TREATMENT OF CLAM-PROCESSING WASTEWATER USING UPFLOW ANAEROBIC SLUDGE BLANKET (UASB) TECHNOLOGY

by

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1.0 INTRODUCTION

Typically, in the United States, clam-processing plants discharge their process wastewater into surface waters that enter salt water creeks or estuaries. The clam-processing industry in Virginia has been mandated by the Virginia State Water Control Board (VSWCB) to improve plant effluent quality. These effluents contain high levels of organics in the range of 1,000 to 10,000 mg/L biological oxygen demand (BOD_5) and 1,000 to 10,000 mg/L total suspended solids (TSS). The proposed VSWCB's monthly average standards which range from 0 mg/L BOD_5: 0 mg/L TSS to 90 mg/L BOD_5: 90 mg/L TSS to as high as 500 mg/L BOD_5: 500 mg/L TSS depending on discharge impact studies. Industry compliance will require a significant investment by an industry which is already operating on a narrow profit margin (1). One large clam-processing facility in Virginia is currently installing an anaerobic wastewater treatment system.

Anaerobic treatment is potentially a low cost method for treating high strength seafood wastes. In any biological treatment system a major expense is the disposal and handling cost of excess biomass or sludge. The total volume of excess sludge may be reduced significantly by incorporating anaerobic processes which convert approximately 15% of the organic material metabolized to cell mass compared to about 50% conversion in aerobic processes (2). Significant sludge cost can also be saved if the wastewater is treated initially with an anaerobic process before polishing with an aerobic system. The savings with anaerobic pretreatment also include reduced aeration cost. During anaerobic digestion a large portion of the organic material in the wastewater is converted to methane gas which potentially could be used to defer plant energy costs.

Sludge disposal and energy costs have risen sharply over the last 10 years and thus have encouraged the development and utilization of new types of anaerobic systems. These new systems are classified as "high-rate" anaerobic processes because they contain highly active bacteria which are dense or attached. They usually operate at mesophilic (25 to 35 °C) or thermophilic (45 to 65 °C) temperatures and have low hydraulic retention times (HRT) of less than 8 hours. These high-rate methods have greatly improved the stability and efficiency of previous anaerobic treatment processes by separating the bacterial cell retention time from the HRT. These systems tolerate a high organic loading rate (OLR) which is the amount of organic material applied per liter of reactor volume per day. Generally the temperature of seafood processing wastewaters, such as effluent from clam-processing facilities with temperatures of 20 to 40 °C (3), are high enough to make high-rate anaerobic treatment economical. High-rate anaerobic systems contain a dense sludge. Typical aerobic mixed liquor suspended solids (MLSS) contain 3-5 kg volatile suspended solids (VSS)/m^3 compared to 30-50 kg VSS/m^3 in high-rate anaerobic systems (4).

One potential high-rate anaerobic system for treating clam-processing wastewater is the Upflow Anaerobic Sludge Blanket reactor (UASB). The UASB was developed in the Netherlands by Dr. Gatze Lettinga in 1977. The fundamental characteristic of the UASB system is a dense granulated sludge which settles well (i.e., Sludge Volume Index (SVI) approaching 20 ml/g) (5). The high cell density in the readily settling granules provides for a very long cell residence time at a short HRT. The UASB is a promising high-rate treatment system for the seafood processing industry due to its simplicity compared to some other high-rate reactors.

Figure 1 shows a schematic of the UASB reactor. Wastewater enters the distribution plumbing near the bottom of the reactor. The wastewater flows upwards through the blanket of granular sludge where the organic materials are degraded. The wastewater then enters a clarification zone which separates the liquid, biogas and solids. This process is called three-phase separation. Here the wastewater exits through the weirs, the biogas exits the top of the reactor, and any entrained granular sludge is settled back down to the sludge bed. The three-phase separator is designed to allow finely dispersed solids to exit with the wastewater so as not to be retained in the granular sludge bed.

The UASB may prove to be an economical option for the clam-processing industry compared to other treatment systems because UASB reactors:

1) have high organic loading rates which permits smaller and less expensive reactors compared to aerobic systems which become oxygen limited at high loadings (6),

2) eliminate the need for expensive biomass support materials (which also take up volume) required by attached film anaerobic reactors,

3) require fewer mechanical components such as mixers, pumps and aerators,

4) can operate at higher inhibitor concentrations, as compared to systems with shorter cell residence times or thinner biofilms,

5) allow extended down-times (more than a year) without feeding with no sludge activity loss (7),

6) yield methane gas which could be used to defer plant energy costs, and

7) produce less sludge than aerobic systems (granular sludge is stabilized and easy to dewater) (7).

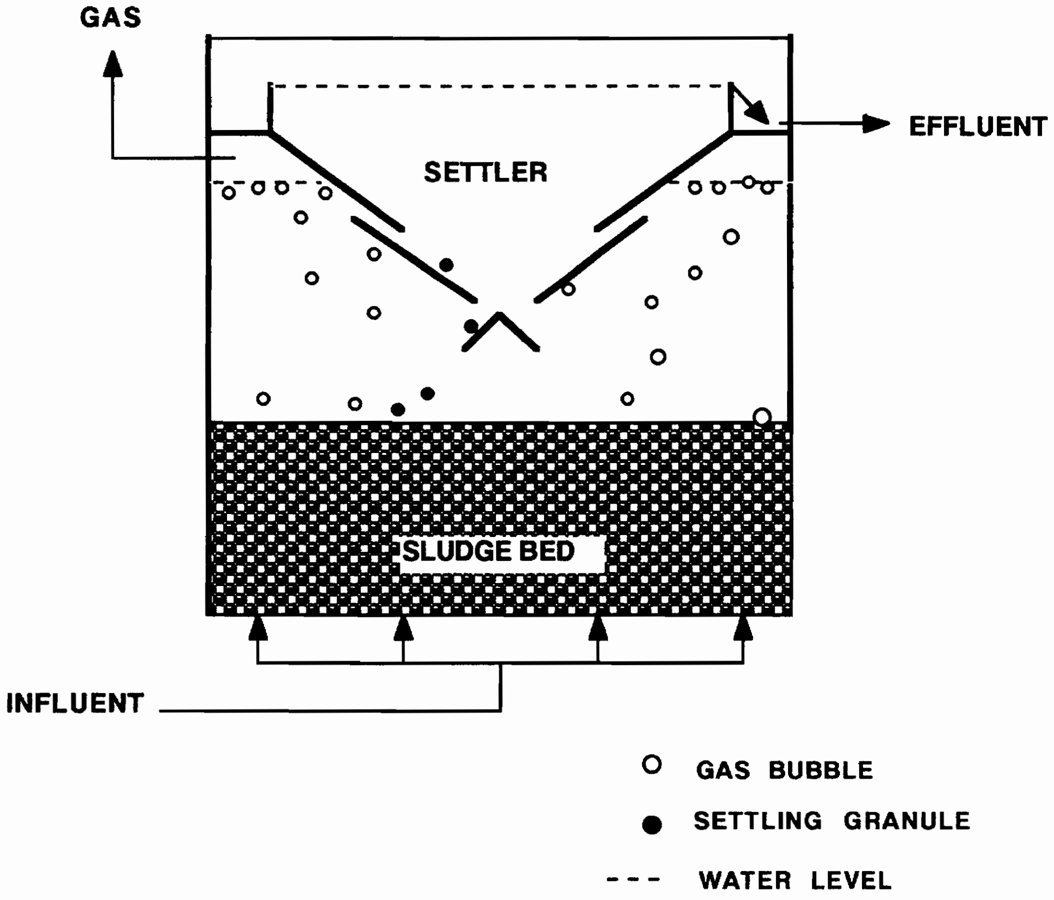

Figure 1. Schematic view of a Full Scale UASB Reactor Design.

Potential disadvantages of the UASB reactor include possible lack of adequate inoculum for start-up, potential washout of granules, sensitivity and higher effluent organic concentrations compared to aerobic systems.

The objective of this research was to establish the applicability of Upflow Anaerobic Sludge Blanket technology by testing the effluent from one of the Virginia clam-processing plants. The research goals were to achieve effluent quality required by VSWCB and to determine design and operating parameters for a UASB treatment process. This project was initiated with the view that the UASB treatment process might prove to be economical for the clam-processing industry.

In order to demonstrate the applicability of the UASB system to clam-processing wastewater, a two-phase research design was implemented. The first phase employed batch reactor studies. The second phase of the research involved the use of a continuous-flow, laboratory-scale, UASB reactor.

Phase A. Batch Reactor Studies

1) The biological methane potential (BMP) of the clam-processing wastewater was determined to evaluate the extent to which the wastewater could be degraded anaerobically.
2) The specific methanogenic activity (SMA) was determined to quantify the bacterial activity of the granular sludge utilizing clam-processing wastewater.

3) A comparison of biodegradability and methane production rates for clam-processing wastewater in the presence and absence of supplemental nutrients was evaluated.
4) Due to the brine used during processing, sludge inhibition and toxicity by sodium chloride (NaCl) was determined.

Phase B. Laboratory scale UASB

1) Biodegradability and methane potential measurements were performed to establish the anaerobic biodegradability of the waste.
2) The maximum organic loading rate was established to ensure the required BOD and TSS removal efficiencies.
3) System operation requirements were investigated to assist in a possible full scale design.

Since the practicality of the UASB process depends not only on the treatment efficiency but also on the volumetric capacity of the reactor (usually expressed as organic loading rate (OLR) with units of gram BOD applied per liter of the reactor per day), the focus of the experiment presented in this report was on determining the maximum OLR.

2.0 LITERATURE REVIEW

2.1 Clam Processing Industry

Prior to the 1987 Clean Water Act amendments, the clam-processing industry was required by the EPA to provide only minimal treatment of effluent by screening and settling for solids removal. Currently the VSWCB is writing National Pollution Discharge Elimination System (NPDES) permit regulations requiring secondary treatment of effluents. Draft permit regulations for one plant call for a maximum monthly average for BOD_5 and TSS to be 500 mg/L each. Daily maximum for BOD_5 and TSS is proposed at 700 mg/L each. Oil and grease requirements are 93 mg/L monthly average with a daily maximum of 400 mg/L. It is proposed that ammonia (NH_3) be monitored but no limits will be imposed at this time.

The clam-processing effluent arises from a combination of manual and mechanical shucking for the separation of clam meat and shells. Wastewater is generated during most stages of clam-processing such as washing and clean-up operations as well as from splashing and overflows from tanks and flumes (8). Wastewater flow is highly variable during a typical clam-processing day based on the amount of clams received, intermittent discharges and non-continuous operations. During one stage of clam-processing, brine tanks are used to float the clam meat away from the settling shells and grit. The brine tanks are dumped and cleaned at the end of the processing day which causes a surge in the salt loading and organic strength of the wastewater. The settled solids from the brine tanks are land filled.

Wastewater flow rates monitored in 1988 in three of Virginia's clam-processing plants ranged from 80,000 to 350,000 gallons per day (gpd). Peak flows up to 40,000 gallons per hour (gph) were observed. Average COD values for the three plants ranged from 3,000 to 7,000 mg COD/L. The TSS values ranged from 1,000 to 10,000 mg/L (8).

Libelo (3) investigated the production of wastewater from several mechanized clam-processing plants in Virginia. He determined that the wastewater generally consisted of fresh water used in processing which contained organics of widely varying concentrations, sand and shell material. It was reported that the COD ranged from 1,000 to 5,000 mg COD/L with spikes as high as 27,000 mg COD/L. Libelo measured the suspended solids in the wastewater and found that it varied widely from 6,000 to 30,000 mg TSS/L. Anaerobic batch studies determined that a COD reduction of 75% was possible using a seven day retention time with about two liters of methane being produced from one liter of clam wastewater, equivalent to a decrease in COD from 21,400 to 5,500 mg COD/L. Libelo determined that the size of the waste treatment system was critical to its feasibility due to very limited space at the various plants.

Guida and Kugelman (9) investigated the treatment of clam-processing wastewater by activated sludge and salt marsh (wet lands) polishing. They determined that the BOD_5, solids, and nutrient removal efficiencies were potentially 100%.

2.2 Anaerobic Fundamentals

A knowledge of anaerobic fundamentals is very important for the design and operation of efficient anaerobic reactors and for an understanding of how upsets occur and how to alleviate them. Methanogens are considered the key link in the chain of microorganisms participating in the anaerobic digestion process.

2.2 1 Microbiology

Complex organic wastes are consumed by a consortium of anaerobic microbes such that the anaerobic degradation of organics is shared by several types of bacteria. Figure 2 shows a schematic of the anaerobic degradation of organic matter (10). First, the complex organic polymers are enzymatically reduced (hydrolysis) to volatile organic acids, alcohols, hydrogen (H_2) and carbon dioxide (CO_2) by hydrolytic bacteria. Then hydrogen producing acetogenic bacteria ferment the volatile organic acids (VOA) and alcohols for their metabolic needs subsequently excreting acetic acid, one-carbon compounds, H_2 and CO_2 (acidogenesis). The homoacetogenic bacteria synthesize acetic acid from H_2 and CO_2 (acetogenesis). Finally, the homotrophic acetoclastic methanogenic bacteria metabolize acetic acid to methane and the autotrophic methanogens utilize the H_2 to produce methane (methanogenesis).

Waste degradation depends on the interdependent metabolisms of anaerobic bacterial groups. One example of this is the interaction between H_2-

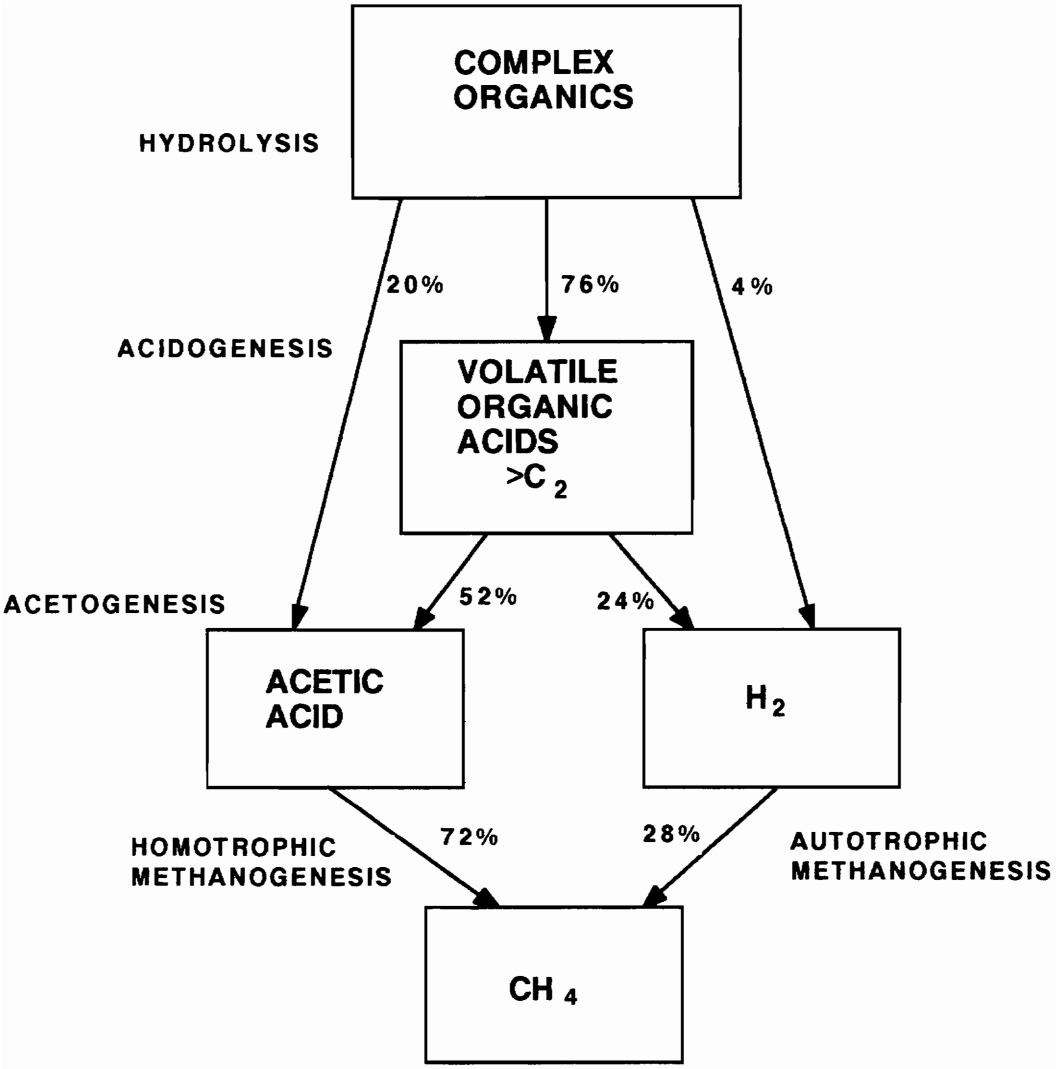

Figure 2. Schematic view of the Anaerobic Degradation of Complex Organic Material to Methane Gas.

forming acetogenic bacteria and the H_2-consuming homoacetogenic bacteria and autotrophic methanogenic bacteria. This relationship is called obligate proton reduction (11). A H_2 partial pressure above $10^{-4.5}$ atmospheres was reported by Dubourgier (12) and McInerney (13) to inhibit methanogenesis. Another relationship critical to digester performance is the utilization of acids by methanogens. If the methanogens can not metabolize the acids produced by other bacterial groups then the acids can accumulate. Disruption of this interdependency is one of the major causes of digester "souring," or build-up of acids, resulting in low pH.

Excess sludge production is considerably less in anaerobic systems than in aerobic systems as measured by the net volatile suspended solids (VSS) production. According to Eckenfelder (14), most wastewaters in an activated sludge system have a net sludge yield of 0.5 kg VSS produced per kg of COD consumed. By contrast, in an anaerobic system the excess sludge yield is approximately 0.1 kg VSS/kg COD consumed. The difference in cell growth is attributed mostly to the energy available in the anaerobic and aerobic environments. Under anaerobic conditions no oxygen is available as an electron acceptor and, thus, carbon atoms are used. This reaction is inefficient and results in end products that still contain energy such as methane.

In anaerobic reactors the free energy change is small compared to that of aerobic degradation. Wolfe (15) reported that in the anaerobic conversion of one mole of glucose to methane and carbon dioxide 5 moles of ATP per mole of glucose are produced, as compared to 36-38 moles of ATP per mole glucose through aerobic metabolism. The energy is conserved in the methane gas during anaerobic digestion. This is the reason for the smaller sludge yield in anaerobic biomass compared to aerobic sludges. The major anaerobic electron acceptor for respiration is CO_2, compared to O_2 for aerobes.

Balch et al. (16) published studies on methanogens' unique DNA structure, intermediate metabolism, and lipid composition which led to a reclassification of methanogens. Methanogens currently belong to the group of prokaryotes called Archaebacteria, or ancient bacteria, believed to be among the earliest life forms, almost 4 billion years old (17). The uniqueness of methanogens is found also in three electron transfer coenzymes specific to methanogens. One of which, F420, has been successfully used to monitor the activity of methanogens in anaerobic reactors due to its fluorescence and adsorption peak at 420 nanometers (10).

2.2.2 Temperature

Temperature is a very important operational parameter in anaerobic digestion. As temperatures increase the rates of metabolism increase within two important temperature ranges. Mesophilic reactors operate most efficiently within a temperature range of 22 - 40 °C. Figure 3 shows the mesophilic sludge activity of *Methanothrix soehngenii* as a function of temperature (18). The critical temperature above which bacterial death occurs (40 °C) is precariously close to the maximum growth rate temperature for this mesophil (37 °C). Thermophilic anaerobic digestion (45 - 65 °C) is similar to mesophilic digestion except the bacteria grow two to four times faster (19). One advantage to the higher temperatures is the potential destruction of pathogens. Full scale thermophilic reactors are currently in use. However, a

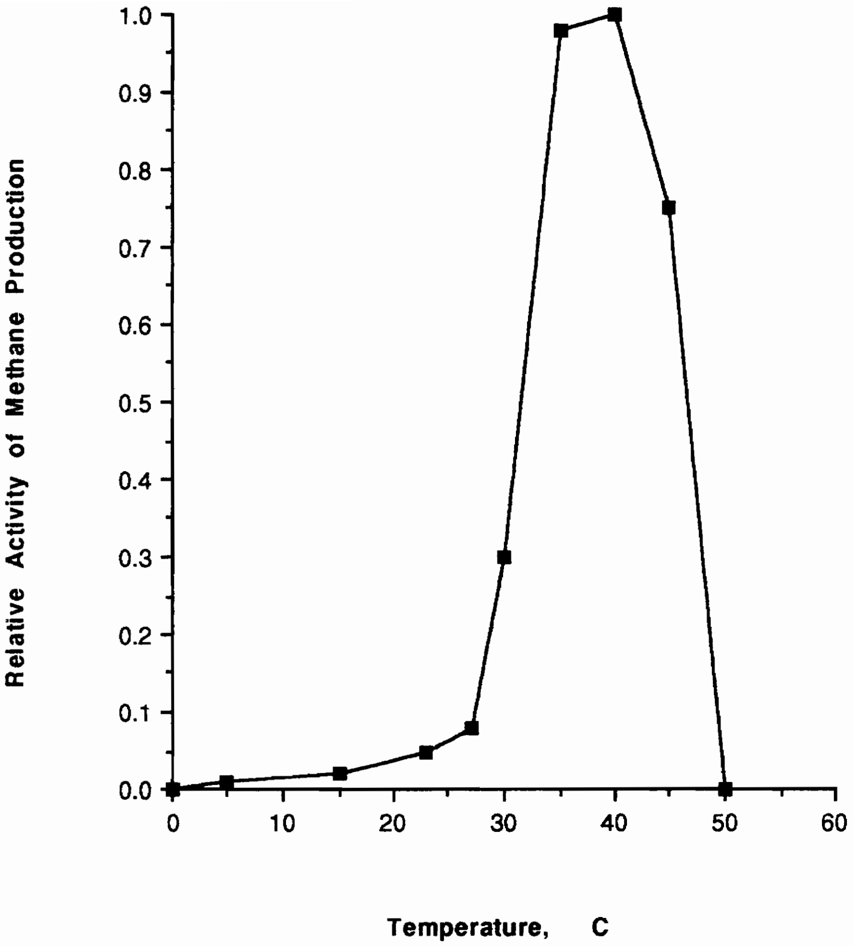

Figure 3. The Activity of Granular Sludge as a Function of Temperature.

significant disadvantage with these reactors is the increased energy input for heating.

2.2.3 pH Effects

Most anaerobes grow best at neutral pH. However, methanogens are particularly sensitive to pH changes. The optimal pH range for methanogens, according to McCarty (20), is 7.0 to 7.2. Low pH conditions (below pH 6.6) result in incomplete metabolism and, thus, a build-up of volatile acids which further drops the pH. At a pH of 6.2 the bacteria experience acute toxicity from the acids. Swings in pH can cause compounds to become toxic by dissociation or combining such as sulfide and ammonium.

An ion bicarbonate equilibrium with the dissolved CO_2 in the reactor helps to keep the pH stable. The CO_2 reacts with hydroxide ions (OH^-) to produce bicarbonate (HCO_3^-) (21). This is called buffer capacity. The generation of CO_2 in the biogas establishes the bicarbonate buffering system. Stumm and Morgan discussed the hydration of CO_2 leading to the formation of H_2CO_3 and the buffering effects of the carbonate system (22).

Alkalinity has been used as an indication of the buffer capacity in anaerobic digesters. Alkalinity of a wastewater is its acid neutralizing capacity and is commonly a function of ammonia, carbonate, bicarbonate and hydroxide concentrations. Alkalinity monitoring will indicate the buffering capacity of the system against acid shocks. According to Standard Methods (23), a properly operating batch anaerobic digester for municipal sludge digestion has an alkalinity greater than 2000 mg/L as $CaCO_3$.

Monitoring the volatile organic acid (VOA) concentration in the effluent either by gas chromatography or by titration methods can serve as a predictive control parameter. The build-up of VOA to toxic concentrations is called reactor souring. Organic acids may staart accumulating before a pH drop can be measured due to the buffering capacity. Observing the build-up of acids and subsequently lowering the organic load to the reactor can prevent souring.

2.2.4 Toxicity in Anaerobic Systems

Toxic compounds can inhibit or poison bacteria by slowing or shutting down their metabolism. The extent of toxicity is usually a function of concentration and pH. Even an essential nutrient can become toxic at high enough concentrations. Methanogens are usually the most sensitive to toxicity. In anaerobic systems toxicity studies are critical in determining process control parameters to prevent inhibition of biological treatment. The process of wastewater neutralization with salts such as sodium hydroxide potentially can add to the toxicity of the wastewater.

Sodium caused inhibition in granular sludge batch assays according to Rinzema (24). At a neutral pH, sodium levels of 5, 10 and 14 g Na^+/L caused inhibitions of 10, 50 and 100%, respectively, when compared to the maximum specific methanogenic activity of granular sludge. Rinzema also reported that sodium toxicity was more severe at pH 8 than at neutrality.

Toxic effects are usually attributed to the cation rather than the anion. McCarty (25) reported that monovalent cations are generally less toxic than divalent cations. He studied anaerobic batch reactors and found that the activity of acetate utilization was decreased significantly by the addition of NaCl. However, McCarty also discovered that the addition of calcium chloride reduced the toxic effects of the sodium chloride. Magnesium was also found to be antagonistic to salt toxicity. Table 1 shows the effects of several cations on methanogens as reported by McCarty (26).

Several anions have been used in laboratory studies that may counteract the toxicity of sodium according to Kugelman and Chin (27). However, it is difficult to determine the effects of several anions combined since they may cause both antagonism and synergism depending on their relative concentrations. Potassium ions have been reported by Kugelman to be either antagonistic or synergistic with respect to sodium toxicity, depending upon the presence of other anions.

The results of several researchers investigating the toxicity of NaCl in anaerobic systems are shown in Table 2. The original paper, written by Rinzema (24), from which Table 2 was extracted contains a detailed list of references. The tabulated data show that anaerobic bacteria are sensitive to Na^+ at concentrations above approximately 4 g/L (0.4% Na^+) resulting in approximately 10% reduction in activity. At concentrations above approximately 9 g/L (0.9% Na^+) a complete cessation of methane production was observed in granular sludge batch experiments.

Table 1. Inhibition Effects of Cations on Methane Fermentation (McCarty, 26)

Cation	Stimulatory Concentration (mg/L)	Moderately Inhibitory Concentration (mg/L)	Strongly Inhibitory Concentration (mg/L)
Calcium	100-200	2,500-4,500	8,000
Magnesium	75-150	1,000-1,500	3,000
Potassium	200-400	2,500-4,500	12,000
Sodium	100-200	3,500-5,500	8,000

Table 2. The Anaerobic Toxicity of Na^+ added as NaCl for the Inhibition of Methane Formation (24).

Sodium Concentration, g /L Inhibition Percentage 10%	 50%	 100%	 Experimental Conditions Substrate, Biomass, Reactor
NA[a]	10	NA	C_2[b], granular sludge, Batch
NA	7	NA	C_2, granular sludge, Batch
NA	4	NA	C_2:C_3:C_4, granular, Batch
NA	8.7	NA	C_2, domestic sludge, Batch
5	10	14	C_2, granular sludge, Continuous
4.5	6.3	9	C_2, granular sludge, Batch
4.8	4	10[c]	C_2, NA, CSTR
6.1	7.4	8[c]	C2, NA, CSTR (15 days)
7	8[c]	NA	C_2, NA, CSTR (70 days)
3.6	6.6	NA	C_2, NA, Batch (shock exposure)
6.7	8.7	NA	C_2, NA, Batch (shock exposure)
3[c]	6	10-11	Dairy, NA, Batch (shock exposure)
13-16	21-24	NA	Sauerkraut waste, granular, UASB
4	7	10[c]	Tomato waste, NA, CSTR
12.5	15	NA	C_2 & ethanol, NA, Anaerobic filter
4.5	6.3	9[c]	C2, granular sludge, Batch

a) not available, b) C_2, acetic acid; C_3, propionic acid; C_4, butyric acid, c) approximate value

2.2.5 Anaerobic Nutrient Requirements

An important environmental condition for efficient anaerobic digestion is the presence of required nutrients in adequate and available concentrations. Recent studies of trace metal requirements for anaerobes have led to the development of more stable reactors through supplementation with iron (Fe), nickle (Ni), manganese (Mn), cobalt (Co), tungsten and selenium. Guiot (28) determined that significant increases in sludge activity were caused by supplementing with trace metals (Fe, Ni, Co, Mn) and concluded that increases were due to increases in both enzymatic cell activity and the number of viable cells in the biomass.

Anaerobic systems are being applied to industrial wastewaters which generally have fewer nutrients than municipal domestic sludge. Speece (29) reported that the anaerobic digestion and stabilization of domestic sludges have not shown nutrient deficiencies. The probability of nutrient deficiency with industrial waste streams is greater and supplementation with nutrients usually is necessary. Speece reported that the presence of required nutrients in the wastewater does not ensure their bio-availability. Nutrient supplementation of apparently adequate nutrients already in the wastewater has produced increased bacterial activity.

2.2.6 Methane Production

The amount and rate of methane derived from COD consumption is a function of temperature, pressure, substrate concentration, nutrients and the presence of toxic materials.

McCarty and Young (30) calculated that 0.348 liters of methane can be produced from 1.0 g of COD consumed at standard temperature and pressure (STP). Methane gas production can be converted from volume (liters) to COD (mg) from the oxygen equivalent formulation as derived by McCarty (31):

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \qquad [1]$$

such that 1 mole (16 grams) of methane is equivalent to 64 grams of COD (2 moles of O_2). The volume of methane gas produced per gram of COD (or BOD_5) removed is calculated by: 1 mole of any gas = 22.4 L (at 0°C and 1 atm) and, thus, 22.4 L of methane (1 mole) equals 64 grams of COD such that 0.348 liters of methane are produced per gram of COD removed. The theoretical COD content of the methane (CH_4-COD) produced is thus 1 gram COD in every liter of gas at STP.

The volume of methane produced at other temperatures can be calculated using Charles' Law:

$$V_2 = V_1 \frac{T_1}{T_2} \qquad [2]$$

where,

V_2 = volume of gas produced at reactor temperature, L
V_1 = volume of gas at standard conditions (22.4 L),
T_1 = temperature at standard conditions (273 K),
T_2 = temperature at reactor conditions, K.

Thus, the COD of methane at 32 °C is calculated to be 0.39 L CH_4 per gram COD or 1 liter of CH_4 from 2.574 grams of COD.

Monitoring toxicity includes observing methane yields over time and observing reductions in those yields. The installation of a cumulative gas meter will allow for the determination of organics removed in terms of methane produced. Two general methods for measuring methane in the laboratory include liquid displacement and gas pressure measurements. Several researchers have used McCarty's formula to convert the liquid displacement measurement of methane into a BOD_5 value in order to more easily compare the production of methane to the disappearance of BOD_5 from the wastewater, (10, 21, 32).

The use of biogas in industry is complicated by the corrosive potential of H_2S forming H_2SO_4. However, the removal of H_2S has been accomplished by "iron sponge" technology at low cost (33).

2.4 Anaerobic Treatment Systems

The anaerobic lagoon is the most commonly used anaerobic technology. Generally they are uncovered ponds into which wastes are discharged without mixing or temperature control. The performance is extremely variable. The performance of lagoons has been improved by using coverings, insulation, solids recycle and biogas collection (34).

Three basic configurations of anaerobic reactors are shown in Figure 4. The most common type of anaerobic reactor is the completely stirred tank reactor, or CSTR, as shown in schematic (a). The CSTR consists of a tank with mechanical mixing. The mean cell residence time in a CSTR is equal to the HRT. Hydraulic residence times of 15 to 30 days are common, resulting in large reactor volumes.

The second configuration shown in Figure 4 is the anaerobic contact process (b). The flow regime is similar to the activated sludge process. By settling and recycling sludge the mean cell residence is greater than the HRT. A degassing chamber is used to aid settling in the clarifier. Since the mean cell residence time has increased, the efficiency of BOD_5 removal is increased. Also, the effluent will contain less TSS than the effluent from the CSTR. Problems with the anaerobic contact process include poor settling of solids due to gas saturation and subsequent release in the clarifier resulting in turbulence.

The third reactor configuration in Figure 4 is the high-rate immobilized biomass type reactor (c). Immobilized biomass reactors contain bacteria concentrations that are independent of the hydraulic flow rate through attachment or suspension. Because the mean cell residence time is much greater than the HRT, these reactors are termed high-rate.

Anaerobic treatment has been assumed to be less tolerant of toxic materials than aerobic treatment. However, this assumption is not always true. Anaerobic treatment technologies may yield greater treatment of some toxic materials than aerobic processes (35). For example, Shink (36) reported that the anaerobic degradation of certain compounds proved to be more efficient

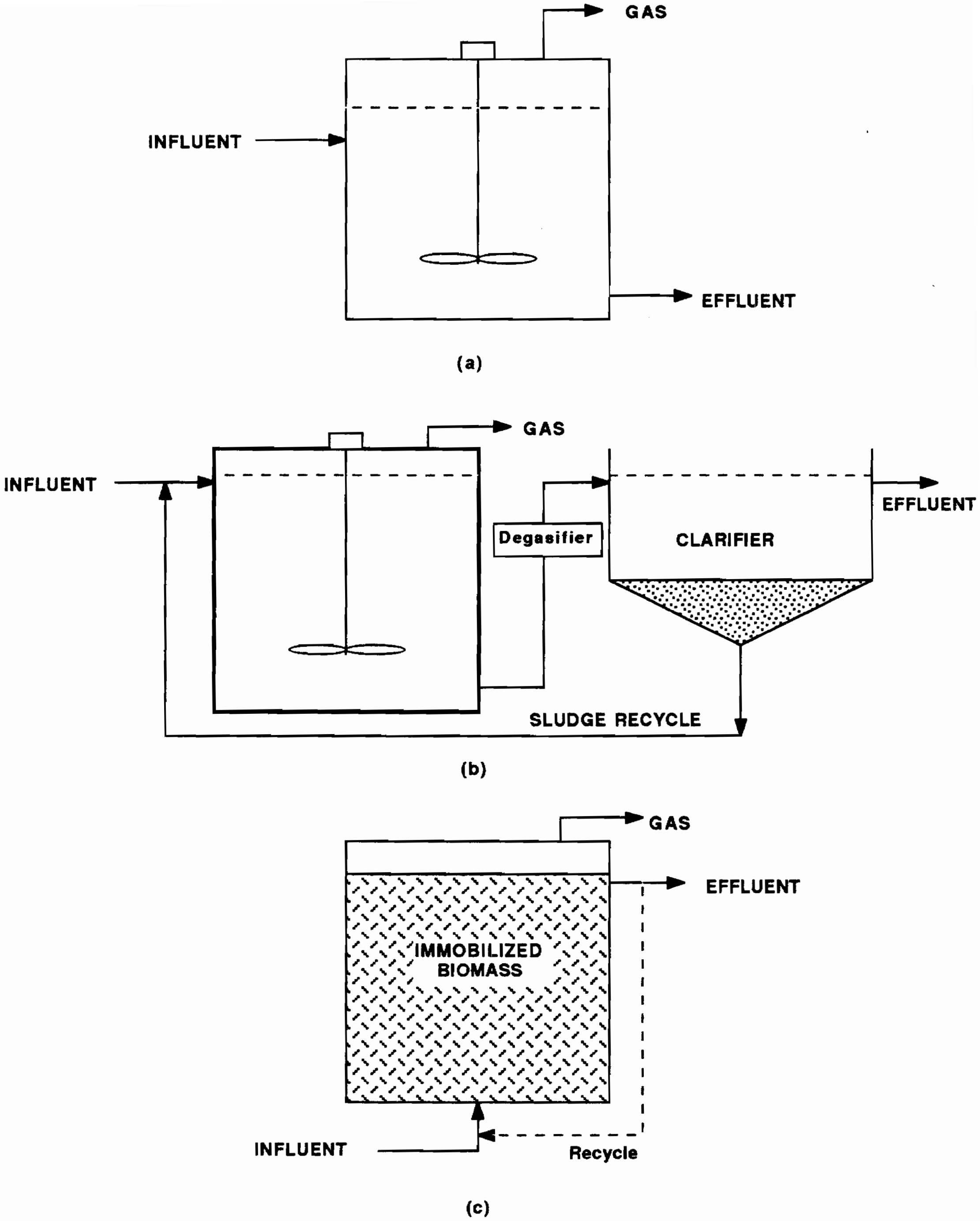

Figure 4. Typical Anaerobic Reactor Configurations: (a) Continuously Stirred Reactor, (b) Anaerobic Contact Process, (c) Attached Film Reactor.

and produced less toxic by-products than the aerobic degradation. Vogel (37) reported the anaerobic degradation of chloroform, trichlorethane and tetrachlorethene, however, no aerobic microorganism was capable of these degradations. The anaerobic reductive dechlorination and dehalogenation pathways of certain compounds was reviewed by Holliger (38). Holliger discussed the untapped potential of anaerobes to mineralize recalcitrant compounds.

2.4.1 High-rate systems

High-rate anaerobic systems use immobilized biomass to achieve high volumetric loading rates and low hydraulic retention time (HRT). Several high-rate systems are currently in use.

1) The Anaerobic Filter, developed by McCarty and Young (39), uses a stationary support material to which bacteria attach. These reactors can be troubled with clogging.

2) The Anaerobic Fixed-Film Expanded Bed, developed by Switzenbaum and Jewell (40), uses a light support material that is expanded (fluidized) in the upward flow of wastewater. Binot (41) also developed a fluidized bed type reactor. These reactors can have high pumping costs.

3) The UASB, developed by Lettinga (42), uses granulated sludge which settles well. These reactors are slow to develop the necessary pelletized biomass if none is available for inoculation from another reactor.

Several combinations and hybrids of these systems have been studied and built for industrial wastewater treatment. One example is the modified sludge blanket reactor (43) consisting of an anaerobic stationary filter above a granular sludge bed. Lettinga (44) compared the operation of the UASB, the anaerobic filter, and the anaerobic attached film expanded bed and the anaerobic fixed film process. He determined that the UASB reactor was more stable.

An expanded bed reactor using granulated activated carbon (GAC) as the support material was studied by Suidan (45). He reported that the GAC adsorbed shock loads resulting in stable reactor operations.

2.4.2 Low-rate systems

Low-rate reactors are essentially anaerobic lagoons in tanks. Low-rate anaerobic systems operate at lower temperatures, degrade wastes at a slower rate, and are loaded to a much lower degree than high-rate systems. Low-rate systems have been used for raw wastes containing high concentrations of suspended solids such as in the stabilization of waste activated sludge from aerobic systems. Generally, low-rate systems are considered if the total suspended solids (TSS) contribute more than 50% to the TCOD of the waste according to Robertson (46). Low-rate systems are loaded much less heavily at approximately 1 g COD/L/d compared to about 15 g COD/L/d for high-rate systems. Typical hydraulic retention times are 10 to 30 days requiring larger reactor volumes than high-rate systems with HRTs less than 8 hours. However, as a result of the reactor volume being larger, it can offer a waste solids retention time of up to two years. The large reactor volume also allows for storage of excess sludge.

2.5 UASB Reactors

Due to the relative newness of the UASB system, designers have used very conservative loadings. Hence, it is very important to test the characteristics of each wastewater, even those generated from similar industries in order to design the UASB system. Laboratory analysis provides the basis from which the proper design of the UASB and other high-rate systems are obtained.

The three phase separation of biogas, effluent and biomass is crucial for operation. The proper design for settling the granules is a key element of the UASB design. The three phase separation device can be simple, but should provide an adequate settling zone for retention of granular sludge in the reactor (47).

The UASB is most easily applied to relatively soluble waste. Generally, the hydrolysis step is the rate limiting step for insoluble waste fractions. The loading rate is necessarily lower for partially insoluble waste .

2.5.1 Biomass Granulation

The mechanism of granulation of anaerobic sludge is unknown (48). Most explanations for granulation include both environmental pressure and the kinetic growth characteristics of specific bacteria species. Growth of bacteria is often discussed in terms of kinetics defined by rates of substrate utilization and rates of growth. Substrate utilization, μ, is the rate at which a specific species of bacteria can remove organic material from the environment. The half velocity constant, K_S, is the concentration of organic material at which the bacteria are using the organic material at half the maximum rate (49). Selective environmental pressures such as substrate concentration in a continuous-flow reactor are highly influential in determining the dominant methanogenic species.

According to Zehnder (18), the type of methanogen most suited for granulation is of the *Methanothrix* type. Rinzema (24) discovered that the most stable granular sludge is composed predominantly of *Methanothrix soehngenii* with the kinetic constants μ equal to 0.1 d^{-1} and K_s equal to 0.2 mM for acetic acid (12 mg BOD_5/L). In direct competition for substrate with *Methanothrix soehngenii* is *Methanosarcina barkeri* with kinetic constants μ equal to 0.45 d^{-1} and K_s equal to 5 millimolar (mM) of acetic acid (300 mg BOD_5/L). Kinetics dictates that a lower substrate level will select for the methanogen with the lower Ks value (50). The methanogen with the lower K_s value demonstrates a stronger affinity and utilization of substrate at low concentrations, which can be an advantage in wastewater treatment (18). Thus, granular sludge is developed at low substrate concentrations in order to select for *Methanothrix soehngenii.*

2.5.2 UASB Operations

A maximum effluent total VOA concentration of 5 milliequivalents per liter (325 mg BOD_5/L as VOA) is used in the design and operation of full scale UASB reactors (51). According to Lettinga (52), the organic loading on the UASB during start-up or after up-set can be increased incrementally when an organic consumption rate of 80% has been consistently observed.

The methane production capability of a UASB reactor is reported in terms of the amount of methane produced in COD units per unit volume of the reactor per unit time. Volumetric gas production rates from the literature averaged 10 to15 g CH_4-COD per liter of reactor volume per day and the specific gas production rates were 1 g CH_4-COD per g VSS of sludge per day (53). The monitoring of biogas production from the UASB is a direct and reliable indicator of the performance of the UASB. If the biogas prduction is less than normal then inhibition may be possible, if the biogas production is higher than normal then an organic overload may be possible.

2.5.3 Biomass Inoculation of UASB Reactors

According to Lettinga (52), the current practice for reactor inoculation is to seed new UASB reactors with granular sludge cultivated in other operating UASB systems. The advantage of inoculating with granulated sludge rather than sewage sludge is a greatly decreased start-up time. Often the wastewater composition is different, thus the environmental conditions of the granular sludge can change in the new treatment system. Lettinga discussed the possible problems that might be encountered when granular sludge is exposed to a different wastewater including: erosion of granule integrity, sludge wash-out, reduced specific methanogenic activity, and gas release problems resulting in flotation of sludge. Generally, the amount of granular seed sludge used to start up new UASB reactors is 10-15 kg VSS/m3. This amount helps ensure that the sludge bed is below the settler compartment and avoids excess washout (10).

2.5.4 Upflow Superficial Wastewater Velocity and HRT in UASB Reactors

The superficial flow velocity (m/hr) of the wastewater flowing upwards through the UASB reactor is the flow rate (m^3/hr) divided by the cross sectional area of the reactor (m^2). For granular sludge, the superficial flow velocity (SFV) of the wastewater cannot be higher than approximately 10 m/hr in order to prevent granular sludge washout (52). Thus, the critical rule used in the design of the wastewater flow was that the SFV through the narrowest part of the reactor must be much less than the granular sludge's downward settling velocity to prevent entrainment and washout of granules.

The HRT is proportional to the superficial flow velocity for a minimum reactor cross sectional area (Z) and reactor volume (V) such that:

$$SFV = \frac{V}{Z * HRT} \qquad [3]$$

where:

SFV = Superficial Flow Velocity, m/hr
V = Volume, m^3
Z = Cross sectional area, m^2
HRT = Hydraulic Retention Time, hr

Another design parameter which effects SFV and HRT is the reactor height limitation. Typical UASB reactors are 6 to 8 meters in height. Most UASB designers do not exceed a height of 10 meters to limit the water head pressure above the granules which might cause diffusion limitations for methane release from the bacteria cells or granules (53).

According to Kooijmans (54), HRT's for municipal wastewaters as low as 2 hours were successfully used in Cali, India.

2.5.5 Dilute Organic Strength Wastewaters in UASB Reactors

A wastewater treatment system for low-strength wastewater (less than 1 g BOD_5/L) is economical if large volumes of the wastewater can flow through the system in short periods of time. The use of fixed film or immobilized biomass systems provides the desired bacteria retention resulting in a significant reactor volume reduction for a given treatment efficiency. Efficient dilute strength organic wastewater treatment has been achieved using UASB reactors.

UASB systems have been used successfully for wastewaters which are relatively dilute, such as municipal wastewaters. Lettinga (55) proved that the UASB will perform well during low loading (less than 1 g COD/L/d) and very short HRT (approximately 2 hours).

The results are inconclusive as to whether extending the retention time, as in low-rate reactors, improves COD removal efficiency for dilute wastewaters (56).

2.5.6 Granular Sludge Production

Anaerobic sludge yields for acetate digestion were reported by Novak and Ramesh (57) to range from 0.12 to 0.4 g VSS/g COD. According the Henze and Harremoes (58), the average cell yield coefficient for granular sludge is 0.18 g VSS/g COD removed at a granular sludge activity rate of 1 to 1.5 kg COD/kg VSS/d. This yield value for a mixed culture is based on adding a yield coefficient of 0.15 for acidifying bacteria and a yield of 0.03 for methanogens.

A mathematical model to calculate volatile solids production in an anaerobic CSTR was presented by Qasim, Warren and Udomsinrot (59) based on reactor gas production:

$$X = \frac{(E\,S - G/5.62)}{1.42} \qquad [4]$$

where,

X = volatile biological solids produced per day, mg/L/d,
G = CH_4 produced per day (STP), ft^3
E = efficiency of BOD_5 utilization, and
S = BOD_5 added per day, lbs/d.

Paasschens, de Vegt and Habets (60) reported the excess sludge production from a full-scale UASB reactor as 0.04 kg solids per kg BOD_5 removed. They reported the results of five years of operating a 2200 m^3 UASB consuming wastewater from a pulp paper mill having a flow of 600 - 650 m^3/hr. The HRT in the reactor was 3.4 hours and the volumetric loading was 10 - 12 kg COD/m^3/d. The excess granular sludge was used to start new UASB reactors. Before installation of the UASB system, the excess sludge was approximately 0.60 kg dry solids/kg BOD_5 consumed from an aerobic system. After installation of the UASB in 1985, the aeration plant was retained as a polishing basin The total excess solids production from the polishing basin were reduced to 0.15 kg solids per kg initial BOD_5 load after pretreatment with the UASB.

According to Kooijmans, Lettinga, and Rodriguez (54), excess granular sludge is easy to dewater. They reported that tests on drying beds resulted in a dry matter content of 25% after two days and 40% after seven days.

3.0 MATERIALS AND METHODS

3.1 Wastewater Description and Sampling

Clam-processing wastewater was obtained from a facility which processes ocean quahogs, *Arctica islandica,* and surf clams, *Spisula solidissima*. Production hours were generally limited to one 8 hour shift, 6 days a week. Production was highly dependent on weather conditions and, thus, lengthy down times might be expected in waste treatment. Samples for this study were collected during the first third of the processing shift, before brine tanks were dumped. Since the brine tanks are a small portion of the total wastewater flow and would essentially contribute a short shock loading, their contribution was not incorporated into the samples. Wastewater was transported in 5 gallon polypropylene carboys at ambient temperatures. The samples were stored frozen at -5 °C until used. The wastewater was thawed, and then clarified by gravity sedimentation for 30 minutes before use in laboratory studies.

3.2 Granular Sludge Biomass

Fresh granular sludge was obtained from an operating UASB reactor at the Floyd Agricultural Energy Corporation (FAEC) in Floyd, VA. A one gallon sample of granular sludge was taken from the 10 foot sample port at the FAEC's UASB. A portion of the sludge was used to start the laboratory UASB at

Virginia Polytechnic Institute and State University. Aliquots of the granular sludge were stored at 4 °C for future analysis and batch experiments.

A schematic of the FAEC UASB is presented in Figure 5. FAEC obtained their granular sludge from Heilerman Brewery, La Crosse, WI in 1986. Heilerman grew their granular sludge from municipal, anaerobic stabilization sludge. The FAEC plant produces anhydrous ethanol from various soluble wastes such as corn syrup and Coca-Cola® syrup. A large fraction of the FAEC wastewater contains ethanol and other by-products of distillation. Inhibition of the granular sludge at FAEC has been noted when caramel type compounds from the Coca-Cola® syrup pass through the UASB system.

3.3 Granular Sludge Characteristics

Figure 6 depicts the apparatus used to elutriate (wash) the granular sludge at ambient laboratory temperatures to remove fine particles. Elutriation is a process of washing, decanting and settling which separates a suspension of solids of varying weight.

Continuous elutriation was performed by inserting a hose connected to a potable water faucet into a 2 liter (L) graduated cylinder with a 5 inch inside diameter. The hose extended to the bottom of the cylinder. A volume of 800 milliliters (ml) of granular sludge was poured into the cylinder. Water flow was adjusted until the granules were fluidized just below the washout velocity of the granules. Small particles and fines were allowed to wash out. The sludge was elutriated for 30 minutes. Aliquots of the washed sludge were analyzed for total and volatile solids.

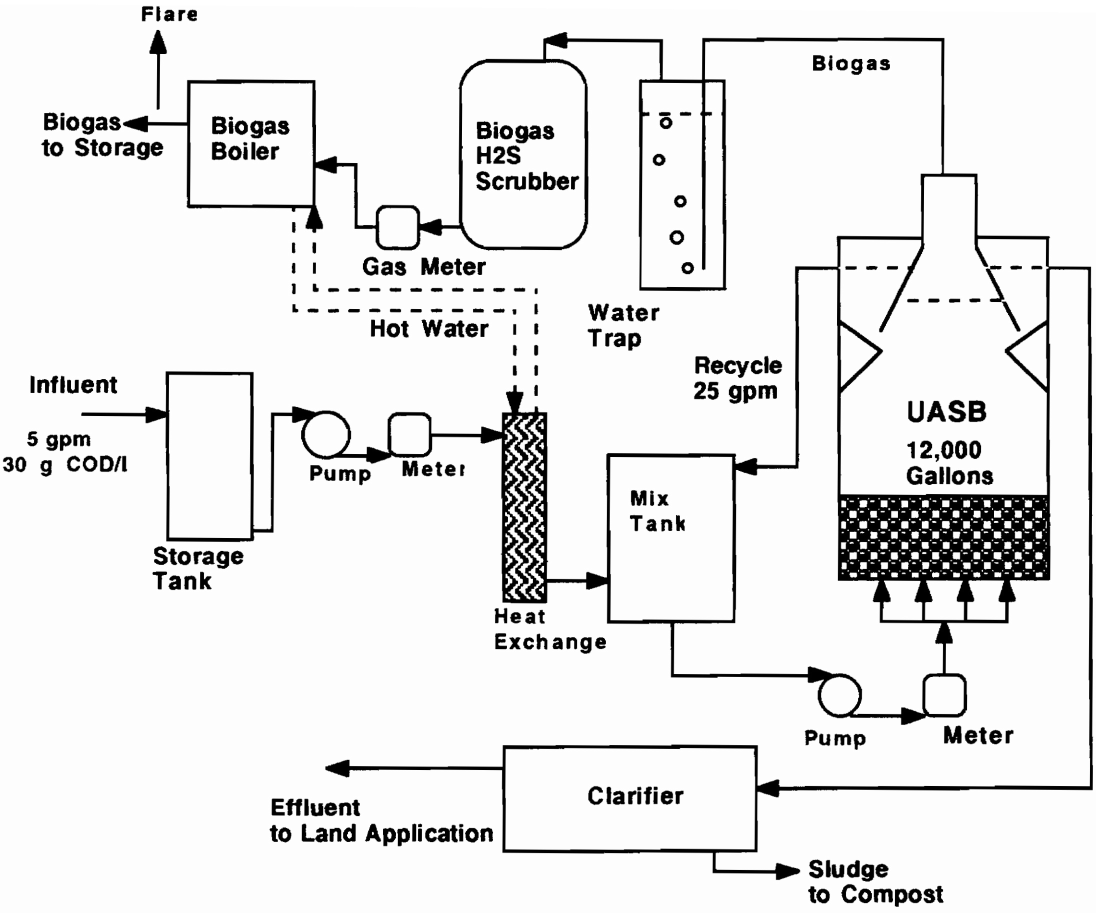

Figure 5. Full Scale UASB System Schematic for FAEC.

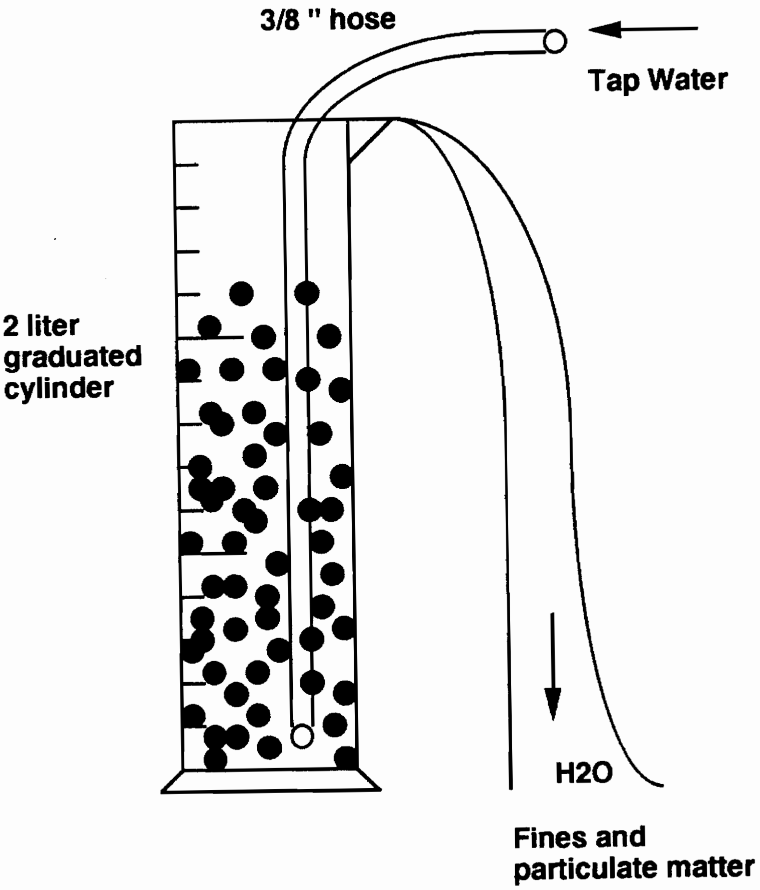

Figure 6. Schematic of the Apparatus for the Elutriation of Granular Sludge.

The settling velocity of the sludge was determined using the same apparatus. The sludge was allowed to fully fluidize and expand to the top of the cylinder without granules overflowing. The flow rate of the tap water was measured and the sludge settling velocity was assumed to be equal to the superficial upward water velocity. The superficial flow velocity of the water was calculated by measuring the cross sectional area of the cylinder in square centimeters and the water flow rate in cubic centimeters per minute.

The sludge was examined under the microscope and total suspended solids (TSS) and volatile suspended solids (VSS) tests were performed.

3.4 Chemical Analysis of Wastewater

All chemical analyses were performed in accordance with Standard Methods (23) unless otherwise noted.

Total solids, volatile solids and suspended solids were analyzed during periods of apparent steady states (when the organic load had been maintained at a constant level for at least 4 days and the effluent COD was within a 15 % margin for 5 days). The pH of the effluent was determined daily using a glass electrode. Alkalinity was determined by titration with 0.02 Normal (N) solution of H_2SO_4.

Total phosphate (TP) was determined by a persulfate digestion procedure. Ortho-phosphate (OP) determinations were performed by the ascorbic acid method using a Beckman Spec 20 at 678 nm. The TP and OP levels were determined at steady state conditions.

Total Kjeldahl Nitrogen (TKN) was determined by a two hour digestion in concentrated H_2SO_4 with a mercury catalyst, followed by distillation into boric acid and titration with 0.02 N H_2SO_4. Ammonia nitrogen (NH_3-N) was determined by distillation and titration with 0.02 N H_2SO_4.

COD was determined by dichromate reflux digestion for two hours at 150 °C. Titrations were performed with 0.05 N ferrous ammonium sulfate and ferroin indicator. BOD_5 was determined in 300 ml dilution bottles inoculated with 2 ml of primary municipal sludge per 5 liters of dilution water. Bottles were incubated at 20 °C for 5 days in a dark cabinet. A YSI dissolved oxygen (DO) meter was used to measure dissolved oxygen before and after incubation.

Sodium was determined by atomic absorption spectrophotometry on a Perkin-Elmer 703. Chloride was determined by ion chromatography (IC) on a Bausch and Lomb Dionex 3000 with a conductivity cell detector. The IC column used was a polystyrene/divinylbenzene stationary phase and a quaternary amine exchange function. The IC eluent was carbonate solution at pH 9.5. The suppressor used was H_2SO_4.

Volatile acids were determined by titration methods as described by van der Laan (61). The sample (50 ml) was filtered and then titrated with 0.1 N HCl to pH 3. The sample then was boiled for three minutes under a reflux condenser to remove CO_2. A final titration with 0.1 N NaOH was performed to pH 6.5.

The milliequivalents (mequ) per liter of total organic acids present, C, was calculated by the following equation:

$$C = \frac{(B*101)-Z-100}{99.23} * \frac{100}{V} \qquad [5]$$

where:

C= Total Organic Acids, milliequivalents
B = Volume of NaOH titrant, ml
Z= Volume of HCl titrant, ml
V= Volume of sample, ml

To convert C to mg/L acetate, it was multiplied by 60 mg/mequ.

Biogas quality was determined by GC using a flame ionization detector (FID) coupled to a molecular sieve 5A packed stainless steel column. Methane (99.0% pure) was injected in various volumes to develop a standard curve.

3.5 Incubation Chamber for Batch and Continuous Experiments

A forced air system was used to circulate air heated by five 100 watt light bulbs in a plenum above the batch reactors. The plenum extended behind the incubation area and the heated air was distributed through perforations in the back wall. Switches activated by a temperature probe automatically controlled the light bulbs and a circulation fan maintained the incubator at 32 ±1 °C.

3.6 Batch assays

In batch assays 500 mL serum flasks sealed with a 32 millimeter (mm) rubber septum and a plastic screw cap were used. The head space of each flask was initially purged with nitrogen gas for 30 seconds at a flow rate of approximately 1 cubic foot per minute. All batch tests were performed in the custom fabricated incubator at 32 °C. A stock synthetic feed solution of VOA was used which consisted of acetic acid, proprionic acid, and butyric acid at 30:20:20 g/kg, respectively. The stock solution contained 12.5 g COD/L after adjusted to pH 7.5 with 50% sodium hydroxide.

Each batch reactor vessel was connected to an individual Marriotte gas measuring system by Tygon® tubing and hypodermic needles. Figure 7 shows a schematic of the batch system design. The Marriotte flask system was used to measure methane without CO_2. The Marriotte system consisted of an inverted serum bottle filled with a 2 N sodium hydroxide solution containing alizarin yellow pH indicator. The biogas entered the flask by way of a needle inserted through the septum. The CO_2 was scrubbed from the biogas and the CH_4 displaced the caustic solution through another needle. Caustic solution was collected in narrow mouth containers fitted with a funnel to prevent evaporative losses. The collection flask was weighed at appropriate intervals, usually every 4 hours depending on the methane production rate, to obtain a cumulative methane production profile. The reactor flasks were shaken once a day.

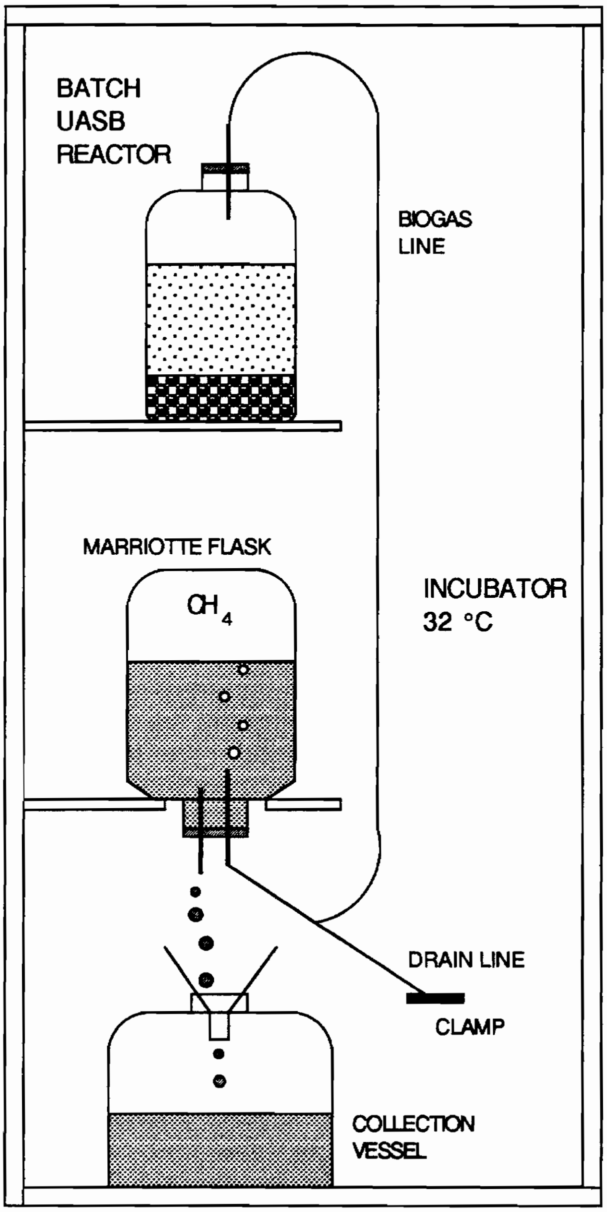

Figure 7. The Experimental Arrangement used in the Anaerobic Batch Reactor Studies.

Granular sludge inoculum was weighed after the free water was allowed to drain by gravity from the sludge through a net for approximately 15 seconds. Anaerobic conditions were not maintained during this period of time.

A nutrient solution containing macro and micro nutrients was made according to the formula in Table 3. The nutrient solution composition was adapted from Speece and McCarty (62). Fifty ml of concentrated nutrient solution were used per liter of wastewater. Chemicals were obtained from Fisher Scientific, Corp.

The weight of displaced hydroxide solution from the Marriotte flask was used to calculate methane gas production. The COD value of the measured volume of methane was calculated from the collection flask weight data using the following formula:

$$CH_4\text{-COD} = \frac{A(W - W_c)}{d} \qquad [6]$$

where,

CH_4-COD = the COD of the wastewater equivalent to the methane produced, mg COD,
A = the theoretical COD value of methane at 32 °C (2.574 g COD/L methane from Equation 2),
W = weight of the displaced liquid from the Marriotte flask, grams,
W_C = weight of liquid displaced from control, grams,
d = density of caustic solution at 32 °C, mg/L.

The cumulative methane production data obtained from each batch experiment was averaged from three replicates. Control values were subtracted from the cumulative values, and it was assumed that any evaporative losses were accounted for in the controls. All batch reactors were inoculated with 5 grams per 500 ml flask of total (gravity drained in a net)

Table 3. Nutrient Solution Composition used in Batch Studies

Nutrients	Concentration in Flask
Macro Nutrients	(g/L)
NH_4Cl Ammonium Chloride	0.28
K_2HPO_4 Potassium Phosphate, dibasic	0.25
$MgSO_4$-$7H_2O$ Magnesium Sulfate, heptahydrate	0.10
$NaHCO_3$ Sodium Bicarbonate	0.40
$CaCl_2$-$2H_2O$ Calcium Chloride, dihydrate	0.10
Trace Elements	(mg/L)
$FeCl_2$-$4H_2O$ Iron Chloride	2.00
H_3Bo_3 Boric Acid	0.05
$ZnCl_2$ Zinc Chloride	0.05
$MnCl_2$-$4H_2O$ Manganese Chloride	0.50
$CuCl_2$-$2H_2O$ Copper Chloride	0.03
$(NH_4)_6Mo_7$-$4H_2O$ Ammonium Molybdate	0.05
$AlCl_3$.$6H_2O$ Aluminum Chloride	0.09
$CoCl_2$-$6H_2O$ Cobaltous Chloride	0.20
$NiCl_2$-$6H_2O$ Nickle Chloride	0.05
$NaSeO_3$-$5H_2O$	0.10
36% HCl	0.001 (ml)

biomass. The resulting volatile solids concentration of the inoculum was 0.475 grams VSS/0.5 L (0.95 g VSS/L). All flasks received 0.25 grams of sodium bicarbonate to help prevent pH fluctuations. The amount of sodium bicarbonate added resulted in the addition of 500 mg/L of $NaHCO_3$ or 137 mg/L of Na^+.

The clam-processing wastewater samples containing the higher TCOD values were used in the batch experiments to ensure that the batch test would not be substrate limited for at least the first 24 hours of the experiments. The clam wastewater was removed from refrigeration and heated to 32 °C prior to addition to the batch reactors. Sludge blanks receiving no substrate contained distilled water, granular sludge and 0.25 grams of sodium bicarbonate.

3.6.1 Specific Methanogenic Activity (SMA)

The SMA measures the methanogenic metabolic activity of a sludge in terms of the rate at which the granular sludge will convert the COD in the wastewater to methane gas. To determine the metabolic activity of the sludge, 5.0 grams of drained granular sludge (equivalent to 0.5 grams of volatile sludge solids) were transferred to each experimental flask. Clam wastewater (500 ml) was added to each of three replicate flasks. A 4:1 ratio of wastewater COD to the VSS concentration of the sludge was selected in order to ensure that the test was not substrate limited. Since the average activity of granular sludge is 1 g COD/g VSS/d, an initial test concentration 4 times higher will prevent the batch test from being substrate limited for the first several days allowing for the determination of maximum substrate utilization rates (53).

3.6.2 Biological Methane Potential (BMP)

The biological methane potential (BMP) of the wastewater is the maximum amount of methane gas that can be produced from one gram of COD of the wastewater. The BMP is similar to the aerobic BOD_5 test which measures the maximum oxygen demand in an aerobic environment. Biological methane potential was used to determine the anaerobic biodegradability of the clam wastewater. Thus, a "BOD_5" or "$BOD_{ultimate}$" can be established on an anaerobic basis. Granular sludge was provided in excess so that the test could be performed under substrate limiting conditions. The test was allowed to proceed until methane production ceased.

3.6.3 Nutrient Deficiency Test

The purpose of this investigation was a screening for nutrient deficiency in the wastewater. Nutrient amended wastewater was compared to unamended wastewater for methane production rates and total gas production. The nutrient solution added was a mixture of inorganic chemical nutrients (see Table 3).

An Initial sludge inoculum of 0.5 g VSS per 500 ml wastewater was used. The sludge was retained in each batch flask between feedings by decanting the supernatant after the methane production leveled off. The sludge was rinsed with distilled water and again decanted. All attempts were made to prevent whole granules from escaping, but fines were allowed to exit with the decanted fluids. Sludge blanks with no clam wastewater and sludge blanks with only nutrient solution were included in the study as controls.

3.6.4 Anaerobic Toxicity Assay (ATA)

A series of wastewater test flasks spiked with increasing amounts of a suspected toxin, NaCl, were monitored to determine cumulative methane production and the rates of methane production. Various concentrations of NaCl dried at 120 °C for two hours were weighed and added to the clam-processing wastewater in the batch experiment. The metabolic activity and total gas production were determined.

A series of wastewater batch reactors was amended with concentrations of NaCl from 4200 mg/l to 12,600 mg/l as Na^+. The cumulative methane production and the metabolic activity with respect to unamended controls were compared.

3.7 Continuous-Flow UASB Experiment

Figure 8 depicts the continuous-flow, UASB experimental design. A one liter, laboratory-scale, UASB reactor was fed clam wastewater at increasing loading rates from 1 g COD/L/day to 17 g COD/L/day. The reactor was constructed of a 0.25 inch thick Plexiglass cylinder with an inside diameter of 3 inches and height of 12 inches. The three phase separation of sludge, liquid and gas was accomplished by use of a funnel. Wastewater flowed through a 7/8 inch PVC "riser" tube with inlet holes below the liquid surface. The riser tube was attached to the bottom of the funnel. The funnel created a quiescent zone to allow settling of sludge back into the reactor. A baffle, attached to the bottom of the riser tube, helped deflect rising granules from entrainment in the

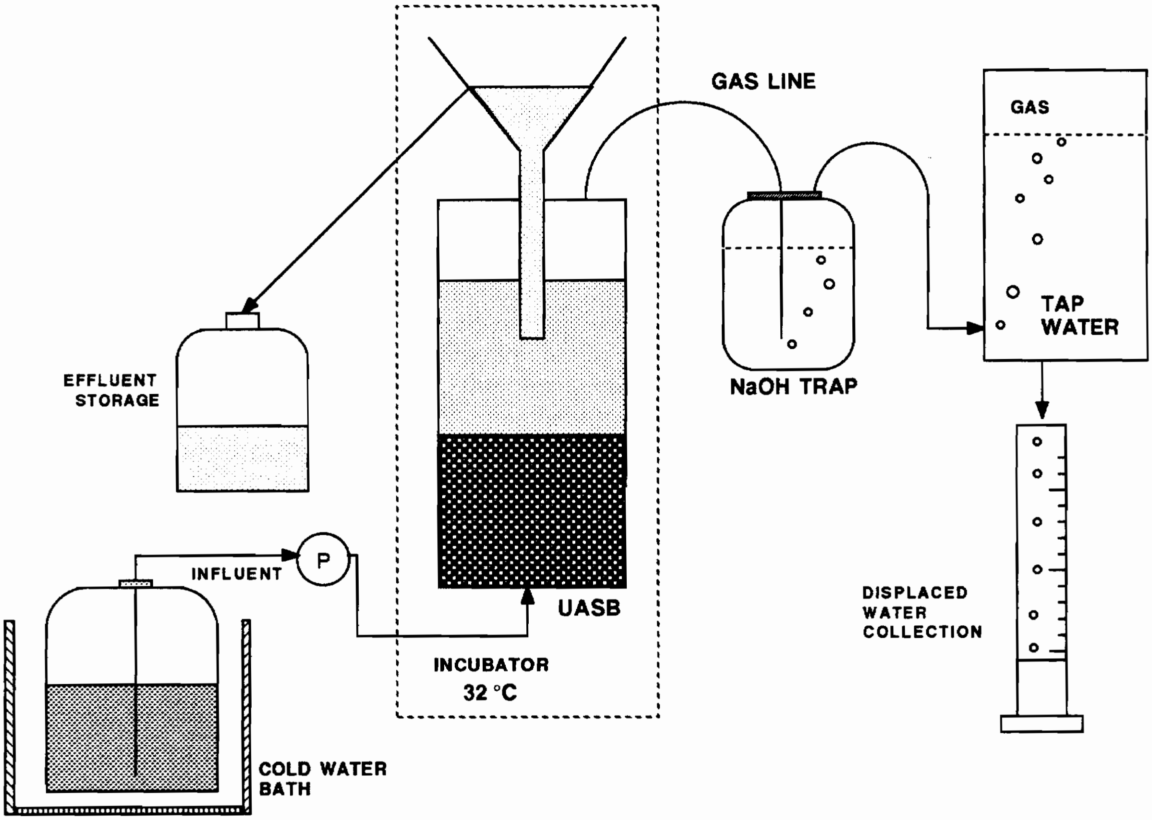

Figure 8. Schematic of the Continuous UASB System Laboratory Design.

liquid. The funnel settling area had a volume of 0.435 L. The effluent flowed into a 2 L clarifier, and then exited to waste. If needed, recycle was taken from the top of the clarifier, 2 inches below the liquid surface.

Gas collected in the head space of the reactor and flowed (by pressure from the evolving biogas) through 1/4 inch Tygon® tubing into a biogas scrubber (modified Marriotte flask). CO_2 and H_2S were removed by a 2 N sodium hydroxide solution containing alizarin yellow pH indicator. The alizarin yellow changes from red (above pH 10) to yellow (below pH 10). The caustic solution was replaced when the indicator turned yellow. The scrubber also served to provide a slight back pressure to keep the gas head space in the UASB reactor at a desired level in order to ensure that the liquid volume in the reactor was 1 liter. The biogas bubbled through the hydroxide solution and CO_2 was absorbed in the caustic (thus dropping the pH). The methane (CO_2 free) then entered the water displacement device. The water displacement device had a height of 55 cm and diameter of 20 cm. The volume of the container was 17.3 liters. Tap water displaced by methane dripped into a calibrated vessel, and the volume of water displaced was recorded to the nearest 10 ml.

Influent wastewater was kept at 10 °C in a water bath to help prevent pre-acidification and degradation. Nutrients were not added to the influent of the continuous-flow reactor.

Biogas quality was determined by gas chromatography using a flame ionization detector. Samples of digester gas were taken through a septum 12 inches from the reactor gas outlet in the Tygon® tubing that led to the CO_2 scrubber. Gas samples were injected at known volumes. The mass of the CH_4 was determined by comparing peak areas to the standard curve developed and the percent CH_4 was calculated. The remainder of the biogas was assumed to be CO_2. The methane production data obtained at ambient laboratory temperatures was converted to volumes at 32 °C for comparison to COD removal data in the 32 °C UASB reactor.

The reactor was inoculated with 300 grams of gravity drained granular sludge which reached a height of 4 inches inside the laboratory reactor, equivalent to a biomass volume of 9 cubic inches. Sludge growth was estimated by measuring the volume and weight increase of the sludge at the end of the experiment. Calculations based on literature growth yield values were also performed. Calculations for the volume of a full scale UASB were performed. A reference table for nomenclature is presented in Appendix D.

4.0 RESULTS AND DISCUSSION

4.1 Clam-Processing Wastewater Characteristics

Wastewater samples collected from the clam-processing plant, at approximately monthly intervals, exhibited wide variations in COD strength and other parameters such as Cl^- and TS. Wastewater flow through the continuous-flow reactor was adjusted accordingly to maintain a steady organic loading rate. Thus, the HRT was varied to obtain the desired organic loading.

Averages and ranges of wastewater influent characteristics are provided in Table 4. The values shown in Table 4 include all samples collected from the clam-processing facility. However, some samples were not used in the laboratory experiments due to high salt content for example. The wide variations in the wastewater COD made it difficult to maintain steady loadings on the laboratory UASB. Steady loading rates are needed for a laboratory study but would not be as necessary in a full scale reactor. COD and BOD_5 levels varied with each collection of wastewater from the clam-processing facility. The COD declined from a high value of 6,000 mg/L to 1,000 mg/L and less, as determined in this investigation. Variations in salt concentration were also observed through the course of this study. Wastewater Na^+ levels ranged from 2,800 mg/L to 10,500 mg/L. Samples containing Na^+ above 5,000 mg/L were not used to feed the laboratory UASB reactor. Chloride ranged from 4,900 to 15,800 mg/L. Total solids ranged from 9.5 to 24.5 g TS/L, with 80-99% being dissolved.

Table 4: Range and Average of Clam-Processing Wastewater Characteristics

Analysis	Range (mg/L)	Average (mg/L)
Total 5 day BOD, $TBOD_5$	350-3,300	1,100
Soluble 5 day BOD, $SBOD_5$	250-3,050	800
Total COD, TCOD	500-6,000	1,700
Soluble COD, SCOD	450-3,650	1,250
Total Solids, TS	9,500-24,500	16,500
Total Suspended Solids, TSS	140-1,140	700
Total Kjeldahl Nitrogen, TKN	100-290	220
Ammonia Nitrogen, NH_3-N	10-200	62
Total Phosphorus, TP	22-64	40
Ortho-Phosphate, OP	21-61	34
Alkalinity as $CaCO_3$	215-650	430
pH (unitless)	6.4-8.6	7.2
Sodium, Na^+	2,800-10,500	4,200
Chloride, Cl^-	4,900-15,800	8,500

4.2 Granular Sludge Characteristics

The settling velocity of the granular sludge inoculum, as determined by the elutriation method, was found to be 15 m/s. This value is within range of the reported values in the literature for granular sludge. The sludge from the laboratory UASB retained the same settling velocity at the conclusion of the experiment. The sludge volume index (SVI) of the inoculm was approximately 30 ml/g after only 1 minute compared to the regular 30 minute test required for aerobic sludge.

The volatile solids level was 9.5% of the drained wet granular sludge and 81% of the total dried solids. The volatile solids are an indication of the active mass of the total drained sludge with the normal range for granular sludge between 7.5 and 14.5% (10). The VS content (81% of the dry solids) and ash content (19% of the dry solids) in the granules remained constant during the course of the continuous laboratory experiment.

The size of the granules varied from about 0.8 to 2.5 mm in size. Typical sludge granules range from 0.4 to 3 mm (63). The sludge retained its glossy black appearance, with no gray colorations, indicating a relatively small population of acidifiers. Whereas, a gray sludge contains a high percentage of acidifiers typical of systems treating unacidified carbonaceous wastewater (64). At the end of the continuous-flow experiment the sludge had a more spherical look, rather than the original, oblong shape. This may have been caused by the relatively low HRT in the laboratory reactor, as compared to the high HRT in the FAEC reactor (greater than 6 hours (65)).

The activity of the full-scale UASB granular sludge reported by FAEC personnel (65) was slightly higher (0.5 g CH_4-COD/g VSS/d) than the average sludge activity measured during the continuous-flow experiment (the sludge activity calculated from the continuous-flow reactor averaged 0.4 g CH_4-COD/g VSS/day). Granular sludge has an activity range of 0.4 to 2.0 g COD/g VSS/day (66). In comparison, the sludge activity calculated from batch experiments using the clam wastewater was 0.9 g CH_4-COD/g VSS/day.

Initial microscopic examination of the sludge showed mostly rod-shaped bacteria assumed to be *Methanothrix soehngenii* (67). Observations of the final sludge showed normal growth of fines and some flocculent biomass. Under extended upflow operations these fines would either wash out or produce new granules.

4.3 Batch Reactor Assays

The data for batch experiments is presented in Appendix A.

4.3.1 Nutrient Studies

Results obtained after the addition of concentrated inorganic nutrient media to the clam-processing waste were compared to those obtained with the unamended waste. Three feedings of nutrients, using the same biomass in each flask, were used to determine indications of nutrient deficiency in the clam-processing wastewater. The data are presented in Figure 9. All three feedings (using the same COD concentrations) showed no significant differences between clam wastewater amended with nutrients and clam wastewater not amended with nutrients. The nutrient recipe (Table 3), obtained from the literature, was formulated to provide nutrients in excess but well below any toxic concentrations. The increase in sludge activity and cumulative methane production during subsequent feedings is attributed to acclimation of the granular sludge to the clam-processing wastewater.

The average total cumulative gas produced after 185 hours was 520 ml for the nutrient amended flasks and 515 ml for the unamended control flasks after three feedings of wastewater. All flasks during the third feeding had the same rate of methane production, 1.1 g CH_4-COD/g VSS/d. Thus, no significant increase in methane production from the addition of nutrients was observed. From the data obtained, the need for adding nutrients was not apparent.

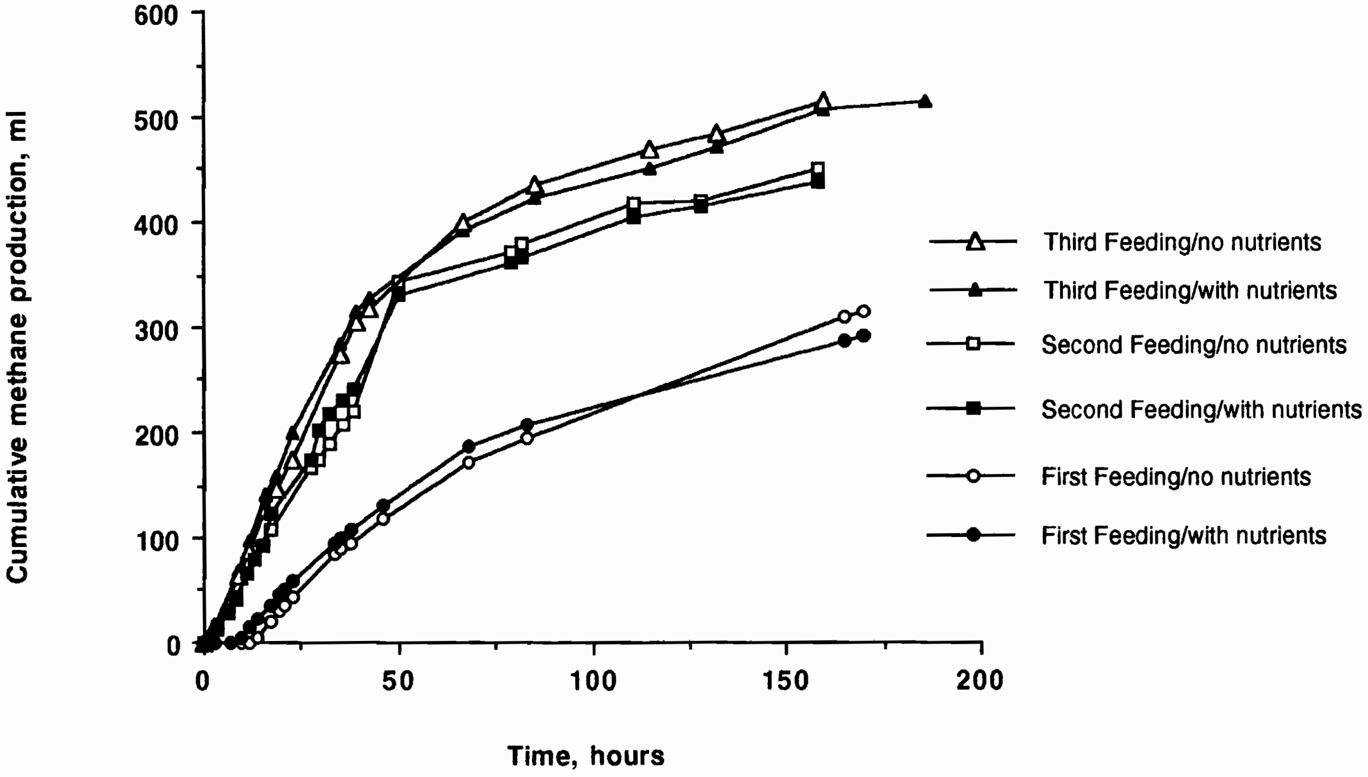

Figure 9. Plot of the Cumulative Methane Production versus Time for the Addition of Nutrients to Clam-Processing Wastewater in Batch Reactors for Three Serial Feedings.

4.3.2. Biological Methane Potential (BMP)

The BMP is an anaerobic test analogous to the aerobic 5 day biological oxygen demand (BOD_5) test for measuring the organic strength of wastewaters. The data from the BMP batch experiment using clam-processing wastewater are presented in Figure 10. The third feeding data plot was used to determine the ultimate potential for conversion of organics to methane (10). The cumulative methane production leveled off at 500 ml CH_4. The cumulative methane production (500 ml CH_4 from 0.5 liter of waste) was used to calculate the anaerobic biodegradability of the clam-processing waste. The "anaerobic BOD_5" was calculated as 2.8 g CH_4-COD/L and the BMP was calculated to be 0.8 g CH_4-COD per g COD in the clam-processing wastewater.

Also, a BMP was calculated for the nutrient amendment batch tests (see Figure 9). The third feeding of nutrients obtained a cumulative methane production of 520 ml from an initial TCOD of 3,200 mg/L in the clam wastewater. The calculated BMP was also 0.8 g COD-CH_4/g COD in the wastewater.

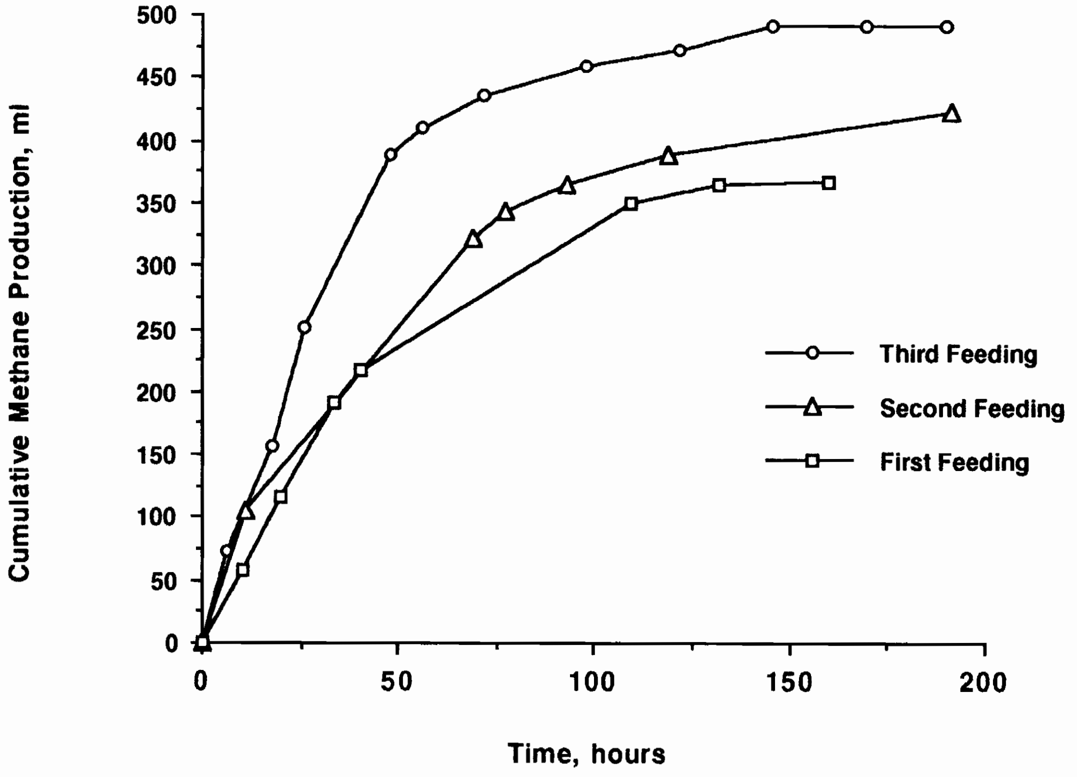

Figure 10. Cumulative Methane Production versus Time for Monitoring the Anaerobic Biodegradability of Clam-Processing Wastewater.

4.3.3 Specific Methanogenic Activity (SMA)

SMA calculations were based on the linear portions (first 30 hours) of the BMP and nutrient tests (third feedings), as shown in Figures 9 and 10. This region was considered representative of the maximum sludge activity rate utilizing the clam-processing wastewater.

The activity rate for the BMP test was 1.24 g CH_4-COD/g VSS/day. The activity calculated from the nutrient test was 0.9 g CH_4-COD/g VSS/day. The average of the two test yields an activity of 1.07 g CH_4-COD/g VSS/day.

The SMA using the VOA solution was less than the SMA using clam-processing wastewater. The relative sludge inactivity using VOA was most likely due to toxicity from the NaOH used to neutralize the volatile acids in the stock solution (~40 g/kg). Thus, the data using VOA were not used for comparative analysis.

4.3.4 Anaerobic Toxicity Assay using Sodium Chloride in Batch Reactors

Figure 11 shows the plots of data obtained from the NaCl toxicity experiment. Methane production rates were significantly depressed by increasing the salt concentration over a three fold rang. This indicates inhibition of methanogenesis by the addition of the NaCl above 5,250 mg/L as Na^+.

The sludge activity rates were 0.4, 0.24, 0.18, 0.06 and 0.04 g CH_4-COD/g VSS/day for salt concentrations of 4,200 (unamended), 5250, 6300, 8400 and 12600 mg/L Na^+, respectively. Thus a 10 fold decrease in activity was observed for a three fold increase in Na^+ concentration.

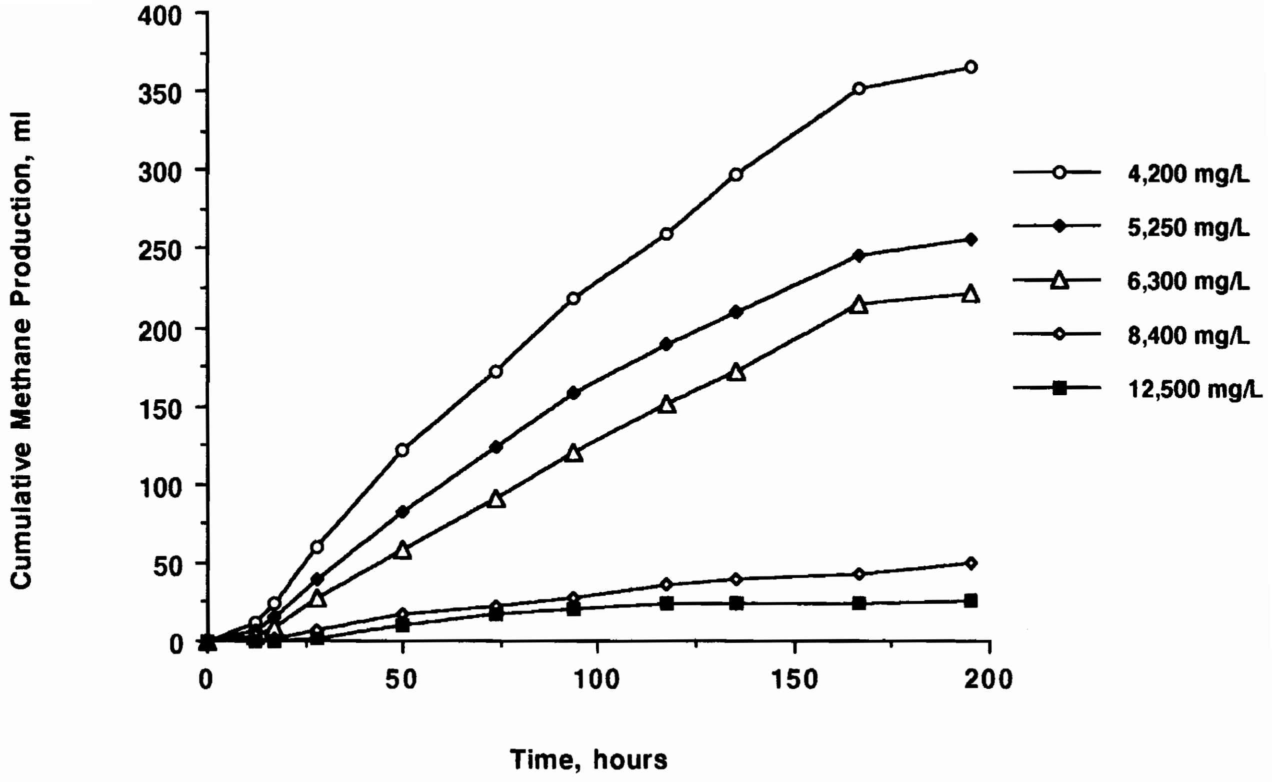

Figure 11. Plot of Variations in the Cumulative Gas Production versus Time as a function of Sodium Concentration.

The flasks containing salt did not reach the same level of ultimate cumulative methane production as the unamended flasks. The maximum theoretical methane production was not achieved in the flasks amended with NaCl. The total cumulative methane production was reduced 22, 45, 53, 90 and 95% from the theoretical for salt concentrations of 4200, 5250, 6300, 8400 and 12600 mg/L Na^+, respectively.

4.4 Continuous-Flow UASB Experimental Results

The results presented in this section cover a period of more than 7 months of continuous operation of the laboratory UASB from September 1989 to March 1990. Unless otherwise noted, the data presented in graph and tabular form were collected from day 100 to day 223 (end of experiment) for the continuous-flow reactor. The data is presented in Appendix B. The data collected before day 100 from the UASB which operated at organic loading rates less than 4.8 g COD/L/d were omitted because of the low loading rates and wide variations due to equipment trouble. After modifications to handle higher flows, reactor performance settled down to predictable and controllable loading rates on about day 100.

Effluent VOA analysis was difficult, thus, effluent COD data was used to monitor the reactor performance. Acetic acid peaks from ion chromatography were masked by carbonate peaks. The FID detector was not able to give reproducible results at the low range of concentrations of interest (less than 100 mg/L). The titration method for total VOA was abandoned for the easier

COD analysis. The total organic acid concentration was estimated from the COD values, assuming that one gram of organic acids contained 1.25 g COD.

The laboratory UASB reactor was inoculated with 300 grams of gravity drained granular sludge and resulted in an organic solids concentration of 28.5 g VSS/L in the reactor.

4.4.1 Organic Loading Rate and Effluent Quality

The organic loading rate (OLR) can be calculated on a reactor volume basis (g COD removed per liter reactor volume per day) or on a sludge volatile solids basis (g COD removed per g VSS per day). More often, the activity is reported per reactor volume since the sludge VSS concentration is changing through growth or washout.

The OLRs during the UASB experiment are summarized in Table 5. The averaged organic loading rates investigated were 4.8, 7.8, 13.8, and 16.3 g COD/L/d. These values are average values due to the fluctuations in the influent COD and the difficulty in adjusting the influent pump to match the rate needed to apply the specific OLR. The flow rate was adjusted to obtain the desired OLR and, thus, the HRT varied. Figure 12 provides the organic loading rates from day 100 of the experiment to day 223 and the methane production rates calculated as CH_4-COD/L/d.

Table 5. Organic Loading Rate (OLR) Schedule for the Laboratory UASB

Time (days)	Average OLR (g COD/g VSS/day)	Average OLR (g COD/L/day)	Average HRT (hours)
100-120	0.2	4.8	7.7
130-145	0.3	7.8	4.9
160-175	0.5	13.8	3.0
200-223	0.6	16.3	3.2

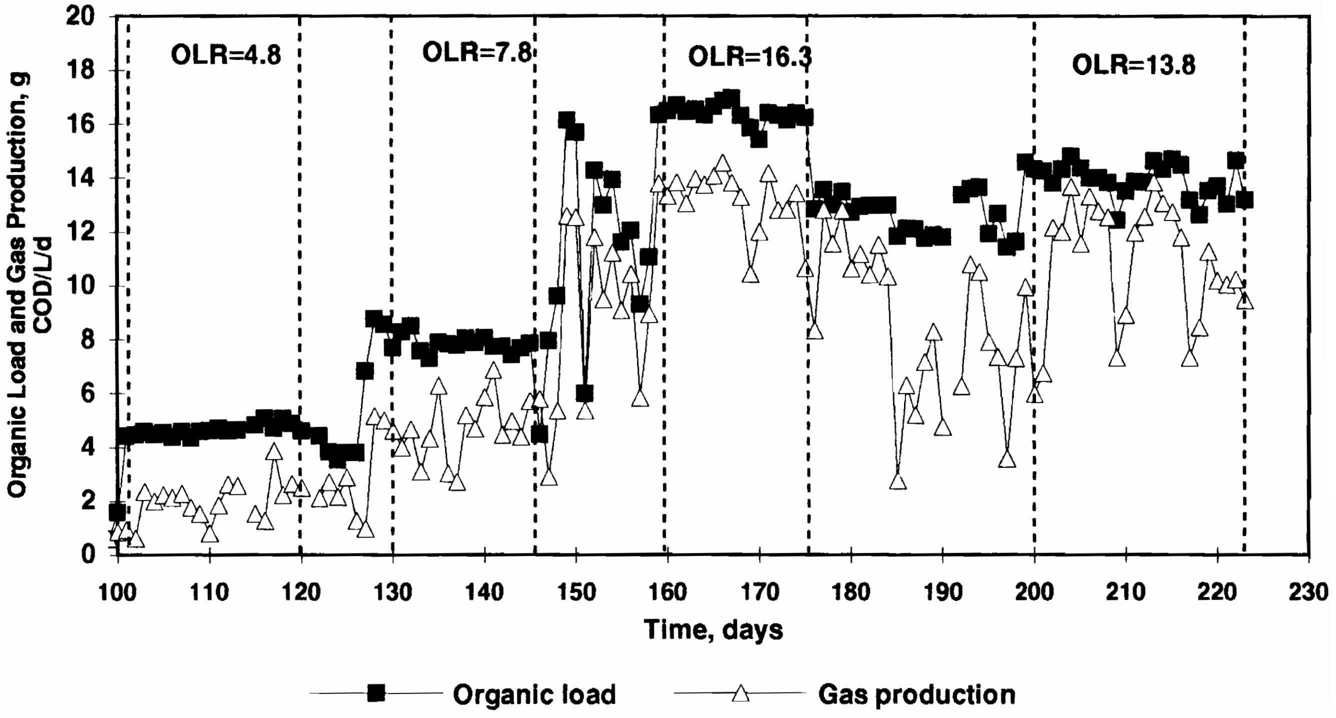

Figure 12. Continuous UASB experimental Organic Loading and Daily Gas Production versus Time.

The loading rate was maintained until steady state conditions were reached, as defined by a constant effluent COD value. Figure 13 shows the daily effluent TCOD and SCOD. Usually 5 days were needed to produce a steady effluent COD after the OLR was changed. Table 6 shows the average effluent quality for each loading rate. The maximum OLR studied that resulted in satisfactory effluent quality within the proposed limits of the VSWCB was OLR=13.8. The effluent quality at OLR=13.8 was for some parameters better than the effluent quality at an OLR=7.8. One possible explanation is that the granular sludge had adapted better to the clam-processing wastewater.

Reduction in COD strength in the wastewater was accompanied by a reduction in nutrient levels. Generally there is less reduction in nutrients in anaerobic processes compared to aerobic systems (68). TKN was not reduced significantly. Ammonia increased in the reducing environment from an average of 62 ppm in the influent samples to an average of 173 ppm in the effluent samples. The reduction of total nitrogen to ammonia nitrogen averaged 50%. The effluent ammonia was typical of reducing environments and could be removed from the effluent by biological nitrification. Aerobic treatment systems also result in effluent ammonia which would also require tertiary treatment. The USEPA requires ammonia (NH_4-N) concentrations in receiving waters to be below 0.02 mg/L to prevent interference with oxygen transport in fish (22). TP and OP were reduced on the average in the wastewater by 22% and 15%, respectively.

The amount of sludge washout due to the low HRT, and consequently high superficial flow velocity, may have contributed to TSS levels in the effluent.

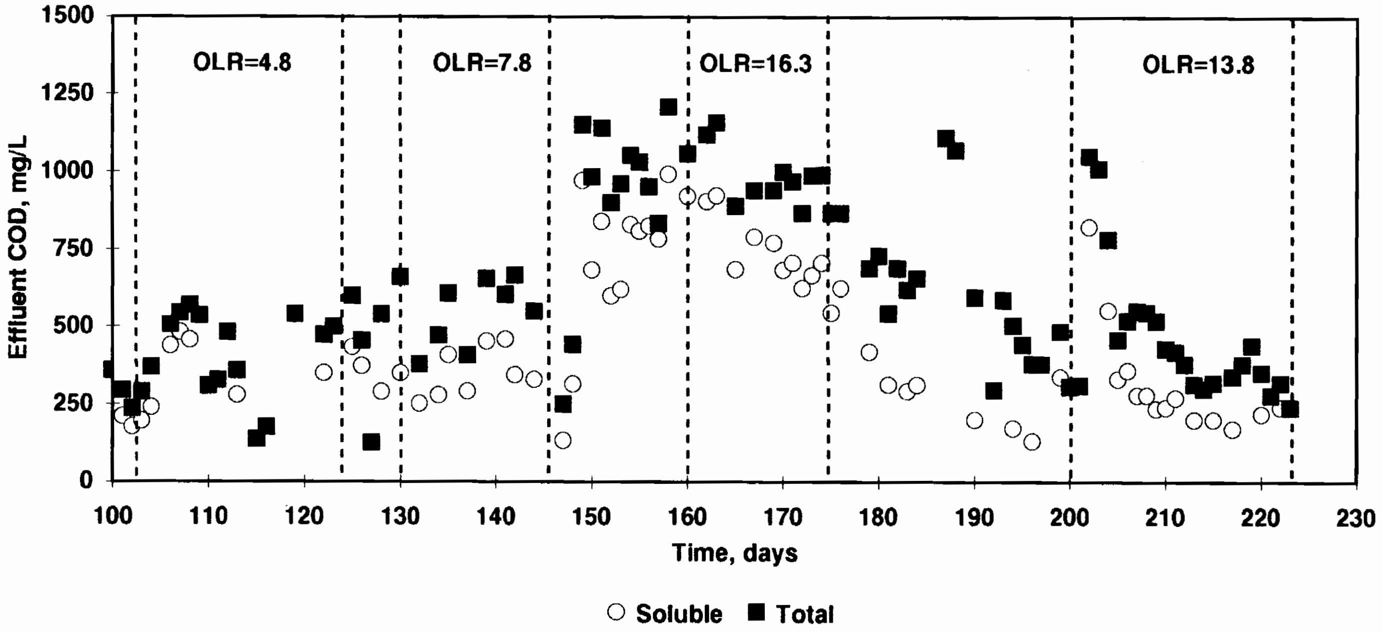

Figure 13. Continuous UASB reactor Effluent COD versus Time.

Table 6. Comparison of the Average Effluent Quality during Steady State Organic Loadings Rates.

Parameter (mg/L)	Organic Loading Rate (g COD/L/d) 4.8	7.8	13.8	16.3
$TBOD_5$	77	200	200	624
$SBOD_5$	67	133	126	508
TCOD	408	543	416	939
SCOD	352	353	253	686
TSS	66	92	86	543
TKN	220	170	230	300
NH_3-N	110	140	165	275
TP	18	23	51	33
OP	17	23	46	30
Alkalinity (as $CaCO_3$)	655	1,280	1,225	1,380
pH	7.3	7.2	7.1	7.1

Washout of fine particles is, however, desired to keep the sludge bed in the reactor free of inactive solids. Average effluent TSS values for each loading rate investigated are shown in Table 6. TSS values were above the maximum proposed effluent requirements (500 mg/L) during the OLR=16.3 period. However, at an OLR=13.8 and a similar HRT, the TSS average was well under the highest proposed VSWCB limit. TSS values were below the 90 mg/L limit for OLR equal to 4.8 g COD/L/d.

BOD_5 concentrations consistently remained below the VSWCB limit of 500 mg/L for organic loadings at and below 13.8 mg COD/L/d.

4.4.2 Treatment Efficiency for COD, BOD_5, and TSS removal

The effluent COD and per cent COD reduction are shown in Figures 13 and 14, respectively. Table 6 shows the average effluent COD and BOD_5 for each OLR. The reactor average efficiency parameters for each OLR studied are summarized in Table 7. The OLR equal to 13.8 g COD/L/d gave the best results for effluent quality (while maximizing the OLR) and resulted in the TCOD averaging 416 mg/L and the SCOD averaging 253 mg/L. $TBOD_5$ averaged 200 mg/L remaining well below 500 mg/L, while $SBOD_5$ averaged 125 mg/L.

Figure 14 shows the percent treatment efficiency for TCOD and SCOD removal. The large fluctuations in the TCOD removal efficiency were caused by effluent suspended solids. SCOD removal efficiency averaged 64, 70, 83 and 61% for OLR's of 4.8, 7.8, 13.8, and 16.3 g COD/L/d, respectively. TCOD removal efficiency averaged 73, 64, 77 and 56% for OLR's equal to 4.8, 7.8, 13.8, and 16.3 g COD/L/d, respectively.

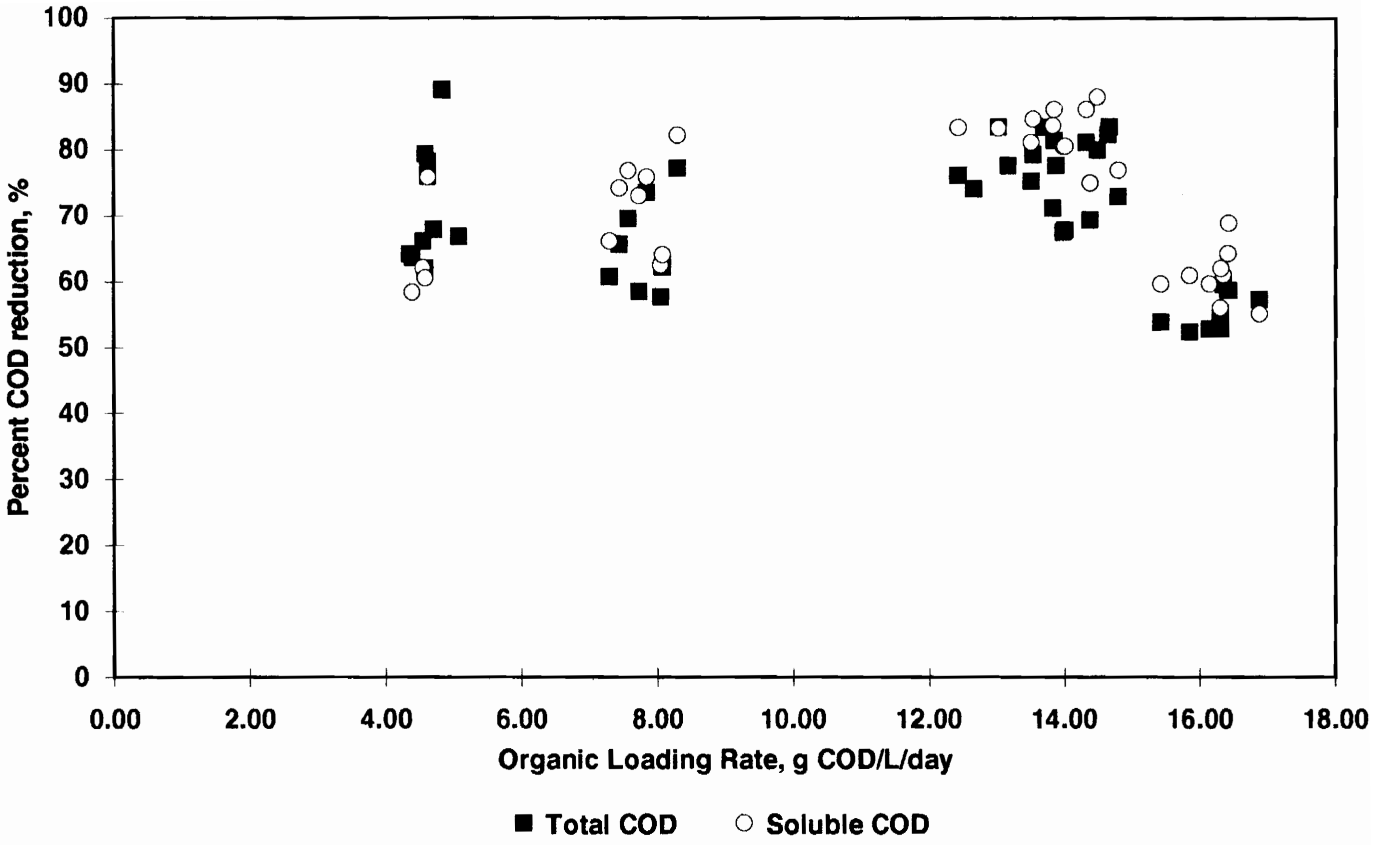

Figure 14. SCOD and TCOD % Reduction in the Continuous Reactor versus Organic Loading Rate during Steady State Periods.

Table 7. Comparison of Average Removal Efficiency Parameters for the Organic Loading Rates Studied.

	Organic Loading Rate (g COD/L/d)			
Parameter	4.8	7.8	13.8	16.3
$TBOD_5$, %	93	82	82	67
$SBOD_5$, %	93	87	87	65
TCOD, %	73	64	77	56
SCOD, %	69	70	83	61
TSS, %	92	69	83	31
Organics removed as CH_4-COD, %	46	63	81	79

The average effluent TSS data for each loading rate are shown in Table 6. The TSS data is plotted in Figure 15. TSS removal efficiency averaged 83% during the period for an OLR of 13.8 g COD/L/d, and averaged only 31% during the OLR of 16.3 g COD/L/d. TSS removal efficiency shown in Table 7 averaged 69% for an OLR of 7.8 g COD/L/d and 92% for an OLR of 4.8 g COD/L/d. An abrupt release of occluded biogas observed within the sludge bed may have influenced effluent TSS results by entraining solids through the settling area. Because the small laboratory reactor had only a few inches of hydraulic head space above the height of the sludge bed, the ability of the laboratory settler device to prevent TSS from entrainment in the effluent may have been hampered at such a small scale.

4.4.3 Gas Production

Figure 12 provides data for gas production as CH_4-COD during the last 123 days of the continuous experiment. During the period of OLR = 13.8, the average ratio of CH_4-COD production rate to organic loading rate was 81%. The ratio during the period when OLR = 16.3 was 79% (see Table 6). Thus, the methane production efficiency was better at OLR = 13.8 than at an OLR = 16.3. Pump failures interrupted the feeding period between days 180 and 200 and resulted in erratic OLRs and methane production.

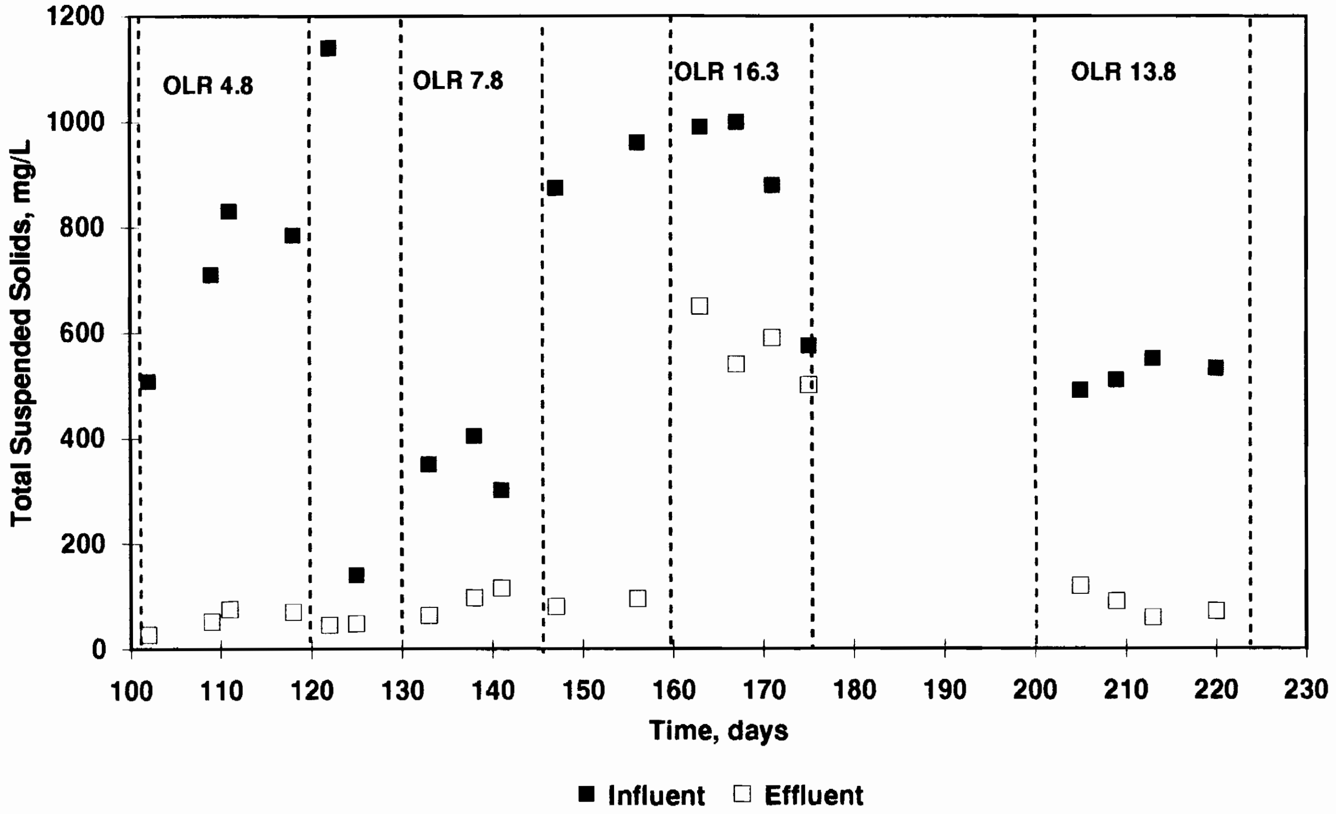

Figure 15. Total Suspended Solids versus Time in the Continuous UASB Reactor.

The quality of the biogas ranged from 70% to 80 % methane. The calibration curve from the gas chromatography analysis is shown in Figure 16. The calibration data is presented in Appendix C. The correlation for the standard curve was very high having an R value of 1.0. The remaining portion of the biogas was assumed to be mostly CO_2. Hydrogen, H_2S and water vapor was assumed to contribute in combination less than 1% of the biogas volume (22).

Calculated average specific activity based on gas production in the continuous reactor during OLR=13.8 resulted in 0.4 g CH_4-COD/g VSS/day (data in Appendix B). The average volumetric activity was 11.2 g CH_4-COD/L/d.

Thus, the estimated gas production for a full scale UASB reactor based on 1000 gallons of clam-processing wastewater is as follows: At an OLR of 13.8 g COD/L/day the average conversion efficiency of wastewater organics to CH_4-COD was 81% (Appendix B). The average COD removal was 1,300 mg COD removed per liter of wastewater. The methane content of the biogas was 80% during this period. Thus, the amount of methane produced from 1000 gallons of wastewater would be approximately 700 ft^3 of biogas (80% methane and 20% CO_2 at STP).

The BTU content of methane is 960 BTU/ft^3 STP. At a cost of $0.50 per 100,000 BTU for industrial natural gas (69), the biogas BTU content of the clam-processing wastewater investigated was potentially worth $2.50/1000 gallons treated. Thus, 1000 gallons of the clam-processing wastewater containing 1000 mg/L of COD metabolized to methane yields 560 ft^3 of methane or 518,400 BTU per 1000 gallons of wastewater treated anaerobically by the UASB process.

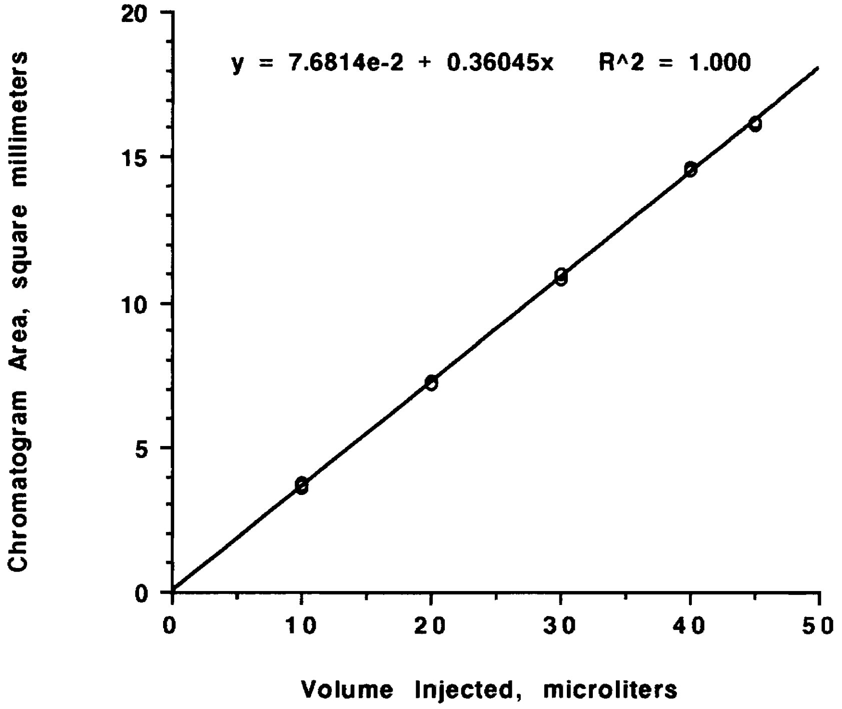

Figure 16. Gas Chromatography Calibration Curve for Methane Composition in Biogas.

4.4.4 HRT and Superficial Flow Velocity

The hydraulic retention time on a daily basis is shown in Figure 17. The average HRT for each organic loading rate is shown in Table 5. The HRT dropped from as high as 11 hours to 3 hours over the course of the experiment. Generally, in order to increase the organic loading to the reactor, the HRT had to be lowered.

The reactor operated well at the low HRT. Sludge wash out was not a problem at the applied superficial velocities (SFVs of 0.3, 0.4, 0.7 and 0.6 m/hr for OLRs of 4.8, 7.8, 13.8 and 16.3 respectively). The 7/8 inch inside diameter of the riser representing the narrowest constriction through which the wastewater flowed resulted in an SFV of 0.65 m/hr at the lowest applied HRT of 3 hours. The TSS in the effluent averaged 86 mg/L during the period when OLR equaled 13.8 g COD/L/d and the HRT equaled 3.2 hours.

4.5 Estimated Sludge Production

The estimated sludge yield for the laboratory UASB was 0.004 g VSS/g COD removed. The sludge growth yield was estimated by dividing the measured increase in sludge by the approximated COD removed during the 7 months of operation. The gravity drained, sludge weight increased from 300 grams (elutriated) to 357 grams (not elutriated) (a 19% gain in wet sludge weight after 7 months). Converting grams of wet weight to grams of VSS (9.5%) resulted in 5.4 g VSS sludge produced. The total estimated COD consumed during the

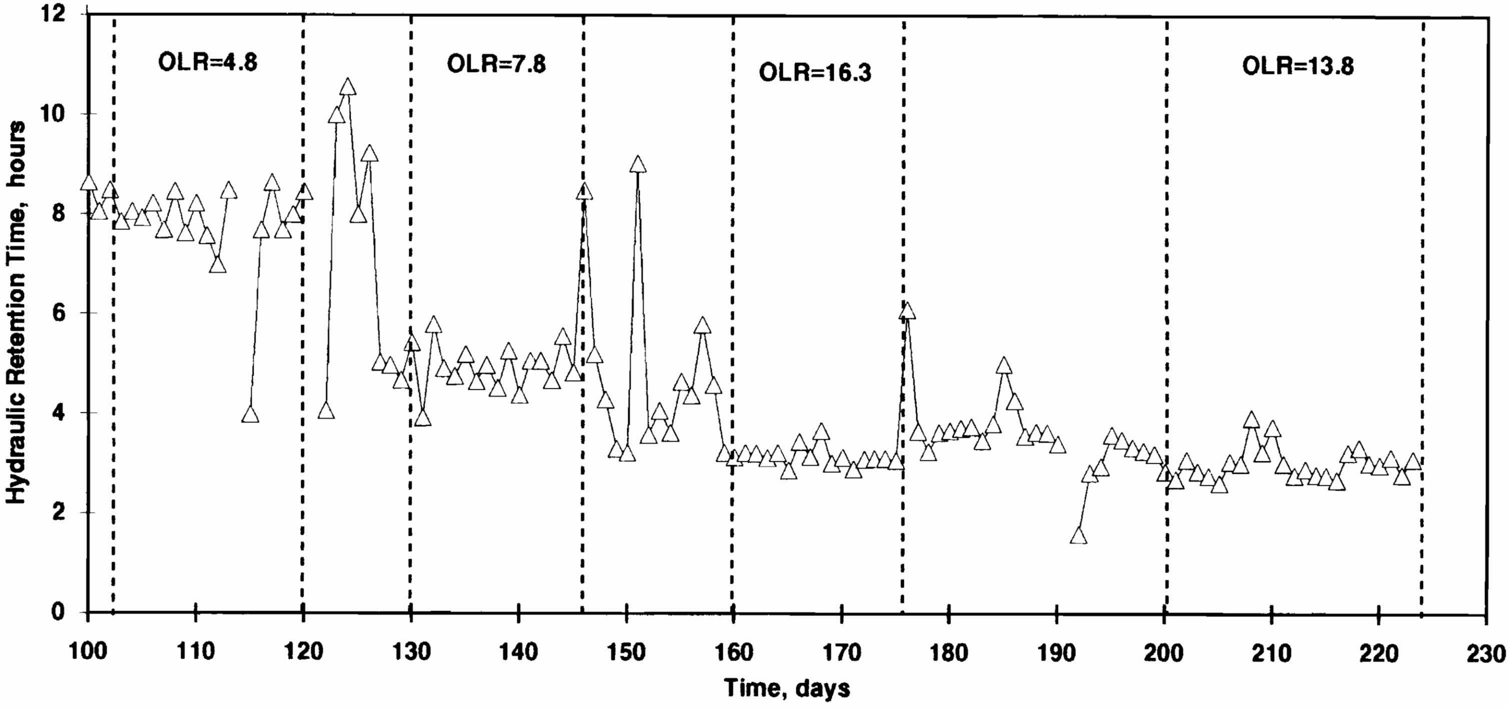

Figure 17. Hydraulic Retention versus Time for the Continuous UASB Experiment.

entire experiment was 1,200 g. Thus, the sludge yield was calculated at 0.005 g VSS/g COD removed. This sludge yield is much lower than the reported yields in the literature. Sludge was not removed from the laboratory UASB during the 223 day experiment.

The sludge height increased from 4 inches to 5.25 inches resulting in a 30 % volume increase during the 7 months of operation. Accurate measurements of sludge volume in the reactor were difficult due to occluded gas pockets in the sludge bed. A small amount of solids resembling clam waste solids was observed in the sludge and may have contributed to an over-estimate of biomass growth. Some granules were observed in the effluent from the reactor, but did not contribute significantly to the TSS of the effluent. Minimal washout of granules was observed.

A theoretical sludge yield for the laboratory UASB was calculated using literature values. For a mixed anaerobic culture utilizing an acetate substrate the maximum cell yield is reported by de Zeeuw to be 0.17 Kg VSS/kg COD removed (4). Based on this reported theoretical yield, the following calculations for sludge production in the laboratory UASB were made: For an average COD removal of 1,200 mg COD/L (4,542 mg COD/gallon wastewater) the theoretical sludge production would equal 204 mg VSS sludge/L (772 mg VSS sludge/gallon) or 772 g VSS sludge/1000 gallons treated. If the settled sludge contained 10% VSS, then the volume of sludge produced would be 7720 g sludge/1000 gallons treated or 17 lbs sludge/1000 gallons, or approximately 2 gallons of wet sludge per 1000 gallons of wastewater treated.

4.6 Other UASB Reactor Operations

The alkalinity of the wastewater varied, but was sufficient to buffer the microbial acidification of the wastewater and prevent wide pH shifts. The alkalinity of the reactor effluent was higher than the influent suggesting that the CO_2 evolved from the biogas participated in a bicarbonate buffer system. No additions of bicarbonate or NaOH chemicals were made after day 100. Prior to day 100, when the loading rate (less than 2 g COD/L/d) and generation of CO_2 (in the biogas) were low, it was necessary to add chemicals to buffer the pH.

The pH of the reactor over the course of the experiment is depicted in Figure 18. An operator error resulted in a large decrease in pH to pH 5.4 (day 185) possibly due to a build up of VOA. However, the reactor recovered after pumping tap water through the system for 9 hours at a flow rate corresponding to approximately 3 HRTs (3 liters/hour) and then restarting the influent flow.

The temperature of the reactor fluid in the settler and the temperature of the biogas stored in the Marriotte flask are shown in Figure 19. The average value for the reactor in the incubator was 32 °C. The biogas measured and stored in the Marriotte flask (see Figure 8) followed ambient laboratory temperatures and averaged 25 °C.

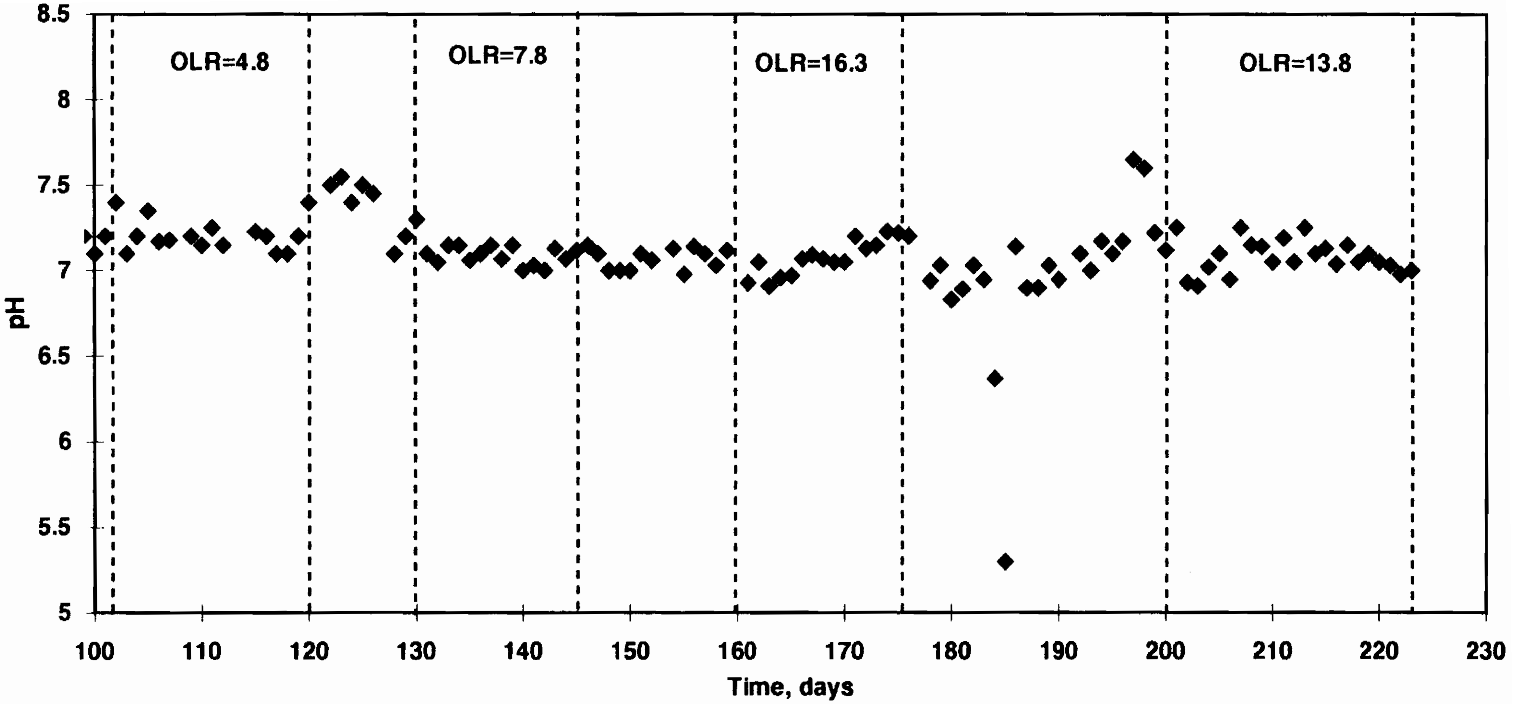

Figure 18. Continuous UASB daily Reactor pH.

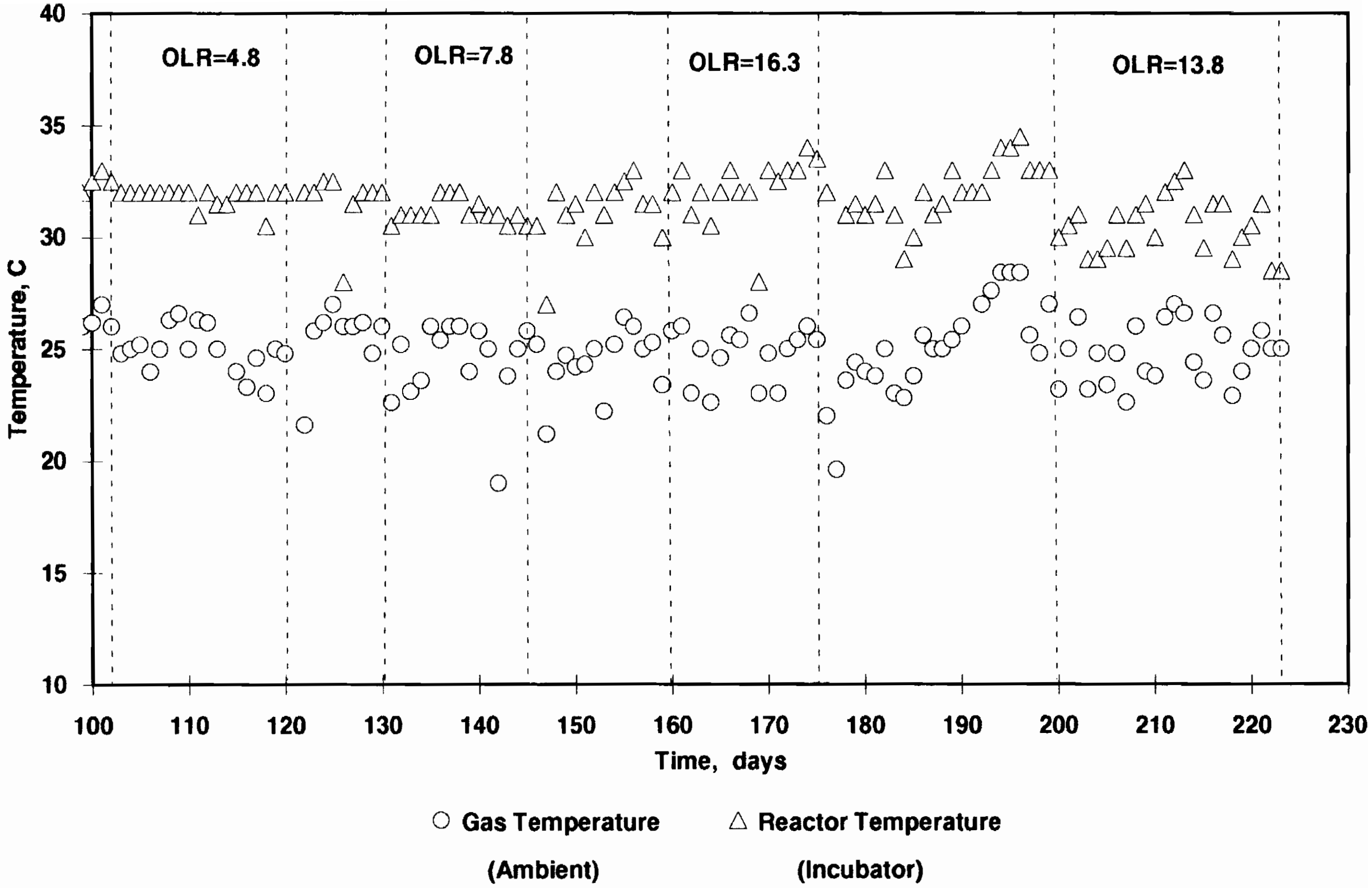

Figure 19. Continuous UASB daily Incubator Temperature and Ambient Gas Temperature.

4.7 UASB Reactor Volume and Cost Estimates

The full-scale UASB reactor volume required to treat the clam-processing wastewater tested was estimated using data obtained during the period of OLR=13.8. The volume of the reactor was calculated using the following assumptions: an average wastewater flow rate of 350,000 gallons of effluent per day (1,325 m^3/day), an average COD removal value of 1.2 g COD removed per liter of wastewater treated (4.5 g COD/gallon), and an organic loading rate of 13.8 g COD/L/d. The reactor volume required was calculated by multiplying 1.2 kg COD/m^3 (COD removed per liter of wastewater treated) times 1,325 m^3/d (wastewater daily flow rate) and dividing by 13.8 kg COD/m^3/d (appropriate OLR) which resulted in a 115 m^3 UASB reactor. Note that no safety factors or considerations for scale-up have been included in the calculations.

Lettinga (44), in 1984, estimated the cost of an UASB system, including facilities for gas utilization, a control room, and heat exchangers to be between $500,000 and $750,000 for a 1,000 m^3 (265,000 gallon) plant. Based on an average inflation rate of 6%, the 1991 cost of the Lettinga system would be $1,127,500. The FAEC treatment system shown in Figure 5 (45 m^3) cost approximately $120,000 in 1986 (70). Hence, the cost of the FAEC system in 1991 U. S. dollars would be approximately $160,500. By interpolating between the cost of the two UASB systems, a 115 m^3 UASB reactor for clam-processing wastewater would cost approximately $250,000.

5.0 CONCLUSIONS

Laboratory testing revealed that a UASB system has the capability of maintaining organic removals of up to 95% for soluble COD at a HRT as low as 3 hours for the clam-processing wastewater tested. No pH adjustment was required due to a stable alkalinity and buffering system within the reactor. Organic loading rates of up to 13.8 g COD/L/d provided effluent quality consistently below 250 ppm BOD_5 and 150 ppm TSS. An organic loading rate of 4.8 g COD/L/d resulted in average effluent values of 77 and 66 mg/L for BOD_5 and TSS, respectively.

Batch experimental results revealed that the metabolic activity of the methanogens was above 1 g CH_4-COD/g VS/d while utilizing the clam-processing wastewater at a sodium concentration of 4,200 mg Na^+/L. Batch experiments showed increasing inhibition of methanogenesis at sodium concentrations above 5,000 mg Na^+/L. Batch experiments indicated that there was no need for nutrient supplementation.

The nature of the wastewater used in the experiments changed over time because several samples were collected at monthly intervals. Due to the changes in the wastewater it was necessary to adjust the HRT to maintain the correct OLR, thus making it difficult to evaluate their individual contributions to effluent quality. The wastewater from one clam-processing facility was tested in this project.

Based on the best conditions of this study the operational parameters provided in Table 8 are recommended for UASB treatment of clam-processing wastewater.

Table 8. Recommended Operating Parameters for a UASB facility treating Clam-Processing Wastewater based on Experimental Results and Literature Values.

Parameter	Range
Organic Load	10 - 13 gram COD/L/day
Temperature	30 - 35 °C
Alkalinity	1,000 to 2,000 mg/l as $CaCO_3$
pH	7.0 - 7.8
Nutrients	no additions required
HRT	greater than 2.5 hours

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Appendix A

Appendix A: Data from Anaerobic Batch Reactor Experiments

First Feeding Nutrient Test

		Clam	Clam	Clam	Clam	Nutrients/ Clam	Nutrients/ Clam	Nutrients/ Clam	Nutrients/ Clam
Date	**Time**	1	2	3	4	5	6	7	8
	hrs	Cumulative methane production, ml							
Jan-19	22	28.29	28.77	28.83	26.68	29.03	27.5	29.33	29.12
Jan-19	23.25	31.62	32.39	36.07	32.22	31.49	28.73	30.61	30.75
Jan-20	0.75	34.11	34.93	38.87	35.08	32.89	30.07	32.44	32.15
Jan-20	5	42.68	44.58	44.81	44.86	39.21	35.71	37.21	36.83
Jan-20	7.5	51.94	53.02	50.29	54.33	45.79	41.62	42.19	41.03
Jan-20	10	64.02	64.57	55.69	67.39	56.59	51.15	51.55	48.29
Jan-20	12	73.54	70.86	59.16	76.85	64.75	58.81	58.74	54.07
Jan-20	15	90.06	90.06	66.26	92.75	80.09	73.76	71.39	63.88
Jan-20	17	99.98	99.47	70.03	101.98	88.97	82.93	78.5	70.12
Jan-20	18.6	106.96	102.96	72.12	107.77	94.93	88.73	83.35	74.53
Jan-20	20.5	115.28	113.47	74.29	115.77	102.68	95.6	89.39	79.58
Jan-21	7.5	155.14	153.37	79.65	157.3	142.2	130.69	125.44	109.92
Jan-21	9	159.77	157.2	79.98	162.23	146.53	135.86	129.57	114.09
Jan-21	12	168.22	159.72	80.16	171.36	155.13	143.86	137.72	121.45
Jan-21	20.25	189.07	181.23	81.06	195.02	178.82	165.52	158.94	141.82
Jan-22	18	240.28	243.43	82.9	251.45	238.17	224.4	211.22	190.03
Jan-23	9	263.6	265.3	83.1	275.2	257.1	246.9	235.5	214.1
Jan-26	19	379.1	379.04	87.51	395.84	352.31	308.01	315.06	296.47
Jan-26	24	388.22	378.93	88.44	405.92	359.32	310.65	326.48	306.5

		Blank	Blank	Nutrient/ Blank	Nutrient/ Blank
Date	**Time**	12	13	14	15
	hrs	Cumulative methane production, ml			
Jan-19	22	28.11	26.87	28.31	28.43
Jan-19	23.25	38.52	31.79	30.52	28.52
Jan-20	0.75	56.67	53.03	40.07	29.72
Jan-20	5	66.41	67.46	43.96	31.12
Jan-20	7.5	67.83	68.48	43.93	31.08
Jan-20	10	68.09	69.26	43.99	31.13
Jan-20	12	68.35	69.84	44.07	31.16
Jan-20	15	69.53	70.86	44.25	31.2
Jan-20	17	69.73	71.05	44.23	31.17
Jan-20	18.6	69.72	71.18	44.23	31.15
Jan-20	20.5	69.69	71.14	44.2	31.12
Jan-21	7.5	69.53	71.8	44.32	31.04
Jan-21	9	69.46	71.8	44.31	31.03
Jan-21	12	69.4	71.75	44.31	30.98
Jan-21	20.25	69.32	71.63	44.32	30.88
Jan-22	18	68.99	74.21	45.56	30.59
Jan-23	9	68.63	74.41	45.7	30.4
Jan-26	19	67.64	77.82	48.29	29.5
Jan-26	24	67.52	81.92	49.75	29.5

Second Feeding Nutrient Test

		Clam	Clam	Nutrient/ Clam	Nutrient/ Clam	Blank	Blank	Nutrient/ Blank	Nutrient/ Blank
DATE	TIME	1	2	5	8	12	13	14	15
	hrs	Cumulative methane production, ml							
Jan-27	4	28.29	28.77	29.03	29.12	28.11	26.87	28.31	28.43
Jan-27	7.5	43.27	50.23	45.72	50	28.11	27.18	31.37	28.43
Jan-27	10	57.5	66.31	58.13	64.42	28.11	27.14	32.17	28.44
Jan-27	12	71.18	80.8	78.72	76.8	28.12	27.28	34.36	28.44
Jan-27	15	100.25	111.66	113.25	102.88	28.12	27.85	39.39	28.44
Jan-27	17	113.69	126.5	123.03	119.28	28.12	28.08	40.21	28.44
Jan-27	19	125.9	141.7	130.2	134.7	28.12	28.08	40.21	28.5
Jan-27	21.5	138.8	159.9	176	153.5	28.12	28.08	40.21	28.5
Jan-28	7.5	189.4	222.8	215	217.3	28.12	28.08	40.21	28.5
Jan-28	9.25	197.7	233	261.5	227.9	28.12	28.08	40.21	28.5
Jan-28	12.25	210.2	248.25	271	244.4	28.12	28.08	40.21	28.5
Jan-28	15.5	227.6	267.2	277.9	265.8	28.12	28.08	40.21	28.5
Jan-28	18.25	241.7	282.2	283.7	282.25	28.12	28.08	40.21	28.5
Jan-29	6.5	330	370	333.9	370	28.12	28.08	40.21	28.5
Jan-30	11	407.8	414.7	360.4	446.3	28.12	28.08	40.21	28.5
Jan-30	14	423.3	417.9	365	449.4	28.12	28.08	40.21	28.5
Jan-31	18.45	490.7	423.5	409.7	483.3	28.12	28.08	40.21	28.5
Feb-1	11.5	498.9	425.3	427.5	484.8	28.12	28.08	40.21	28.5
Feb-2	18	542.6	438.8	457.6	500.6	28.12	28.08	40.21	28.5

Third Feeding Nutrient Test

Date	Time	2	4	5	7	12	13	14
	hrs	Cumulative methane production, ml						
Feb-2	23	28.77	26.68	29.03	29.33	28.11	26.87	28.31
Feb-3	1.5	45.04	50.11	48.98	48	32.16	27.2	28.33
Feb-3	8	92.9	114.8	98.32	101	34.4	27.2	28.33
Feb-3	11	117.8	143.6	120.4	125.2	34.6	27.2	28.33
Feb-3	14.75	161.1	192.5	160.5	167.3	34.6	27.2	28.33
Feb-3	17.5	167.6	222.3	171.5	195.5	35.3	28.1	28.33
Feb-3	21.75	207.5	263.6	186.3	237	35.3	28.1	28.33
Feb-4	9.75	294.2	346.1	290	332.9	35.3	28.1	28.33
Feb-4	14.5	327.8	375.2	313.3	369.1	35.8	28.1	28.33
Feb-4	17.75	341	385.1	327.1	383.7	35.8	28.1	28.33
Feb-5	18.5	414.7	442.5	410.7	462.2	35.8	28.1	28.33
Feb-6	13	447.9	471.6	452.5	494.2	35.8	28.1	28.33
Feb-7	18.5	477.3	497.8	497.2	515	35.8	28.1	28.33
Feb-8	12	496.7	515.4	511.6	531	35.8	28.1	28.33
Feb-10	15.25	527.4	558.9	553.4	550.1	35.8	28.1	28.33
Feb-11	17.5	527.8	572.7	564.3	555.1	35.8	28.1	28.33

Biological Methane Potential Data (BMP)

	Third Feeding				
Date	Time	Clam	Clam	Blank	Blank
	hrs	Cumulative methane production, ml			
9-14-89	9	29.29	28.83	29.03	27.5
9-14-89	15	99.7	110	30.02	27.9
9-15-89	3	183	188	30.8	28.4
9-15-89	11	280	281.9	31	28.7
9-16-89	9	410.5	422.6	31.1	28.9
9-16-89	17	432.55	443.4	31.1	28.9
9-17-89	9	460.3	467.1	31.1	28.9
9-18-89	10.5	485.5	492.6	31.1	28.9
9-19-89	10.5	502.3	500.8	31.1	28.9
9-20-89	10.5	520	520	31.1	28.9
9-21-89	10.5	521	522	31.1	28.9
	Second Feeding				
'9-5	12	28.42	28.81	29.03	27.5
'9-5	23	127.48	138.69	29.1	27.7
'9-7	8.75	338.85	363.7	29.1	27.7
'9-7	17	359.8	385.1	29.1	27.7
'9-8	9	383.4	405.1	29.1	27.7
'9-9	10.5	410.2	424.4	29.1	27.7
'9-12	11	453.3	449	29.1	27.7
	First Feeding				
'8-30	0	28.42	28.82	29.03	27.5
'8-30	9	85.5	89.5	29.5	27.5
'8-31	19	139	150.4	29.5	27.5
'8-31	8.5	208.5	230	29.55	27.5
'9-3	15.5	232.6	260.1	29.55	27.5
'9-4	12	356.7	400.1	29.55	27.5
	11	375.2	411	29.55	27.5

Salt toxicity batch experiment

May 1990 mg/L Na+

		4200	4200	4200	8400	8400	8400	12000	12000	12000	Blank	Blank	Blank
DATE	TIME	Flask 1	2	3	4	5	6	7	8	9	13	14	15
	hrs	Cumulative gas production, ml											
5-16	17	28.29	28.77	28.83	26.68	29.03	27.5	29.33	29.12	28.42	26.87	28.31	28.43
5-16	22	35.85	36.5	35.25	31.7	35.9	32.5	34.7	35.2	33.2	31.8	32.85	33.4
5-17	10	50.7	56.5	60.3	35	37.9	35.9	37.75	38.1	35.2	34.25	35.1	36.95
5-17	18	78.3	93.3	92.4	38.5	48.8	39.85	39.85	40.75	38.9	35.1	36.5	37.1
5-18	17	152	171.5	167.8	56.3	66.3	58.3	57	51	56	40	41	49
5-19	11	203.6	220.4	214	67.9	71.8	68.65	70.6	62.4	62.5	44.2	43.2	52.6
5-20	8	255.8	276.3	262.3	75.7	79.4	76.5	77.6	70.6	67.6	47.2	45.7	55.7
5-21	17	322.1	338.6	324.3	90.1	93.8	89.9	85.3	78.6	74.5	53.4	51.7	60
5-22	9	341.5	352.9	342.2	94	98.65	93.5	87.2	80	76.2	53.7	52.4	62.1
5-23	18	370.6	375.4	366.9	98.2	104	96.5	90.2	80	76	53.2	53.3	62
5-25	18	383.4	393.8	369.6	102.6	113.6	107.8	93.5	82.5	79.5	55.9	56.2	64.6

Salt Toxicity

mg/L Na+

		4200	4200	5250	5250	5250	6300	6300	6300	Blank	Blank	Blank
DATE	TIME	Flask 2	3	4	5	6	7	8	9	10	11	12
	hrs	Cumulative methane production, ml										
5-25	20.5	28.77	28.83	26.68	29.03	27.5	29.33	29.12	28.42	29.14	28.81	28.11
5-26	9	48.6	47.5	40.6	43.2	41.4	38.8	38.4	42.7	38.5	37.4	33.6
	13.5	62.3	62	51.1	54.1	50.8	46.2	44.7	50	41.8	39.4	35.5
5-27	0.5	101.6	102.9	81	84.7	79.6	69.3	68.2	73	49.9	45.5	39.3
	22.5	163.6	169.6	128	130.8	122.3	102.4	103.6	108	56.1	50.3	39.9
5-28	10.5	192.5	195	150	152.6	142.4	120.7	119.4	123	57.4	51.3	40
	22	217.3	220	169	173	164	139.4	135.5	137	57.4	52	40.9
5-29	18	258.6	271.5	204	209.6	200.7	174.5	166	163.7	57.4	53.6	41.3
5-30	18	302.5	312	233.4	241.3	234.5	208	195.8	192.4	57.4	53.6	41.6
5-31	11.5	344.2	345	250.8	261.5	255.3	229.7	217.7	213.5	57.4	53.6	41.6
6-1	19	399.7	400	292	293.6	291.2	263.2	264.1	262	57.4	53.6	41.6
6-3	12	417.4	410	300.4	303	302	270	271	269	57.4	53.6	41.6

Appendix B

Appendix B: Data from Laboratory Scale UASB Reactor Experiment

Date	Time	Day	Time Period hr	pH	Infl COD g/l	Infl L	Infl L/d	Temp Reactor °C	Gas ambient L	Temp Gas °C	Gas 32 °C L
10-Dec	12	100	25	7.1	0.6	2.78	2.67	32.5	0.35	26.2	0.35
11-Dec	12.25	101	24.25	7.2	1.5	2.98	2.95	33	0.365	27	0.37
12-Dec	11	102	22.75	7.4	1.5	2.83	2.99	32.5	0.23	26	0.23
13-Dec	11	103	24	7.1	1.5	3.06	3.06	32	0.9	24.8	0.92
14-Dec	11	104	24	7.2	1.5	2.98	2.98	32	0.76	25	0.77
15-Dec	11	105	24	7.35	1.5	3.03	3.03	32	0.86	25.2	0.87
16-Dec	11	106	24	7.17	1.5	2.92	2.92	32	0.82	24	0.84
17-Dec	11.5	107	24.5	7.18	1.5	3.12	3.06	32	0.89	25	0.90
18-Dec	11	108	23.5		1.5	2.84	2.90	32	0.67	26.3	0.68
19-Dec	11.75	109	24.75	7.2	1.5	3.15	3.05	32	0.61	26.6	0.62
20-Dec	10.5	110	22.75	7.15	1.5	2.92	3.08	32	0.3	25	0.31
21-Dec	10.75	111	24.25	7.25	1.5	3.17	3.14	31	0.72	26.3	0.73
22-Dec	13.5	112	26.75	7.15	1.5	3.43	3.08	32	1.12	26.2	1.13
23-Dec	11.5	113	22		1.5	2.83	3.09	31.5	0.9	25	0.92
24-Dec	12	114			1.5			31.5			
25-Dec	12	115	48.5	7.23	1.625	6.01	2.97	32	1.19	24	1.21
26-Dec	12	116	24	7.2	1.625	3.12	3.12	32	0.48	23.3	0.49
27-Dec	11	117	23	7.1	1.625	2.78	2.90	32	1.42	24.6	1.45
28-Dec	11	118	24	7.1	1.625	3.12	3.12	30.5	0.85	23	0.87
29-Dec	11	119	24	7.2	1.625	3	3.00	32	1.01	25	1.03
30-Dec	11	120	24	7.4	1.625	2.84	2.84	32	0.96	24.8	0.98
31-Dec	11	121			1.625						
1-Jan	15	122	52	7.5	1.625	5.9	2.72	32	1.74	21.6	1.79
2-Jan	12.5	123	21.5	7.55	1.425	2.4	2.68	32	0.93	25.8	0.94
3-Jan	10.5	124	22	7.4	1.425	2.27	2.48	32.5	0.76	26.2	0.77
4-Jan	13.5	125	27	7.5	1.425	3	2.67	32.5	1.25	27	1.26
5-Jan	13	126	23.5	7.45	1.425	2.6	2.66	28	0.48	26	0.49
6-Jan	17	127	28		1.67	4.76	4.08	31.5	0.44	26	0.45
7-Jan	15	128	22	7.1	1.67	4.82	5.26	32	1.82	26.2	1.84
8-Jan	15	129	24	7.2	1.67	5.13	5.13	32	1.91	24.8	1.94
9-Jan	14	130	23	7.3	1.67	4.42	4.61	32	1.7	26	1.72
10-Jan	19.5	131	29.5	7.1	1.67	6.1	4.96	30.5	1.86	22.6	1.91
11-Jan	15	132	19.5	7.05	1.67	4.14	5.10	31	1.45	25.2	1.47
12-Jan	15	133	24	7.15	1.55	4.88	4.88	31	1.18	23.1	1.21
13-Jan	16.75	134	25.75	7.15	1.55	5.05	4.71	31	1.77	23.6	1.81
14-Jan	14.5	135	21.75	7.06	1.55	4.62	5.10	31	2.19	26	2.22
15-Jan	15	136	24.5	7.1	1.55	5.16	5.05	32	1.19	25.4	1.21
16-Jan	14	137	23	7.15	1.55	4.82	5.03	32	1	26	1.01
17-Jan	14.5	138	24.5	7.07	1.55	5.3	5.19	32	2.04	26	2.07
18-Jan	12	139	21.5	7.15	1.55	4.56	5.09	31	1.6	24	1.63
19-Jan	14	140	26	7	1.6	5.47	5.05	31.5	2.44	25.8	2.47
20-Jan	13.5	141	23.5	7.03	1.6	4.73	4.83	31	2.58	25	2.62
21-Jan	13	142	23.5	7	1.6	4.73	4.83	31	1.64	19	1.70
22-Jan	15.5	143	26.5	7.13	1.6	5.13	4.65	30.5	2.09	23.8	2.13
23-Jan	13	144	21.5	7.07	1.6	4.31	4.81	31	1.51	25	1.54
24-Jan	14	145	25	7.12	1.65	4.96	4.76	30.5	2.29	25.8	2.32
25-Jan	15	146	25	7.15	1.65	2.83	2.72	30.5	2.31	25.2	2.35
26-Jan	14	147	23	7.1	1.65	4.62	4.82	27	1.05	21.2	1.08
27-Jan	13	148	23	7	1.65	5.59	5.83	32	1.96	24	2.00
28-Jan	12.75	149	23.75	7	2.2	7.26	7.34	31	4.76	24.7	4.84
29-Jan	13.75	150	25	7	2.2	7.43	7.13	31.5	4.99	24.2	5.09

Date	Time	Day	Time Period hr	pH	Infl COD g/l	Infl L	Infl L/d	Temp Reactor °C	Gas ambient L	Temp Gas °C	Gas 32 °C L
30-Jan	13.25	151	23.5	7.1	2.2	2.66	2.72	30	2	24.3	2.04
Jan-31	14	152	24.75	7.06	2.2	6.69	6.49	32	4.66	25	4.74
Feb-1	14	153	24		2.2	5.9	5.90	31	3.6	22.2	3.70
Feb-2	15	154	25	7.13	2.2	6.6	6.34	32	4.48	25.2	4.55
Feb-3	14.5	155	23.5	6.98	2.2	5.16	5.27	32.5	3.42	26.4	3.46
Feb-4	14.5	156	24	7.14	2.2	5.47	5.47	33	4	26	4.05
Feb-5	14	157	23.5	7.1	2.2	4.14	4.23	31.5	2.19	25	2.23
Feb-6	15	158	25	7.03	2.2	5.23	5.02	31.5	3.57	25.3	3.63
Feb-7	15	159	24	7.12	2.2	7.43	7.43	30	5.25	23.4	5.37
Feb-8	15.5	160	24.5		2.2	7.65	7.49	32	5.22	25.8	5.29
Feb-9	15	161	23.5	6.93	2.2	7.43	7.59	33	5.2	26	5.27
Feb-10	15	162	24	7.05	2.2	7.48	7.48	31	4.96	23	5.08
Feb-11	15.5	163	24.5	6.91	2.2	7.68	7.52	32	5.45	25	5.54
Feb-12	15.5	164	24	6.96	2.2	7.43	7.43	30.5	5.22	22.6	5.35
Feb-13	18	165	26.5	6.97	2.2	8.36	7.57	32	5.94	24.6	6.05
Feb-14	15.75	166	21.75	7.07	2.2	6.95	7.67	33	5.06	25.6	5.13
Feb-15	15.5	167	23.75	7.09	2.2	7.63	7.71	32	5.24	25.4	5.32
Feb-16	11.75	168	20.25	7.07	2.1	6.55	7.76	32	4.31	26.6	4.36
Feb-17	13	169	25.25	7.05	2.1	7.94	7.55	28	4.17	23	4.27
Feb-18	14	170	25	7.05	2.1	7.65	7.34	33	4.78	24.8	4.86
Feb-19	15.33	171	25.33	7.2	2.1	8.25	7.82	32.5	5.69	23	5.82
Feb-20	15.25	172	23.92	7.13	2.1	7.74	7.77	33	4.88	25	4.96
Feb-21	15.33	173	24.08	7.15	2.1	7.71	7.68	33	4.92	25.4	5.00
Feb-22	15	174	23.67	7.23	2.1	7.71	7.82	34	5.09	26	5.16
Feb-23	15.25	175	24.25	7.22	2.1	7.82	7.74	33.5	4.12	25.4	4.18
Feb-24	12.75	176	14.75	7.2	2	3.94	6.41	32	1.94	22	1.99
Feb-25	12	177	23.25		2	6.56	6.77		4.66	19.6	4.83
Feb-26	15.25	178	27.25	6.94	2	7.38	6.50	31	5	23.6	5.11
Feb-27	14.75	179	23.5	7.03	2	6.6	6.74	31.5	4.78	24.4	4.87
Feb-28	15.5	180	24.75	6.83	2	6.55	6.35	31	4.18	24	4.26
Mar-1	15.5	181	24	6.89	2	6.46	6.46	31.5	4.26	23.8	4.35
Mar-2	15.2	182	23.7	7.03	2	6.41	6.49	33	3.93	25	4.00
Mar-3	16.75	183	25.55	6.95	2	6.91	6.49	31	4.66	23	4.77
Mar-4	16	184	23.25	6.37	2	6.29	6.49	29	3.8	22.8	3.89
Mar-5	9.5	185	17.5	5.3	1.8	4.79	6.57	30	0.77	23.8	0.79
Mar-6	15	186	20	7.14	1.8	5.61	6.73	32	2.01	25.6	2.04
Mar-7	15	187	24	6.9	1.8	6.72	6.72	31	1.99	25	2.02
Mar-8	15.2	188	24.2	6.9	1.8	6.58	6.53	31.5	2.76	25	2.81
Mar-9	15.25	189	24.05	7.03	1.8	6.61	6.60	33	3.18	25.4	3.23
Mar-10	17	190	25.75	6.95	1.8	7.03	6.55	32	1.96	26	1.99
Mar-11	0	191			1.8			32			
Mar-12	15	192	46	7.1	1.7	15.05	7.85	32	4.64	27	4.69
Mar-13	16.5	193	25.5	7	1.7	8.5	8.00	33	4.42	27.6	4.46
Mar-14	16.75	194	24.25	7.17	1.7	8.11	8.03	34	4.11	28.4	4.13
Mar-15	17	195	24.25	7.1	1.8	6.69	6.62	34	3.09	28.4	3.11
Mar-16	16.5	196	23.5	7.17	1.8	6.89	7.04	34.5	2.79	28.4	2.80
Mar-17	16.7	197	24.2	7.65	1.6	7.2	7.14	33	1.38	25.6	1.40
Mar-18	18.5	198	25.8	7.6	1.7	7.35	6.84	33	3	24.8	3.05
Mar-19	15.5	199	21	7.22	1.7	7.51	8.58	33	3.35	27	3.38
Mar-20	15.5	200	24	7.12	1.7	8.42	8.42	30	2.28	23.2	2.33
Mar-21	17	201	25.5	7.25	1.7	8.9	8.38	30.5	2.75	25	2.80
Mar-22	16	202	23	6.93	1.7	7.77	8.11	31	4.48	26.4	4.53
Mar-23	16	203	24	6.91	1.7	8.42	8.42	29	4.56	23.2	4.66
Mar-24	16	204	24	7.02	1.7	8.7	8.70	29	5.23	24.8	5.32

Mar-25	18	205	26	7.1	1.7	9.16	8.46	29.5	4.77	23.4	4.88
Mar-26	17	206	23	6.95	1.7	7.88	8.22	31	4.88	24.8	4.97
Mar-27	16.3	207	23.3	7.25	1.7	8	8.24	29.5	4.7	22.6	4.82
Mar-28	17.75	208	18	7.15	1.7	6.1	8.13	31	3.61	26	3.66
Mar-29	19.5	209	25.75	7.14	1.8	7.41	6.91	31.5	3	24	3.06
Mar-30	16	210	20.5	7.05	1.8	6.41	7.50	30	2.9	23.8	2.96
Mar-31	15.5	211	23.5	7.19	1.7	7.99	8.16	32	4.5	26.4	4.55
Apr-1	17	212	25.5	7.05	1.7	8.65	8.14	32.5	5.14	27	5.19
Apr-2	16	213	23	7.25	1.7	8.25	8.61	33	5.1	26.6	5.16
Apr-3	16.5	214	24.5	7.1	1.7	8.6	8.42	31	5.1	24.4	5.20
Apr-4	16.5	215	24	7.13	1.7	8.65	8.65	29.5	4.85	23.6	4.95
Apr-5	17.75	216	25.25	7.04	1.7	8.96	8.52	31.5	4.78	26.6	4.83
Apr-6	16.75	217	23	7.15	1.7	7.42	7.74	31.5	2.69	25.6	2.73
Apr-7	16	218	23.25	7.05	1.7	7.21	7.44	29	3.11	22.9	3.18
Apr-8	15	219	23	7.1	1.7	7.63	7.96	30	4.12	24	4.20
Apr-9	16.5	220	25.5	7.05	1.7	8.56	8.06	30.5	4.14	25	4.21
Apr-10	15.5	221	23	7.03	1.7	7.34	7.66	31.5	3.69	25.8	3.74
Apr-11	16	222	24.5	6.98	1.7	8.8	8.62	28.5	4	25	4.07
Apr-12	18.5	223	26.5	7	1.7	8.56	7.75	28.5	4	25	4.07
		'Average values (not including first 4 to 5 days of OLR period)									
	average OLR=4.8			7.2	1.5	3.2	3.0	31.8	0.9	25.0	0.9
	average OLR=7.8			7.1	1.6	4.8	4.9	31.2	1.8	24.6	1.8
	average ORL=16.3			7.1	2.1	7.6	7.7	32.2	5.0	24.8	5.0
	average OLR=13.8			7.1	1.7	8.0	8.1	30.5	4.2	24.9	4.3
Date	**Time**	**Day**	**Period**	**pH**	**Infl COD gr/l**	**Infl liters**	**Infl L/d**	**Temp Reactor °C**	**Gas ambient °C**	**Temp Gas °C**	**Gas 32 °C**

OL gCOD/ L/day	Gas g CH4- COD/ L/day	Infl TSS mg/l	Effl TSS mg/l	%TSS Removed	Infl TCOD mg/L	Effl TCOD mg/L	Infl SCOD mg/l	Effl SCOD mg/l	%TCOD Removed	%SCOD Removed	Effl TBOD mg/L
1.60	0.87										
4.42	0.94				1500	295	1160	210	80.3	81.9	
4.48	0.63	508	28	94.5	1500	235	1160	178	84.3	84.7	80
4.59	2.35				1500	290	1160	195	80.7	83.2	
4.47	1.99				1500	370	1160	240	75.3	79.3	
4.55	2.25				1500		1160				
4.38	2.15				1500	508	1160	439	66.1	62.2	
4.58	2.28				1500	544	1160	483	63.7	58.4	
4.35	1.78				1500	570	1160	458	62.0	60.5	
4.58	1.54	710	52	92.7	1500	537	1160		64.2		
4.62	0.83				1500	309	1160		79.4		
4.71	1.85	830	75	91.0	1500	327	1160		78.2		
4.62	2.62				1500	482	1160		67.9		
4.63	2.57				1500	360	1160	280	76.0	75.9	83
					1500		1160				
4.83	1.54				1625	137	1260	100	91.6	92.1	
5.07	1.26				1625	178	1260		89.0		70
4.71	3.88				1625		1260				
5.07	2.24	785	70	91.1	1625		1260				
4.88	2.64				1625	540	1260		66.8		
4.62	2.51				1625		1260				
					1625		1260				
4.43	2.12	1140	46	96.0	1625	474	1260	351	70.8	72.1	
3.82	2.71				1425	500			64.9		
3.53	2.16				1425						
3.80	2.88	140	48	65.7	1425	600		434	57.9		106
3.78	1.28				1425	455		373	68.1		
6.81	0.98				1670	128			92.3		
8.78	5.17				1670	540		290	67.7		
8.57	4.99				1670						
7.70	4.62				1670	660	1420	350	60.5	75.4	
8.29	3.99				1670		1420				
8.51	4.66				1670	381	1420	253	77.2	82.2	108
7.56	3.10	351	64	81.8	1550		1210				
7.30	4.33				1550	472	1210	280	69.5	76.9	
7.90	6.29				1550	608	1210	410	60.8	66.1	
7.83	3.04				1550		1210				
7.80	2.72				1550	410	1210	292	73.5	75.9	
8.05	5.20	405	97	76.0	1550		1210				
7.89	4.68				1550	655	1210	454	57.7	62.5	214
8.08	5.87				1600		1280				
7.73	6.89	301	116	61.5	1600	604	1280	460	62.3	64.1	280
7.73	4.47				1600	665	1280	345	58.4	73.0	
7.43	4.97				1600		1280				
7.70	4.40				1600	550	1280	330	65.6	74.2	
7.86	5.73				1650		1400				
4.48	5.79				1650		1400				
7.95	2.90	874	80	90.8	1650	250	1400	133	84.8	90.5	
9.62	5.36				1650	443	1400	315	73.2	77.5	153
16.14	12.58				2200	1150	1760	970	47.7	44.9	
15.69	12.55				2200	983	1760	682	55.3	61.3	

OL gCOD/ L/day	g CH4- COD/ L/day	Infl TSS mg/l	Effl TSS mg/l	%TSS Removed	Infl TCOD mg/L	Effl TCOD mg/L	Infl SCOD mg/l	Effl SCOD mg/l	%TCOD Removed	%SCOD Removed	Effl TBOD mg/L
5.98	5.35				2200	1140	1760	840	48.2	52.3	
14.27	11.81				2200	900	1760	600	59.1	65.9	
12.98	9.50				2200	960	1760	620	56.4	64.8	
13.94	11.23				2200	1052	1760	830	52.2	52.8	
11.59	9.08				2200	1032	1760	809	53.1	54.0	354
12.03	10.42	960	95	90.1	2200	952	1760	825	56.7	53.1	
9.30	5.84				2200	833	1760	783	62.1	55.5	
11.05	8.95				2200	1210	1760	991	45.0	43.7	
16.35	13.79				2200		1760				
16.49	13.33				2200	1059	1760	921	51.9	47.7	500
16.69	13.83				2200		1760				
16.46	13.05				2200	1120	1760	905	49.1	48.6	
16.55	13.95	990	650	34.3	2200	1160	1760	924	47.3	47.5	
16.35	13.75				2200		1760				
16.66	14.08				2200	890	1760	685	59.5	61.1	660
16.87	14.56				2200		1760				
16.96	13.82	1000	540	46.0	2200	940	1760	790	57.3	55.1	
16.30	13.28				2100		1750				
15.85	10.43				2100	940	1750	770	55.2	56.0	570
15.42	12.00				2100	1000	1750	683	52.4	61.0	
16.42	14.18	880	590	33.0	2100	968	1750	706	53.9	59.7	
16.31	12.79				2100	867	1750	625	58.7	64.3	642
16.14	12.80				2100	988	1750	665	53.0	62.0	
16.42	13.44				2100	990	1750	706	52.9	59.7	
16.25	10.64	575	500	13.0	2100	867	1750	545	58.7	68.9	
12.82	8.33				2000	867	1620	625	56.7	61.4	165
13.54	12.80				2000		1620				
13.00	11.56				2000		1620				
13.48	12.78				2000	690	1620	420	65.5	74.1	
12.70	10.63				2000	728	1620		63.6		
12.92	11.18				2000	542	1620	314	72.9	80.6	
12.98	10.40				2000	690	1620		65.5		
12.98	11.52				2000	620	1620	293	69.0	81.9	
12.99	10.33				2000	657	1620	314	67.2	80.6	310
11.82	2.77				1800		1450				
12.12	6.29				1800		1450				
12.10	5.20				1800	1110	1450		38.3		
11.75	7.15				1800	1070	1450		40.6		
11.87	8.28				1800		1450				
11.79	4.76				1800	595	1450	200	66.9	86.2	290
					1800		1450				
13.35	6.28				1700	297	1440		82.5		
13.60	10.78				1700	587	1440		65.5		
13.64	10.51				1700	505	1440	174	70.3	87.9	
11.92	7.90				1800	443	1450		75.4		
12.67	7.36				1800	380	1450	132	78.9	90.9	
11.42	3.57				1600	380	1292		76.3		
11.62	7.30				1700		1440				
14.59	9.94				1700	484	1440	339	71.5	76.5	
14.31	5.99				1700	306	1440		82.0		300
14.24	6.76				1700	310	1440		81.8		225
13.78	12.16				1700	1050	1440	823	38.2	42.8	
14.31	11.99				1700	1010	1440		40.6		
14.79	13.68				1700	782	1440	554	54.0	61.5	

14.37	11.57	490	120	75.5	1700	460	1440	332	72.9	76.9	
13.98	13.32				1700	520	1440	360	69.4	75.0	
14.01	12.75				1700	552	1440	280	67.5	80.6	198
13.83	12.54				1700	548	1440	280	67.8	80.6	
12.43	7.33	510	90	82.4	1800	520	1445	236	71.1	83.7	210
13.51	8.91				1800	430	1445	240	76.1	83.4	
13.87	11.95				1700	420	1440	272	75.3	81.1	190
13.84	12.56				1700	380	1440		77.6		
14.63	13.83	550	60	89.1	1700	316	1440	200	81.4	86.1	
14.32	13.08				1700	300	1440		82.4		
14.71	12.73				1700	320	1440	200	81.2	86.1	200
14.48	11.81				1700		1440				
13.16	7.32				1700	340	1440	172	80.0	88.1	
12.65	8.45				1700	380	1440		77.6		
13.53	11.27				1700	440	1440		74.1		
13.70	10.18	532	72	86.5	1700	352	1440	220	79.3	84.7	
13.02	10.03				1700	280	1440		83.5		
14.65	10.24				1700	320	1440	240	83.5	83.3	
13.18	9.47				1700	240	1440		83.5		
4.7	2.1	775.0	65.7	91.6	1546.9	408.4	1197.5	352.0	73.2	69.8	76.5
7.8	4.7	352.3	92.3	73.1	1583.6	543.1	1263.6	353.0	65.6	71.9	200.7
16.3	13.0	818.3	543.3	30.7	2133.3	938.9	1753.3	686.1	55.7	60.8	624.0
13.8	11.2	520.5	85.5	83.4	1710.0	415.8	1440.5	252.7	76.9	82.5	199.5
OL gTCOD/ L/day	**Gas gCOD/ L/day**	**Infl TSS mg/l**	**Effl TSS mg/l**	**%TSS Removed**	**Infl SCOD mg/L**	**Effl TCOD mg/L**	**Infl SCOD mg/l**	**Effl SCOD mg/l**	**Total %COD Removed**	**Soluble %COD Removed**	**Effl TBOD mg/L**

Effl SBOD mg/L	Infl TBOD mg/L	Infl SBOD mg/L	HRT hours	SFV m/hr	Efficiency Gas/OL Ratio
			8.6	0.2	0.55
			8.1	0.2	0.21
75			8.5	0.2	0.14
			7.8	0.3	0.51
			8.1	0.2	0.44
			7.9	0.2	0.49
			8.2	0.2	0.49
			7.7	0.3	0.50
			8.5	0.2	0.41
			7.6	0.3	0.34
			8.2	0.2	0.18
			7.6	0.3	0.39
			7.0	0.3	0.57
66	1220	940	8.5	0.2	0.55
			4.0	0.5	0.32
60			7.7	0.3	0.25
			8.6	0.2	0.82
			7.7	0.3	0.44
			8.0	0.2	0.54
			8.5	0.2	0.54
			4.1	0.5	0.48
			10.0	0.2	0.71
			10.6	0.2	0.61
35			8.0	0.2	0.76
			9.2	0.2	0.34
			5.0	0.4	0.14
			5.0	0.4	0.59
			4.7	0.4	0.58
			5.4	0.4	0.60
			3.9	0.5	0.48
95	570		5.8	0.3	0.55
			4.9	0.4	0.41
			4.8	0.4	0.59
			5.2	0.4	0.80
			4.7	0.4	0.39
			5.0	0.4	0.35
			4.5	0.4	0.65
112	1410	1200	5.3	0.4	0.59
			4.4	0.4	0.73
193	1310	1200	5.1	0.4	0.89
			5.1	0.4	0.58
			4.7	0.4	0.67
			5.6	0.4	0.57
			4.8	0.4	0.73
			8.5	0.2	0.36
			5.2	0.4	0.36
93	1900	1575	4.3	0.5	0.56
			3.3	0.6	0.78
			3.2	0.6	0.80

Effl SBOD mg/L	Infl TBOD mg/L	Infl SBOD mg/L	HRT hours	SFV m/hr	Efficiency Gas/OL Ratio
			9.0	0.2	0.90
			3.6	0.6	0.83
			4.1	0.5	0.73
			3.6	0.5	0.81
207	1725	1690	4.7	0.4	0.78
			4.4	0.4	0.87
			5.8	0.3	0.63
			4.6	0.4	0.81
			3.2	0.6	0.84
465			3.1	0.6	0.81
			3.2	0.6	0.83
			3.2	0.6	0.79
			3.1	0.6	0.84
			3.2	0.6	0.84
505			2.9	0.7	0.85
			3.5	0.6	0.86
			3.1	0.6	0.81
			3.7	0.5	0.81
510	1890	1675	3.0	0.7	0.66
			3.1	0.6	0.78
			2.9	0.7	0.86
510	1900	1280	3.1	0.6	0.78
			3.1	0.6	0.79
			3.1	0.6	0.82
			3.1	0.6	0.65
160			6.1	0.3	0.65
			3.7	0.5	0.95
			3.3	0.6	0.89
			3.6	0.5	0.95
			3.7	0.5	0.84
			3.7	0.5	0.87
			3.7	0.5	0.80
			3.5	0.6	0.89
280			3.8	0.5	0.80
			5.0	0.4	0.23
			4.3	0.5	0.52
			3.6	0.6	0.43
			3.6	0.5	0.61
			3.6	0.5	0.70
265			3.4	0.6	0.40
					0.00
			1.6	1.2	0.47
			2.8	0.7	0.79
			3.0	0.7	0.77
			3.6	0.6	0.66
			3.5	0.6	0.58
			3.3	0.6	0.31
			3.3	0.6	0.63
			3.2	0.6	0.68
250			2.9	0.7	0.42
165			2.7	0.7	0.47
			3.1	0.6	0.88
			2.9	0.7	0.84
			2.8	0.7	0.92

			2.6	0.8	0.80
			3.0	0.6	0.95
112			3.0	0.7	0.91
			3.9	0.5	0.91
120			3.2	0.6	0.59
			3.7	0.5	0.66
130			3.0	0.7	0.86
			2.8	0.7	0.91
	1100	960	2.9	0.7	0.95
			2.8	0.7	0.91
140			2.8	0.7	0.87
			2.7	0.7	0.82
			3.2	0.6	0.56
			3.3	0.6	0.67
			3.0	0.7	0.83
			3.0	0.7	0.74
			3.1	0.6	0.77
			2.8	0.7	0.70
			3.1	0.6	0.72
67.0	1220.0	940.0	7.7	0.3	0.46
133.3	1096.7	1200.0	5.0	0.4	0.6
508.3	1895.0	1477.5	3.2	0.6	0.79
125.5	1100.0	960.0	3.0	0.7	0.81
Effl SBOD mg/L	**Infl TBOD mg/L**	**Infl SBOD mg/L**	**HRT hours**	**SFV m/hr**	**Gas/OL Ratio N/M**

Appendix C

Appendix C: Gas Chromatography Data for Methane Composition

99% Methane

Volume, microliters	**Chromatogram Area, square mm**
10	3.61
10	3.77
10	3.66
20	7.32
20	7.26
20	7.21
30	10.93
30	10.82
30	10.93
40	14.66
40	14.59
40	14.57
45	16.1
45	16.22

Appendix D

Appendix D: Nomenclature

ATA = Anaerobic Toxicity Assay

ATP = Adenozine Tri phosphate

BMP = Biological Methane Potential

BOD_5 = 5 day test for Biological Oxygen Demand

BTU = British Thermal Unit

CO_2 = Carbon Dioxide

COD = Chemical Oxygen Demand

CSTR = Continuously Stirred Reactor

DO = Dissolved Oxygen

DS = Dissolved Solids

EPA = Environmental Protection Agency

FAEC = Floyd Agricultural Energy Corporation

FID = Flame Ionization Detector

GAC = Granulated Activated Carbon

GC = Gas Chromatograph

gpd = gallons per day

gph = gallons per hour

H_2 = Hydrogen gas

HRT = Hydraulic Retention Time

L = Liters

mequ = milliequivalents

ml = milliliter

MLSS = Mixed Liquor Suspended Solids

mm = millimeter

NH_3-N = Ammonia

NPDES = National Pollution Discharge Elimination System

OLR = Organic Loading Rate

OP = Ortho-phosphate

Org-N = Organic Nitrogen
ortho-P = Ortho-phosphate
ppm = parts per million
PVC = Polyvinyl Chloride
R = Linear Regression Correlation Coefficient
SFV = Superficial Flow Velocity
SMA = Specific Methanogenic Activity
SVI = Sludge Volume Index
TCOD = Total COD
TKN = Total Kjeldahl Nitrogen
TP = Total Phosphate
TS = Total Solids
TSS = Total Suspended Solids
TVS = Total Volatile Solids
UASB = Upflow Anaerobic Sludge Bed reactor
V = Volume
VOA = Volatile Organic Acid
VS = Volatile Solids
VSS = Volatile Suspended Solids
VSWCB = Virginia State Water Control Board

VITAE

Jessi Lind Tisinger was born December 15, 1959 in Roanoke, Virginia. She attended Cave Spring High School graduating in 1978. She received a B.S. in Mechanical Engineering in 1983 from VPI & SU. While studying for her undergraduate degree she completed the cooperative education program with the U. S. Forest Service and the Smithsonian Institution.

Ms Tisinger spent 6 years working for Floyd Agricultural Energy Corp. in various capacities including wastewater treatment manager. She spent an internship in The Netherlands studying UASB technology at the Agricultural Institute of The Netherlands in Wageningen with Dr. Gatze Lettinga during the summer of 1986.

Currently Ms Tisinger is Research Engineer for Sybron Biochemicals, Inc. of Salem, Virginia.

Jessi Tisinger

TREATMENT OF CLAM-PROCESSING WASTEWATER USING UPFLOW ANAEROBIC SLUDGE BLANKET (UASB) TECHNOLOGY

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(Abstract)

The Upflow Anaerobic Sludge Bed reactor (UASB) has been used successfully by the food processing, pulp and paper and municipal wastewater industries. High organic strength wastewater, limited space, extended down times and sludge handling and disposal have been critical factors in waste treatment system selection.

This study investigated the performance of a laboratory scale UASB reactor for treating clam-processing wastewater. Virginia state effluent regulations for BOD_5:TSS ranged from 0:0 to 90:90 to as high as 500:500 depending on the facility location.

It was found that at a volumetric organic loading rate of 13.8 g COD/L/d the BOD_5 removal efficiency averaged 87% and TSS removal efficiency averaged 83%. The average effluent values for BOD_5 and TSS were 200 ppm and 90 ppm, respectively. The conversion efficiency of COD to methane gas was 81%. At an organic loading rate of 4.8 g COD/L/d the effluent values averaged 77 and 66 mg/L for BOD_5 and TSS, respectively.

Methane production was inhibited at Na^+ concentrations above 5,000 ppm in batch experiments. At Na^+ concentrations above 12,500 ppm inhibition was essentially 100%. Nutrient enrichment did not affect methane production.