of Illinois Champagne-Urbana, January, 1965.

78. Norman, J., "Depollution of Pig Waste Combined with Energy Recovery." *Global Bioconversions*. Donald L. Wise(ed.). CRC Press, Inc: Boca Raton, Florida, 1987.

79. *Standard Methods for Examination of Water and Wastewater*. 16th edition, American Public Health Association, Washington D.C., 1985.

80. Morrison, F.B., *Feeds & Feeding*. 21st edition, Morrison Publishing Co., Ithaca, New York, 1954.

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