To the Graduate Council:

I am submitting herewith a thesis written by Lara Jacqueline Beal entitled "Laboratory Scale Testing of an Anaerobic Waste Treatment System for a Confectionery Wastewater". I have examined the final copy of this thesis for form and content and recommend that it be accepted in partial fulfillment of the requirements for the degree of Master of Science, with a major in Biosystems Engineering.

Dr. D. Raj Raman, Major Professor

We have read this thesis and
recommend its acceptance

Dr. Ronald Yoder

Dr. Kevin Robinson

Accepted for the Council:

Associate Vice Chancellor and
Dean of The Graduate School

Laboratory Scale Testing of an Anaerobic Waste Treatment System for a Confectionery Wastewater

A Thesis

Presented for the

Master of Science

Degree

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Lara Jacqueline Beal

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To:

My parents Len and Judy Beal

for their ever present love and support throughout my education.

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ABSTRACT

Two bench scale anaerobic treatment systems were tested for their ability to reduce the COD concentration of confectionery wastewater. System 1 consisted of an Upflow Anaerobic Sludge Blanket (UASB) followed by a Downflow Anaerobic Filter (DFAF); System 2 consisted of two DFAFs in series. The UASB was operated at 35°C; all other reactors operated at ambient (ca. 25°C) temperature. Each system had a total hydraulic retention time (HRT) of 2.4 days. The wastewater was a mixture of two waste streams, and had a COD of 30 g L^{-1}.

System performance was characterized by chemical oxygen demand (COD) removal rates, pH, and biogas production rates. During the latter part of the study, volatile fatty acid content, alkalinity and biogas composition were monitored to help control reactor performance.

System 1 achieved 98% COD removal efficiency and operated consistently at the required organic loading rate of 12 kg COD m^{-3} d^{-1}. The UASB, in particular, showed a high degree of treatment with a consistent COD treatment efficiency of 95% while loaded at 19 kg COD m^{-3} d^{-1}. Solids retention time (SRT) in the UASB was estimated as 33 d, and the specific substrate utilization rate was approximately 0.67 g COD (g VSS d)$^{-1}$. The secondary reactor in System 1 effectively polished effluent COD to below 0.4 g L^{-1}. The System 2 primary reactor never achieved organic loadings greater than 6 kg COD m^{-3} d^{-1}.

The secondary DFAF reactors contained differing media types: brick cubes and plastic rings. A comparison could not be made between the media types.

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

CHAPTER I

INTRODUCTION

Economics and government regulations drive industrial facilities to search for increasingly efficient and cost effective techniques for on-site treatment of organic wastes. For certain wastes, anaerobic digestion may be a feasible alternative. McCarty (1982) defines anaerobic digestion as the use of biological processes in the absence of oxygen for the stabilization of organic materials by conversion to methane and inorganic end products.

High-rate anaerobic processes continue to gain popularity as feasible treatment alternatives for certain industrial wastewaters. Because these technologies increase the retention time of the slow-growing microorganisms involved in anaerobic processes, high-rate systems achieve increased efficiency and reliability compared to their standard-rate counterparts.

A confectionery processor in the southeastern US produces two waste streams that must be treated prior to discharge to a publicly owned treatment works, a high-strength low-flow stream and a relatively low-strength high-flow stream. On-site treatment of the low-strength stream consists of an aerobic equalization basin, a dissolved air floatation unit, and an activated sludge system. The system requires high energy and synthetic chemical inputs. Recent increases in waste production have caused the efficiency and reliability of the onsite process to decrease. The high-strength waste is land applied off-site by a contractor at a significant cost to the food processor.

Planned expansions to the processing plant require a treatment facility capable of handling the increased load. Also, the current system incurs large operational costs. A high-rate anaerobic digestion process could potentially reduce the energy and sludge disposal requirements below those required by the aerobic system. If properly designed, such a system could potentially treat both waste streams, which have a combined concentration of 30 g COD L^{-1}.

Two high rate reactors were considered as alternatives for treating the combined wastewater, an upflow anaerobic sludge blanket (UASB) and a downflow anaerobic filter (DFAF). Standard-rate anaerobic processes, such as lagoons, require long hydraulic retention times (HRT) to allow for the slow growth of the microorganisms involved. In standard-rate processes, the HRT and the solids retention time (SRT) are equal. In high-rate anaerobic processes, the HRT and the SRT are uncoupled, allowing the HRT to be greatly reduced while the SRT is maintained.

When compared to an aerobic process, high-rate anaerobic systems have several advantages and disadvantages. Speece (1996) gives the following advantages and disadvantages for anaerobic systems:

Advantages

1. reduction of excess biomass disposal costs
2. reduction of nitrogen and phosphorus supplementation costs
3. minimization of installation space
4. conservation of energy
5. elimination of off-gas air pollution

Disadvantages

1. increased start-up time for development of biomass
2. insufficient effluent quality for surface water discharge (in some cases)
3. low kinetic rates at low temperatures
4. no nitrification is possible.

The UASB was one of the proposed high-rate processes making-up part of the treatment system. The UASB reactor relies on the slow upward movement of wastewater through granulated sludge that settles in the lower portion of the reactor. The sludge bed, composed of granular consortiums of microorganisms with good settling characteristics, performs 80 - 90% of the COD removal, though it only occupies 30% of the reactor volume. Rising biogas and biomass particles filled with biogas create good mixing conditions within the reactor. A gas/solid separator at the top of the reactor prevents large amounts of biomass from leaving the reactor (Hwang & Hansen, 1992).

The DFAF was the second high-rate anaerobic process to be considered. Carbon degradation in the DFAF occurs as wastewater moves downward through the system contacting biomass growing on media placed within the reactor. In the DFAF, the HRT is uncoupled from the SRT via biomass attachment to the media.

Before implementing a full-scale anaerobic treatment system, treatability studies must be performed to determine the effectiveness of a process on a given wastewater. The objectives of this treatability study were as follows:

1. to compare the performance of a relatively high-technology reactor (UASB operated at 35°C) with a simpler one (DFAF operated at 25°C)

2. to understand the start-up, operation, and monitoring of high-rate anaerobic systems
3. to determine the potential of DFAF reactors as polishing reactors for primary reactors
4. to determine the effects of two different media types (plastic rings and clay cubes) in second stage DFAFs.

To address these issues, four lab scale reactors were operated for periods of time varying from five to ten months; reactor operation time depended on when each was brought on line during the study. Several water quality parameters were measured to monitor the reactors operations, including: chemical oxygen demand (COD), volatile fatty acid concentrations (VFA), total alkalinity, bicarbonate alkalinity, and pH. Biogas composition and production were also measured.

CHAPTER II

REVIEW OF LITERATURE

Anaerobic Process Microbiology

For complete degradation of organic matter to CO_2 and CH_4 during anaerobic digestion, an orchestrated consortium of metabolic activities occurs (Figure 1). Conversion of complex wastewater to methane requires three distinct biochemical steps; hydrolysis, fermentation and anaerobic oxidation, and methanogenesis. Digester feedstock and operation parameters influence the populations of bacteria present within the reactor, and therefore the concentration of various intermediates and end products.

Hydrolysis degrades complex polymers into smaller sub-units capable of being transported into a cell. It is catalyzed by extracellular enzymes secreted by the microorganism (Pavlostathis and Gomez, 1991). When carbohydrates, proteins, and lipids are hydrolyzed, simple sugars, amino acids, and long chain fatty acids result, respectively.

Simple sugars, amino acids, and long chain fatty acids are further degraded to propionate, butyrate, acetate, CO_2 and H_2 through fermentation and anaerobic oxidation. The short chain fatty acids (propionate and butyrate) are often referred to as volatile fatty acids (VFAs). Anaerobic oxidation also refers to the subsequent conversion of VFAs to acetate or CO_2 and H_2, after fermentation has occurred, for utilization by methane forming bacteria. Some CO_2 and H_2 may be converted to acetate via homoacetogenesis (Pavlostathis and Gomez, 1991).

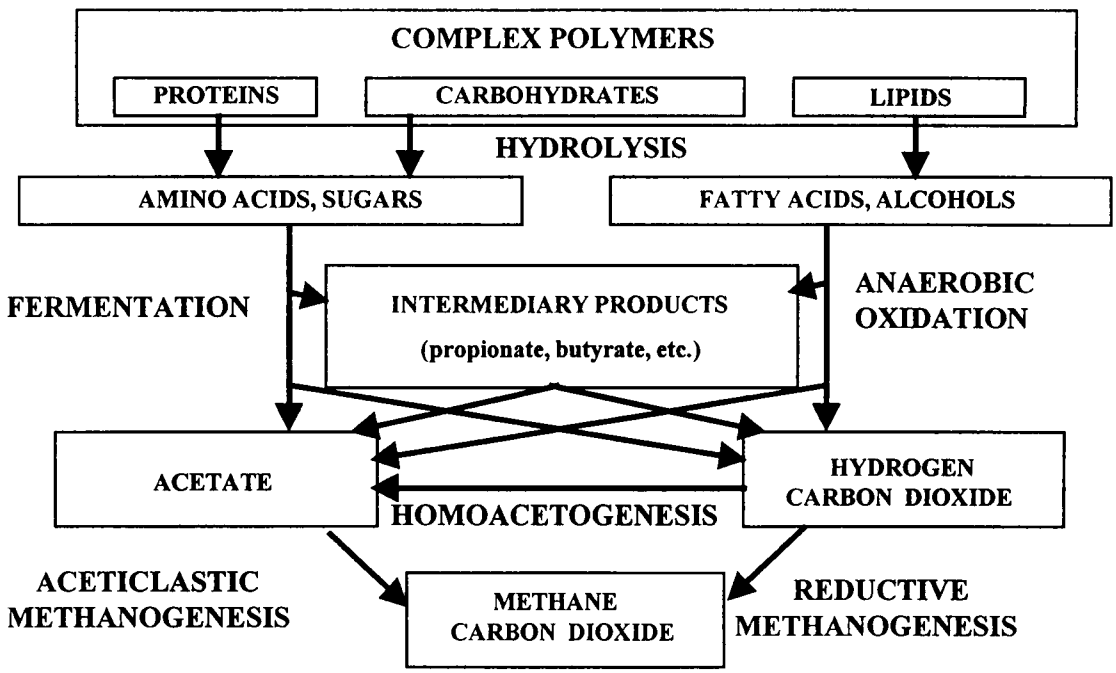


Figure 1. Biochemical pathways in the anaerobic digestion process, derived from Pavlostathis and Gomez (1991)

Methanogenesis is the final process occurring in anaerobic digestion. As a result, acetate, CO_2, and H_2 are converted to CH_4 and CO_2. Aceticlastic methanogenesis involves the decarboxylation of acetate to form CH_4. Reductive methanogenesis (hydrogen oxidation) is the formation of methane from CO_2 and H_2 (Pavlostathis and Gomez, 1991). Reductive methanogenesis is an important step in anaerobic digestion because it regulates the hydrogen producing reactions by controlling the partial pressure of hydrogen in the reactor (Gaudy and Gaudy, 1980).

A typical failure mode occurring in anaerobic reactors involves excess production of short chain fatty acids by the fermenting microorganisms. When the organic load to a reactor rises, the concentration of hydrolyzing and fermenting bacteria increases faster than the concentration of methanogenic bacteria. The result is an increased H^+ concentration, further slowing the methanogen's growth rate. A spiral effect ensues wherein the pH continues to decrease, causing reactor failure (Speece, 1996).

The H_2 concentration within the reactor also affects the process. Conversion of propionate to acetate and H_2, via anaerobic oxidation, is thermodynamically impossible unless the H_2 concentration is less than 10^{-4} atmospheres (100 ppm). Conversion of H_2 to methane is not thermodynamically possible unless the H_2 concentration is greater than 10^{-6} atmospheres (1 ppm). Therefore, for proper reactor function, the H_2 concentration should remain between 10^{-4} and 10^{-6} atmospheres. Single-phase anaerobic processes maintain an average H_2 concentration of 10^{-4} atmospheres (Speece, 1996). During overload conditions, Hawkes et al. (1994) increased H_2 concentrations above 10^{-4} atmospheres.

Removal Kinetics

The Monod equation describes microbial growth rates as follows:

$$\mu = \frac{\mu_{max} S}{K_s + S} - b \qquad (1)$$

where μ is the specific growth rate (hr^{-1}), μ_{max} is the maximum specific growth rate (hr^{-1}), S is the substrate concentration (mg COD L^{-1}), K_s is the half velocity constant (mg COD L^{-1}) and b is the decay rate (hr^{-1}). The Monod equation relates microbial growth rate to substrate affinity and concentration.

Microbial yield describes the rate of growth with respect to substrate consumption. Yield is determined as follows:

$$Y = \frac{dX}{dS} \qquad (2)$$

where X is the microbial concentration measured as volatile suspended solids (mg VSS L^{-1}); units for Y are mg VSS (mg COD)$^{-1}$. A UASB treating distilled sugar cane molasses maintained a yield of 0.17 mg VSS (mg COD)$^{-1}$ (Driessen et al., 1995). Cail and Barford (1985) achieved a yield of 0.13 mg VSS (mg COD)$^{-1}$ treating cane juice stillage in a UASB.

The specific substrate utilization rate (U) identifies the rate of substrate degradation per unit biomass. The U value is determined as follows (Pavlostathis and Gomez, 1991):

$$U = \left(\frac{1}{X}\right)\frac{dS}{dt} = \frac{\mu}{Y} = \frac{U_{max}S}{K_s + S} \qquad (3)$$

where U_{max} is the maximum specific substrate utilization rate (mg COD (mg VSS)$^{-1}$ d^{-1}); (Note that some authors represent U_{max} as k). The UASB studied by Driessen et al. (1995) digesting sugar cane molasses operated with an average U value of 0.86 mg COD (mg VSS)$^{-1}$ d^{-1}. A U value of 0.77 mg COD (mg VSS)$^{-1}$ d^{-1} was reported by Cail and Barford (1985) for a UASB treating cane juice stillage. An upflow anaerobic filter treating ice-cream waste water operated at a U of 0.28 mg COD (mg VSS)$^{-1}$ d^{-1} (Hawkes et al., 1995).

Temperature Effects

Growth kinetics of microorganisms in the anaerobic digestion process may be altered by temperature. In general, higher temperatures lead to higher microbial activity until an optimum temperature is reached. Increasing temperatures beyond the optimum results in a decrease in biological activity (Pavlostathis and Gomez, 1991). The Arrhenius equation is a widely used equation for relating microbial activity to temperature. It is as follows:

$$V = A\exp\left(\frac{-E_a}{RT}\right) \tag{4}$$

where V is a process rate (hr^{-1}), A is the frequency factor (hr^{-1}), E_a is the apparent activation energy (kcal mol^{-1}), R is the universal gas constant (0.001987 kcal mol^{-1} K^{-1}), and T is the absolute temperature (K). Hinshelwood (1946) modified the Arrhenius equation and proposed the following model to describe the effect of temperature on net microbial activity thus accounting for the two opposing processes, cell synthesis and cell degradation:

$$V = A_1\exp\left(\frac{-E_{a1}}{RT}\right) - A_2\exp\left(\frac{-E_{a2}}{RT}\right) \tag{5}$$

where $E_{a2} >> E_{a1}$ (subscripts 1 and 2 refer to synthesis and degradative processes, respectively). Using the Hinshelwood equation, Pavlostathis and Gomez (1991) showed how the Arrhenius equation could be used to model methanogenic activity. The equation developed is as follows:

$$\ln k = \ln k_1 + a_1(T - 30) \tag{6}$$

where k is the relative methanogenic activity and a is an energy constant. For low temperatures, T < 30°C, degradation processes are negligible. The modified Hinshelwood

equation (Figure 2) shows an exponential increase in biomass activity with increasing temperature; activity peaks at the optimum temperature and then abruptly declines with further temperature increases.

Anaerobic reactors generally operate in either the mesophilic (30 - 40°C) or the thermophilic temperature range (50 - 60°C). However, Stronach et al. (1986) concluded that operation in the thermophilic range was more sensitive to process variations and less stable than operation in the mesophilic range. To operate anaerobic reactors below the mesophilic range, such as 20°C, longer start-up times were required because microbial growth rates were slower (Stronach et al., 1986). In contrast to Pavlostathis and Gomez (1991), Lettinga (1978) stated that at temperatures below 20°C less effective degradation takes place because hydrolysis was retarded.

Mass Transfer Effects

Anaerobic filters and UASBs operate via the formation of consortiums of aggregated biofilms and granules to uncouple the HRT and the SRT. However, as the aggregate size increases, the surface area-to-volume ratio decreases and either the influx of substrates or the efflux of products can become the rate limiting step in treatment. For this reason, it is useful to estimate mass transfer limitations of the reactor.

The dimensionless Damkohler number (Da) describes the significance of mass transfer limitations. The Da is the ratio of maximum reaction velocity (V_{max}) to maximum flux rate (J_{max}) (Geankopolis, 1993). Pavlostathis and Gomez (1991) provide the following

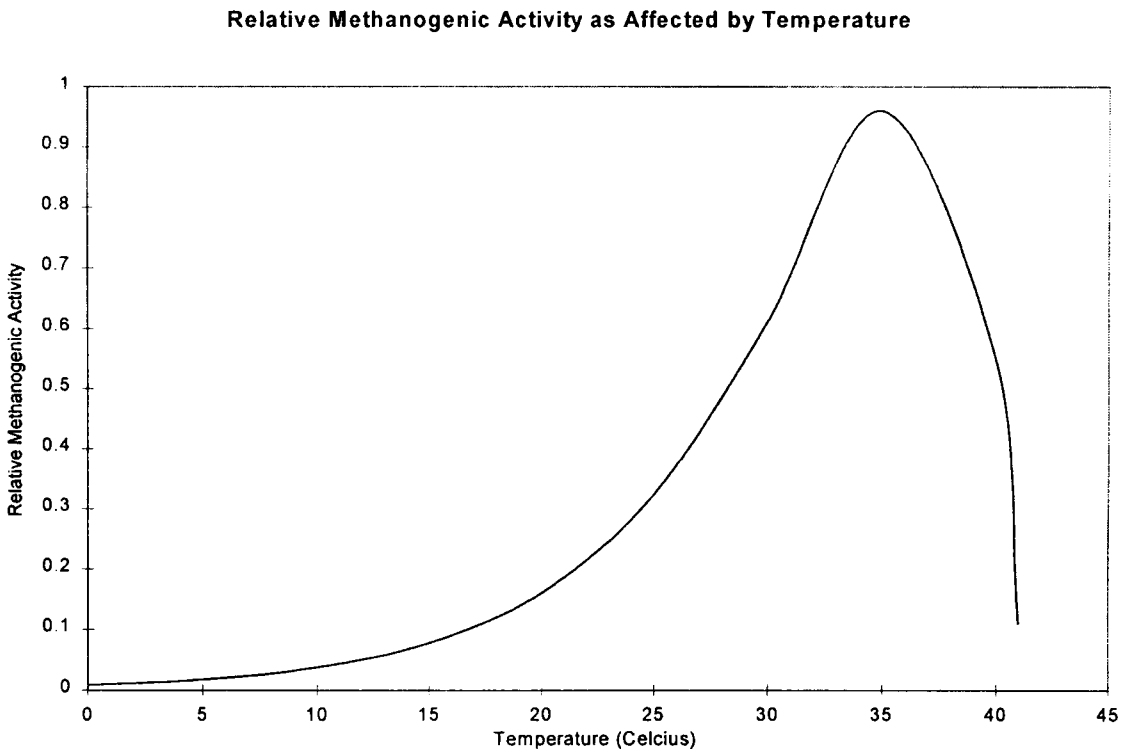


Figure 2. Graph of modified Hinshelwood equation.

equation, which is a modified Da, for describing the effects of external mass transfer limitations in UASB reactors:

$$Da = \frac{V''_{max}}{k_l K_s} \tag{7}$$

where V''_{max} is the maximum substrate uptake rate per unit of aggregate surface area, k_l is the mass transfer coefficient and K_s is the half velocity constant. A Da >> 1 describes a system whose rate of bioconversion is limited by diffusion. A Da ≈ 1 describes a system whose bioconversion rate and diffusion rate are comparable. A Da < 1 describes a desirable system where no diffusion limitations occur. Finally, a Da << 1 describes a system which is limited by its bioconversion rate.

Start-Up

As already described, effective anaerobic digestion requires a consortium of bacteria. The microorganisms present in anaerobic processes have differing generation times. Therefore, start-up of anaerobic systems is more complex than start-up of aerobic systems. Specifically, the anaerobic oxidizing bacteria and the methanogens have low growth rates compared with the hydrolyzing and fermenting bacteria (Wilkie and Colleran, 1988). When bacterial proportions are not correct (not enough acid utilizing microbes present), build-up of excess acid occurs in the system causing a pH decrease possibly resulting in reactor failure. Careful monitoring and proper start-up procedures can circumvent possible failures due to acid build-up.

Start-up of the UASB reactor requires the formation of granular sludge essential for a properly functioning reactor. Because of this, lengthy start-ups may be required. Granule formation results from complex physicochemical and microbial interactions. It has been suggested that formation is initiated by bacterial adhesion to inert matter or to one another through physicochemical interactions (Stronach et al., 1986). Extracellular polymers secreted by the bacteria form the initial granule by strengthening the loosely aggregated formations. The initial granule then grows larger to form a compact, mature granule (Tay and Yan, 1996). The process requires the maintenance of favorable conditions within the reactor. Granule formation occurs within many substrates, but it generally occurs in reactors treating carbohydrate rich or VFA type wastewaters (Cayless et al., 1989). Granule formation time varies. Driessen et al. (1995) reported 6-8 months for granulation of sludge in a full-scale UASB treating cane juice stillage inoculated with digested sewage sludge. However, granules were developed in a 3-L reactor seeded with bacteria from a anaerobic lagoon treating cheese-whey within 2 months (Kalyuzhni et al., 1997).

Inoculating the UASB reactor with granulated biomass from a preexisting source circumvents the initial start-up time required for granule formation. The range of inoculum volume required should be 10 – 30% of the reactor volume (Hickey et al., 1991). However, care should be taken to find granular sludge grown within characteristics similar to the substrate and environment of the reactor to be started. In an experiment performed on five different substrates (sucrose, acetate, casein, butanol, and ice cream) inoculated with a single origin granular sludge, Morgan et al. (1990)

concluded that biomass subjected to feedstock for which it was not acclimatized resulted in deterioration of sludge performance for an average of 20 days, at which time the granules slowly started to recover.

Start-up of an anaerobic filter requires formation of a biofilm on media within the reactor. When a clean inert surface is exposed to fluid flow containing organic material, nutrients, and microorganisms, an organic monolayer rapidly adsorbs to the surface conditioning it for microbe attachment. As microbes grow and proliferate, they produce extracellular polymeric enzymes that form a structural matrix referred to as a biofilm. The biofilm's density increases as the organic load increases (Stronach et al., 1986). Biological matrices form on a wide variety of materials including plastics, metals, and ceramics.

Time for biofilm formation in anaerobic reactors varies. An anaerobic filter treating distillery wastes was seeded with organisms from a pilot plant operated on-site (Szendrey, 1983). The reactor was on-line after four months. Kikkeri and Viraraghavan (1991) were able to develop a biofilm in reactors treating dairy wastewater after two months of operation; the 10-L reactors were seeded with digested sewage sludge. When these reactors were seeded with biomass from a similar type wastewater, a lengthy acclimation period was not required. However, if the wastewater source for the sludge differed, an acclimation period was required (Hickey et al., 1991). Acclimatization of sludge was similar in both reactor types. Monitoring of COD removal efficiencies and VFA production was important during this period. Van Den Berg and Lentz (1980) started their DFAF with a 25-d HRT to achieve a low organic loading rate. The loading

rate was increased in 5 – 25 % steps depending on the VFA production. The feed was stopped when the VFA concentration increased above 600 mg acetic acid L^{-1}; feed was restarted when the VFA concentration decreased below 400 mg acetic acid L^{-1}. An anaerobic filter treating distillery waste was batch loaded for 35-days and recirculated continuously (Szendrey, 1983). The organic load on the digester was increased when the COD removal efficiency stabilized. Cail and Barford (1985) started a UASB with a 28-day acclimatization period. Loading was then slowly increased. Feed was halted when VFA concentrations reached above 1000 mg L^{-1}; feed was increased when VFA concentrations were below 100 mg L^{-1}. Theses studies show that reactors can be effectively started by closely monitoring reactor performance during acclimization.

Based on conclusions previously drawn from other researchers, both of the primary reactors used in this project were inoculated to enhance start-up. The UASB received granulated sludge from a reactor treating carbohydrate wastewater, and the DFAF received anaerobic microorganisms concentrated on wheat germ. Initially, the reactors were lightly loaded. Increases in organic loading rate (OLR) occurred as COD treatment efficiency increased and VFA concentrations decreased.

Organic Loading and Removal Rates

Organic loading and removal rates provide a means for comparing reactor effectiveness between systems and can be used to roughly estimate reactor volumes. The OLR indicates the mass of organic carbon entering the reactor per unit volume per unit time (kg m^{-3} d^{-1}). The organic removal rate (ORR) measures the mass of organic carbon being degraded within the reactor per unit volume per unit time (kg m^{-3} d^{-1}). Loading and

degradation rates are a function of the influent and effluent substrate concentrations and the HRT, as follows:

$$OLR = \frac{S_{in}}{HRT} \tag{8}$$

$$ORR = \frac{(S_{in} - S_{out})}{HRT} \tag{9}$$

where S_{in} is the influent substrate concentration and S_{out} is the effluent substrate concentration. By substituting V/Q for HRT into the above calculations, the rates become:

$$OLR = \frac{S_{in}Q}{V} \tag{10}$$

$$ORR = \frac{(S_{in} - S_{out})Q}{V} \tag{11}$$

where Q is the influent flow rate (L d^{-1}) and V is the reactor volume (L). Influent concentration and flow rate are measurable parameters. Using data from previously constructed systems, the OLR can be estimated and used to approximate a reactor volume. During operation, the OLR can be controlled by varying flow rate or influent concentration. The loading rate and the removal rate can be used to determine the treatment efficiency (TE) as follows:

$$TE = \frac{ORR}{OLR} \times 100 \tag{12}$$

Stronach et al., (1986) suggest typical OLRs for UASBs and anaerobic filters treating various wastewaters. Loading rates for UASBs range between 10 – 30 kg COD m^{-3} d^{-1}; loading rates for anaerobic filters range between 1 – 40 kg COD m^{-3} d^{-1}. Table 1 provides a list of OLRs and COD TE for UASBs and DFAFs treating carbohydrate wastewaters at mesophilic temperatures. Generally, the UASB operates at higher OLRs than the DFAF.

Initially, the UASB was selected as the primary reactor for this study base on the e TE of other reactors with OLRs similar to those required by the confectionery plant. Use of the DFAF as a primary reactor came after observing the effective operation achieved by the DFAF as a secondary reactor. The reactors were sized using OLR information.

Table 1. Organic loading rates for UASB and DFAF reactors treating carbohydrate type wastewaters.

Reactor Type	Author	OLR kg COD m^{-3} d^{-1}	COD conc. mg L^{-1}	Treatment Efficiency	Wastewater
UASB	Cail & Barford, 1985	22.5	27,000	>70%	Cane juice stillage
UASB	Driessen et al., 1995	22	22,000	88%	Sugar cane juice
UASB	Driessen et al., 1995	10	50,000	70%	Sugar cane molasses
DFAF	Hawkes et al., 1995	6	5,000	67%	Ice cream wastes
DFAF	Jhung & Choi, 1995	12.3	30,000	64%	Synthetic
DFAF	Young, 1983	13	130,000	-----	Rum distillate

Reactor Monitoring

To achieve a consistently high degree of waste stabilization and a high conversion of waste to methane, the optimal microbial populations must be maintained. An adequate monitoring system for an anaerobic treatment process combines on-line and off-line measurements, and requires an understanding of the mechanisms controlling process behavior (Hickey, 1991). Common measurements used to monitor anaerobic treatment processes include: 1) VFA production, 2) total and bicarbonate alkalinity, 3) pH, 4) COD reductions, and 5) biogas composition and production (Stover, 1998). Combined, these measurements help achieve stable reactor operation.

An increase in volatile fatty acid concentration can cause reactor instability if adequate system buffering is not available. Large increases in VFA concentrations are indicative of methanogenic population failure due to shock loadings or nutrient depletion (Stronach et al., 1986). As explained earlier, system overloads create high VFA concentrations because hydrolyzing and fermenting bacteria rapidly increase with increased substrate, but the acetotrophic bacteria do not. Thus, excess volatile acids are produced, which tend to consume alkalinity and ultimately decrease the pH. A decrease in pH can lead to further reductions in methanogenesis, and to costly system failures. Well operating anaerobic digestion processes maintain VFA concentrations between 100 and 300 mg acetic acid L^{-1} (Speece, 1996).

Because all correctly operating systems contain VFAs, they must be properly buffered. Alkalinity is a measure of the solution's capacity to neutralize acids; it establishes the system's buffering capacity. Total alkalinity includes the bicarbonate

alkalinity and the alkalinity created by the salts of VFAs. However, though the salts of volatile fatty acids may constitute a major fraction of the total alkalinity, they are not available for the neutralization of additional acids (Speece, 1996). For this reason, the bicarbonate alkalinity of a system should also be measured when determining the system's buffering capacity. Bicarbonate alkalinity should be at least 1000 mg $CaCO_3$ L^{-1} for successful operation, and it should always be greater then the volatile fatty acid concentration (Hawkes et al, 1993).

The pH expresses the H^+ concentration. However, to be of use for process control the data must be combined with the volatile fatty acid and the alkalinity measurements. The methanogens, necessary for anaerobic digestion, prefer nearly neutral conditions; thus, the optimum pH range in anaerobic reactors is from 6.5 to 8.2; conditions outside of this range cause a steep decline in the rate of methane production (Speece, 1996).

Monitoring carbon removal determines the process's treatment efficiency and organic removal rate. Chemical oxygen demand is often used to measure carbon removal in anaerobic reactors because it is a more rapid and precise test than 5-day biochemical oxygen demand (BOD_5) (Szendrey, 1983). Anaerobic systems perform under a wide range of removal efficiencies dependent on the biodegradability of the wastewater and the organic loading rate. Van Den Berg and Lentz (1980) achieved 90% removal of COD in a DFAF treating bean blanching wastewater. A full-scale anaerobic filter being fed distillery effluent operated at an average 82% efficiency in its first year of operation (Szendrey, 1983). Cail and Barford (1985) achieved removal in excess of 95 % soluble COD while operating a 10-L UASB.

Biogas generated in the reactor should be monitored for its production rate and its methane composition. The CH_4 production rate provides an estimation of COD removal and energy conversion. The stochiometric maximum methane production rate is 0.35 L CH_4 (g COD)$^{-1}$ (Pavlostathis and Gomez, 1991). Hawkes et al., (1995) recovered 0.27 L CH_4 (g COD)$^{-1}$ in a DFAF treating ice cream wastewater. A yield of 0.32 L CH_4 (g COD)$^{-1}$ was achieved in a UASB treating cane juice stillage (Cail and Barford, 1985).

Initially, reactor performance was monitored via COD removal, pH, and biogas production. After a reactor failure and further research about anaerobic reactor monitoring, VFA concentration and alkalinity measurements were performed, and biogas composition was also monitored.

CHAPTER III

MATERIALS AND METHODS

Wastewater Composition

Effluent from the confectionery processing plant in this study consisted mainly of carbohydrates existing in the form of sucrose, dextrose, fructose, lactose, and starch. The plant's effluent pH varied from 4 - 6. The low-strength wastewater currently being treated on-site maintained a flow around 340,000 L d^{-1} with a COD ranging from 10 – 20 g L^{-1}. The high-strength wastewater being treated off-site via land application flowed from the plant at approximately 26,000 L d^{-1} with a COD concentration around 250 g L^{-1}. When the two streams were combined, the COD concentration averaged 30 g L^{-1} and the flow rate became 366,000 L d^{-1}.

Because the wastewater mainly consisted of carbohydrates, nutrients must be added to achieve microbial growth. Nitrogen and phosphorus were added to the wastewater in the form of ammonium nitrate and diammonium phosphate at a COD:N:P ratio of 200:4:1(Monroy et al., 1995; Hawkes et al., 1995). To increase the alkalinity of the wastewater, sodium bicarbonate was added as required.

Reactor Set-up

Two anaerobic systems consisting of four reactors were built and operated for the study (Figure 3). Table 2 lists reactor volume, HRT, and fill medium for both systems. System 1 consisted of a UASB reactor operating at 35°C followed by a DFAF operated

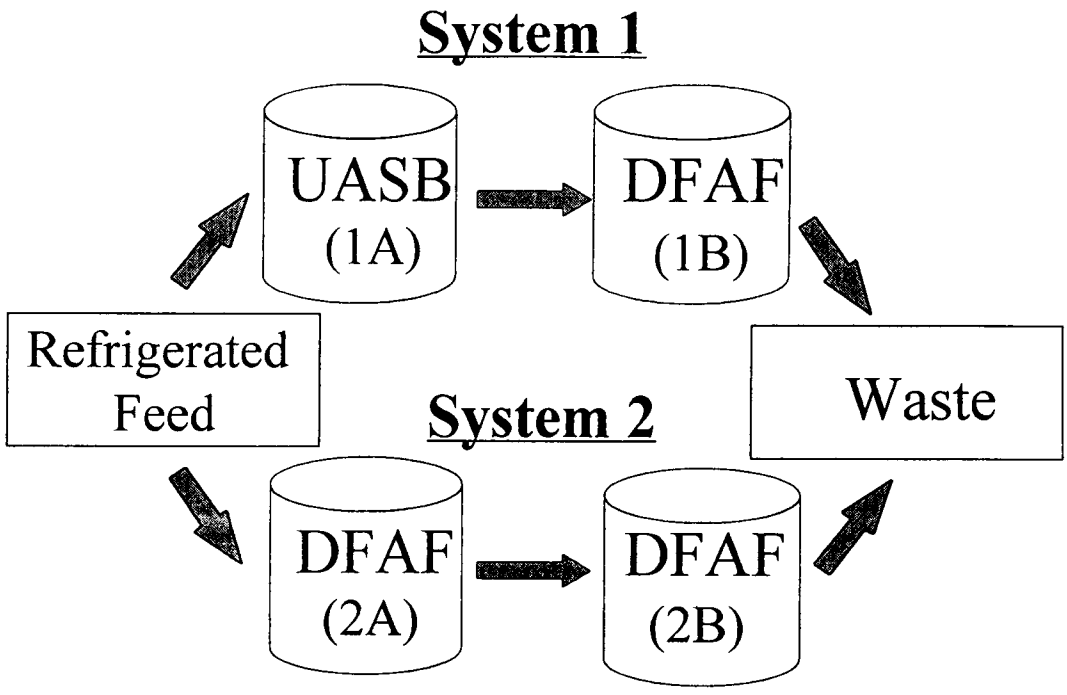


Figure 3. Overview of experimental set-up.

Table 2. Information about the reactors used in the study.

	Reactor Volume (L)	Working Volume (L)	HRT (d)	Fill Medium
System 1				
UASB-1A	24 (20*)	24 (20*)	2 (1.6**)	----------
DFAF-1B	25	10	0.8	25 cm^2 brick pieces
System 2				
DFAF-2A	40	20	1.6	25 cm^2 brick pieces
DFAF-2B	25	10	0.8	Plastic rings

* Due to reactor clogging, the volume of the UASB was reduced to 20 L in October of 1997.
** The HRT was reduced to 1.6 days when the volume of the reactor was reduced.

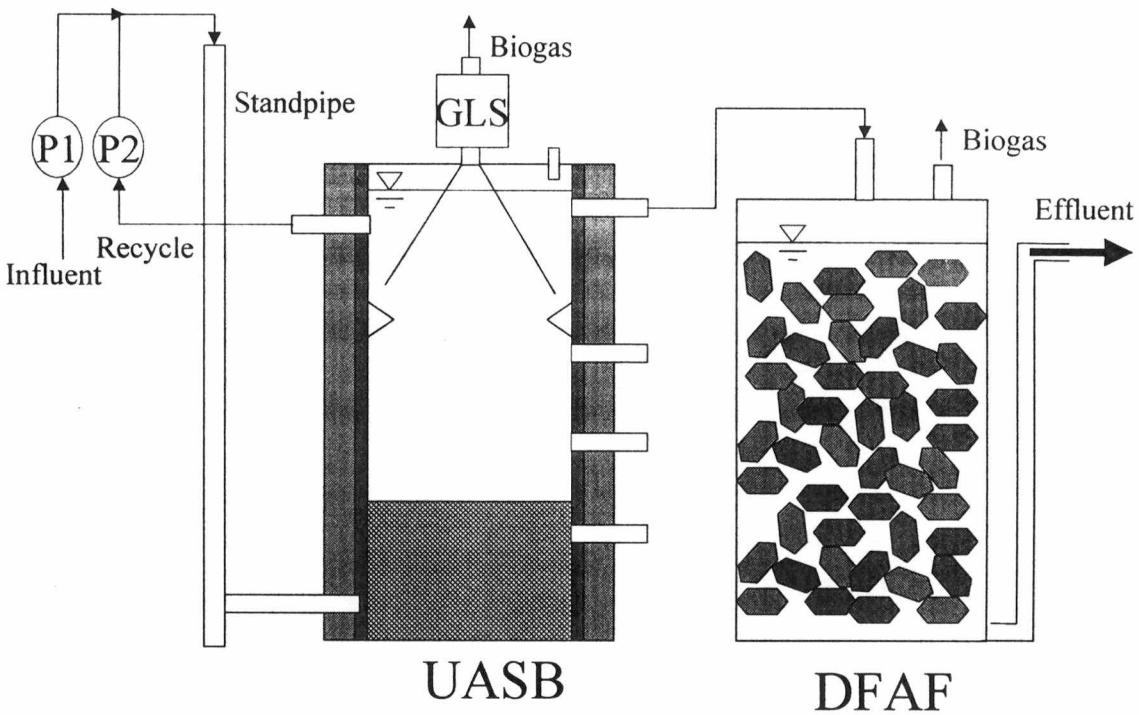
at room temperature (varying from 22 – 25°C). System 2 consisted of two DFAFs at room temperature.

Substrate for the reactors was retrieved from the confectionery processing plant as needed. The high-strength and the low-strength stream were stored separately until mixed. The streams were mixed at the same ratio in which they flow from the plant, 9-L low-strength : 1-L high-strength. The nutrients (N & P) and the sodium bicarbonate were dissolved into solution and added to the feed mixture. When necessary, the solution was diluted with tap water to achieve the appropriate organic loading rate.

The reactor influent was stored at 5°C to prevent pre-digestion of the substrate. A peristaltic pump (Masterflex Model 7553-80) transported wastewater through tubing (L/S 16 Tygon and L/S 16 Pharmed) to the standpipe on the UASB and to the DFAF in parallel with it; secondary reactors were gravity fed with effluent from the primary reactors. The distance between the refrigerated waste and the reactors allowed the tubing to act as a heat exchanger, bringing the substrate up to ambient temperature prior to entering the reactors. The UASB had a recycle line which was controlled with a peristaltic pump (Masterflex, Model 7553-70).

System 1

The UASB in System 1 was built of 9.5-mm thick clear acrylic tubing creating an initial volume of 24 L (reactor diameter-0.203 m , reactor height-0.762 m). Figure 4 provides a drawing of the UASB. The reactor had eight ports. The two uppermost ports located on the side of the reactor) were used for effluent and recirculation. The lowest port (located on the side of the reactor) was used for loading. Two ports located on the



(a) (b)

Figure 4. The reactors in System 1 are: (a) the UASB-1A and (b) the DFAF-1B. P1 was a peristaltic pump (Masterflex, Model 7553-80) used for influent and P2 was a peristaltic pump (Masterflex, Model 7553-70) used for recycle. The GLS (gas liquid separator) was an expansion area used to prevent blockage of the gas line.

top of the reactor were used for gas collection, one of which was also the entrance point for a thermocouple. One of them was the main gas collection port; it was located above the gas/solid separation device. A baffle in the reactor directed gas towards an inverted funnel for measurement. The other gas port was on the outside of the gas/solid separation device to remove trapped biogas that missed the separator. An expanded fitting above the main gas collection port (the GLS) prevented the gas line from clogging when large bubbles of gas were emitted from the sludge bed bringing with them granulated bacteria that could prevent gas flow. The remaining ports allowed for sampling the reactors contents.

Initially, the reactor contained a diffusion plate at the bottom of the reactor to prevent channeling of the flow; however, the plate often clogged and was removed two months after the reactor was brought on-line. The diffuser was removed by cutting 8 cm off the bottom of the reactor, effectively reducing the reactor volume to 20 L. A stand-pipe was placed next to the reactor to monitor blockage formation in the influent tubing and to divert flow if a blockage occured. Blockage formations were indicated when the liquid level in the stand-pipe rose above the normal operating level. A recycle line was added to the reactor in the later part of the study to dilute the incoming substrate to concentrations below 10 g COD L^{-1}. John Owens (1997) at the University of Florida, Gainsville, indicated that the sludge bed should not receive a concentration above 10 g COD L^{-1} and that effluent recirculation should be implemented to maintain these conditions at increased loadings.

Figure 5 shows the liquid levels at various points in the system that must be maintained to insure proper overflow from the UASB to the DFAF. Initially, problems arose because the gas outlet to the tipping bucket gas meter from the UASB was submerged under 12.7 cm (5 in) of water, which pressurized the system. To prevent drainage of the reactor to below the effluent port and subsequent escape of unmeasured biogas, the UASB effluent tube had to be 12.7 cm higher than the reactor liquid level.

The DFAF in System 1 was constructed of schedule 40 PVC pipe creating a volume of 25 L (reactor diameter-0.254 m, reactor height-0.508 m). A drawing of the DFAF was shown in Figure 4. It had three ports. One port on the top was for influent and the other top port was for gas release. The port at the bottom of the reactor was for effluent. The DFAF was filled with broken clay bricks (specific surface area of 670 cm^2 $(L\ bulk\ volume)^{-1}$) giving a void volume of 10 L. Clay bricks were selected as the packing for the reactor because of their ready availability, abrasive surface, and iron content.

System 2

The primary DFAF in System 2 had the same working volume (20 L) and retention time as the UASB parallel to it in System 1. Because it was packed with clay bricks (specific surface area of 670 cm^2 $(L\ bulk\ volume)^{-1}$), it was twice the external size of the UASB. The reactor was constructed of schedule 40 PVC (reactor diameter-0.254 m, reactor height-1.0 m). Figure 6 shows a drawing of the primary DFAF. The filter had six ports. Influent entered through one of the top ports, and effluent was removed through the lowest port. A second port at the top of the reactor allowed gas measurement. There

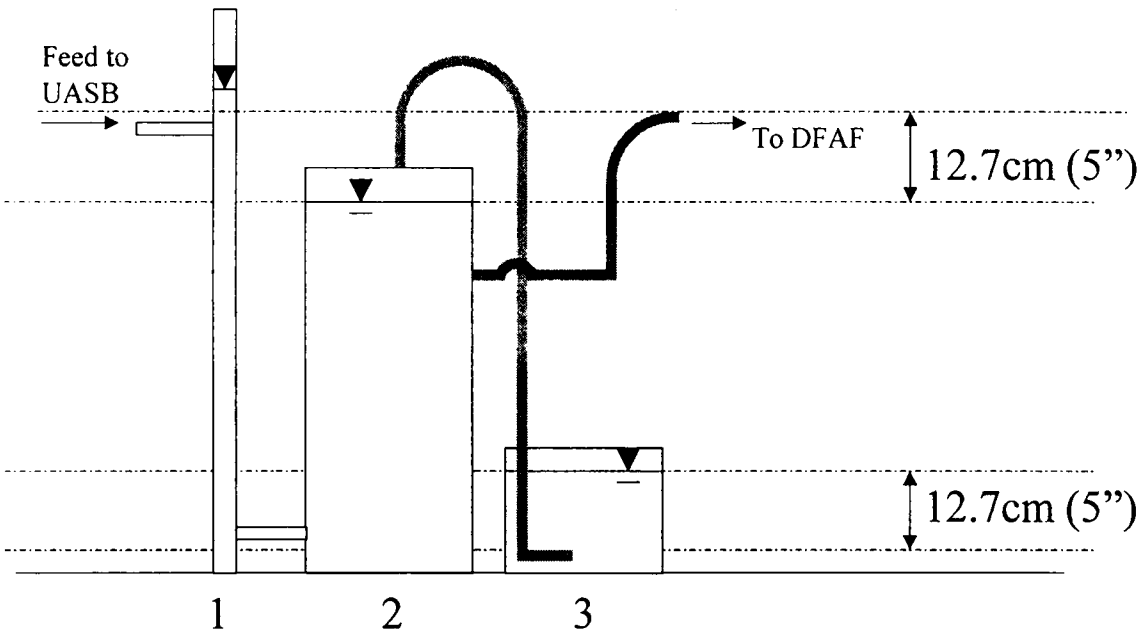


Figure 5. Pressure exerted on the UASB by the gas meter. It must be overcome for the system to operate properly. The standpipe (1) was open to the atmosphere, its liquid level was determined by the UASB effluent tube's height. The UASB (2) was pressurized by the height of water above the gas outlet line for the tipping bucket gas meter (3) submerged under 12.7 cm of water. The UASB's effluent tube must be 12.7 cm higher than the reactor's liquid level to allow proper overflow.

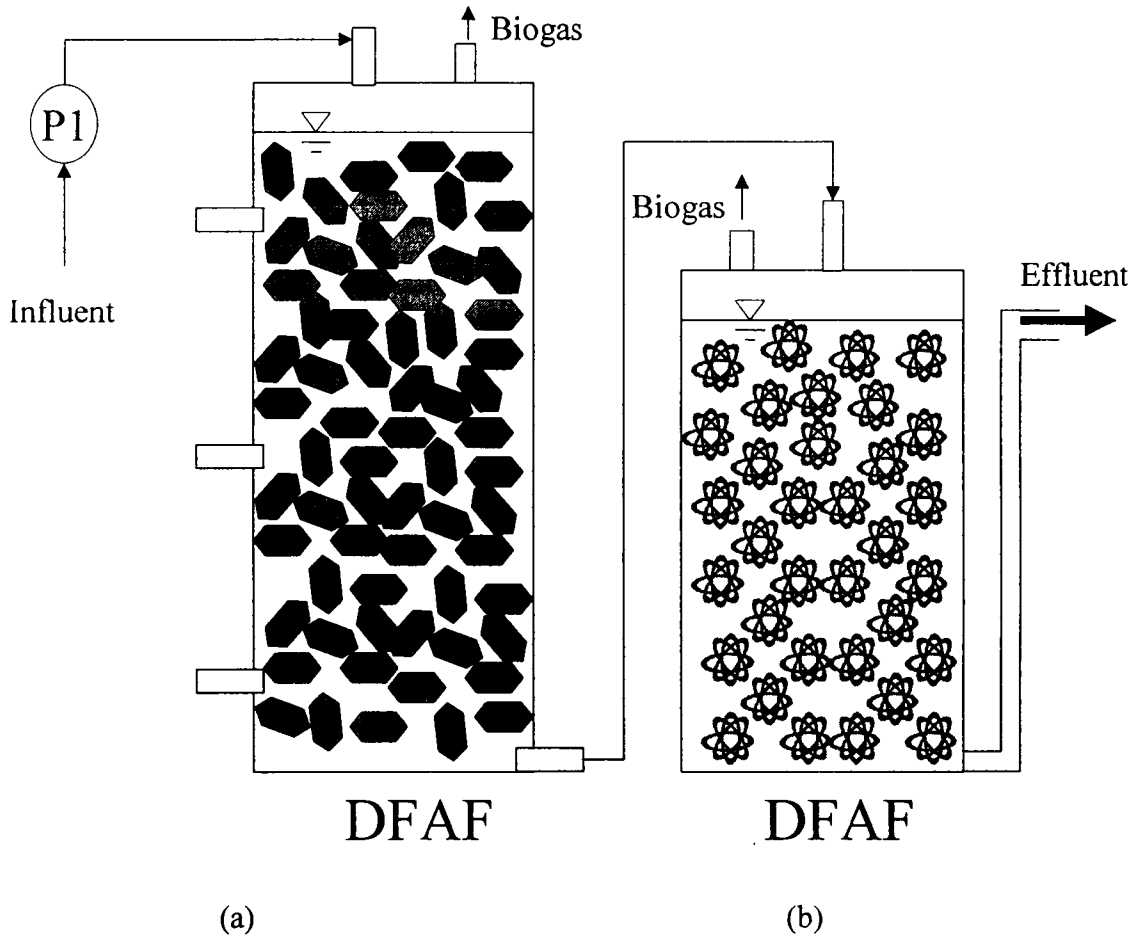


Figure 6. The reactors in System 2: (a) the DFAF-2A and (b) the DFAF-2B.
P1 was a peristaltic pump (Masterflex, Model 7553-80) used to load influent to the reactor.

were three ports along the side of the reactor used for collecting samples.

The secondary DFAF in System 2 was identical to the secondary DFAF in System 1 except the media consisted of plastic rings (specific surface area of 1050 cm^2 (L bulk volume)$^{-1}$). Figure 6 shows a drawing of the reactor.

Reactor Control

A Campbell Scientific 21 X Micrologger was used to monitor and control the reactors (Campbell Scientific, Inc., 1992). Appendix A provides the program used for collecting data. The 21X maintained the UASB's temperature at 35°C. It also monitored the DFAFs' temperatures. Appendix B provides a description of the circuit used to control the 50 W (0.5 A) heating blanket covering the UASB. The biogas metering device (described later) used a magnetic switch connected to the datalogger enabling determination of biogas production rates.

Reactor Operation

The four reactors were brought on-line at various times throughout the 16 months of this project. Early in the study, a UASB was the only reactor being used for treatment. As its operation stabilized, the additional reactors in the study were eventually brought on-line. Because of the time required to start-up and debug the reactors, System 2 was operated for only four months. Table 3 provides a chronology of the reactor operations.

<u>System 1: UASB-1A</u>

The UASB reactor was initially constructed of 0.25 m Schedule 40 PVC. The reactor was not seeded with granulated bacteria, rather it was seeded with a combination of anaerobic dairy lagoon sludge and anaerobic sewage sludge. An attempt was made to

Table 3. Experimental timeline.

	SYSTEM 1		SYSTEM 2	
	UASB - 1A	DFAF - 1B	DFAF - 2A	DFAF - 2B
June 97		Start - up		
July 97		↓		
Aug. 97				
Sept. 97	Start - up	Restart after no loading		
Oct. 97	↓	↓		
Nov. 97				
Dec. 97	Failure			
Jan. 98	Restart after failure			
Feb. 98	↓	↓	Start-up	
Mar. 98			↓	Start-up
Apr. 98			Failure followed by restart	↓
May 98			↓	
June 98				

grow granules at 25°C. However, several problems with this reactor prevented granule growth. Start-up for this UASB always included batch loading for a period of 1 – 2 weeks followed by continuous loading once treatment began to occur. During five months of operational experience with this reactor, the following problems were encountered:

1. clogging of tubing at hose-barb/tubing interface
2. inadvertent pressurizing of gas due to hydraulic problems
3. low reactor liquid levels compounded by an inability to monitor those levels
4. lack of operator expertise regarding system biochemistry.

Due to these problems, several changes were made. In May 1997, a new UASB reactor was constructed and started. This reactor preformed much better than the prototype and was in use 16 months later. The new reactor was built of clear acrylic to enable monitoring of the hydraulic problems that had occurred in the previous system. During start-up, growth of granulated bacteria was again attempted at 25°C; the same seed sludge was used. During the two months that followed, the hydraulic problems that plagued the first reactor were worked out. However, advances in the system's treatment ability were still hampered by lack of system biochemistry monitoring and granule formation.

In September 1997, the UASB was restarted and inoculated with granular sludge from Richard Speece's reactor at Vanderbilt University in Nashville, Tennessee. The reactor's temperature was increased to 35°C. Eight liters (33% of the reactor's volume) of granular sludge were added to the reactor to expedite start-up. During this stage of the

study, only pH, COD, and biogas production were being monitored. The reactor was initially batch loaded with substrate containing a COD of 100 mg L^{-1}. Within a week of granulated sludge inoculation, 50% carbon removal began occurring in the reactor. Because of this, continuous loading started at an OLR of 0.5 kg COD m^{-3} d^{-1}. Over the course of 2.5 months, OLR was gradually increased; 30% step increases in influent concentration were supplied when treatment efficiency recovered from the previous increase in loading. The OLR was increased by raising the substrate concentration; the HRT remained unchanged. Figure 7 shows the organic loading rate and substrate concentrations during the September start-up. By December 1997, the reactor operated at 82 % efficiency with an OLR of 5 kg COD m^{-3} d^{-1} (10,000 mg COD L^{-1}). However, by mid-December the pH dropped to 5.2 and the reactor was not able to recover. When measuring pH alone, the VFA concentration climbed to high levels before reactor overloading was noticed. Combining pH with VFA concentration and alkalinity measurements provided better information than measuring pH alone.

In January 1998, the UASB underwent restart with 2 L of granulated bacteria from a brewery added to the granules that existed in the reactor after the previous system failure. A batch load was placed in the reactor with substrate containing a COD of 1500 mg L^{-1}. After one week, 50% removal occurred in the reactor. Continuous loading was started at an OLR of 0.8 kg COD m^{-3} d^{-1}. Figure 8 represents the influent organic loading rate during start-up. At this time in the study, VFAs, alkalinity, pH, COD, and biogas composition and production were analyzed at two- to three-day intervals. The reactor went through complete start-up within three months via 30% step increases in loadings

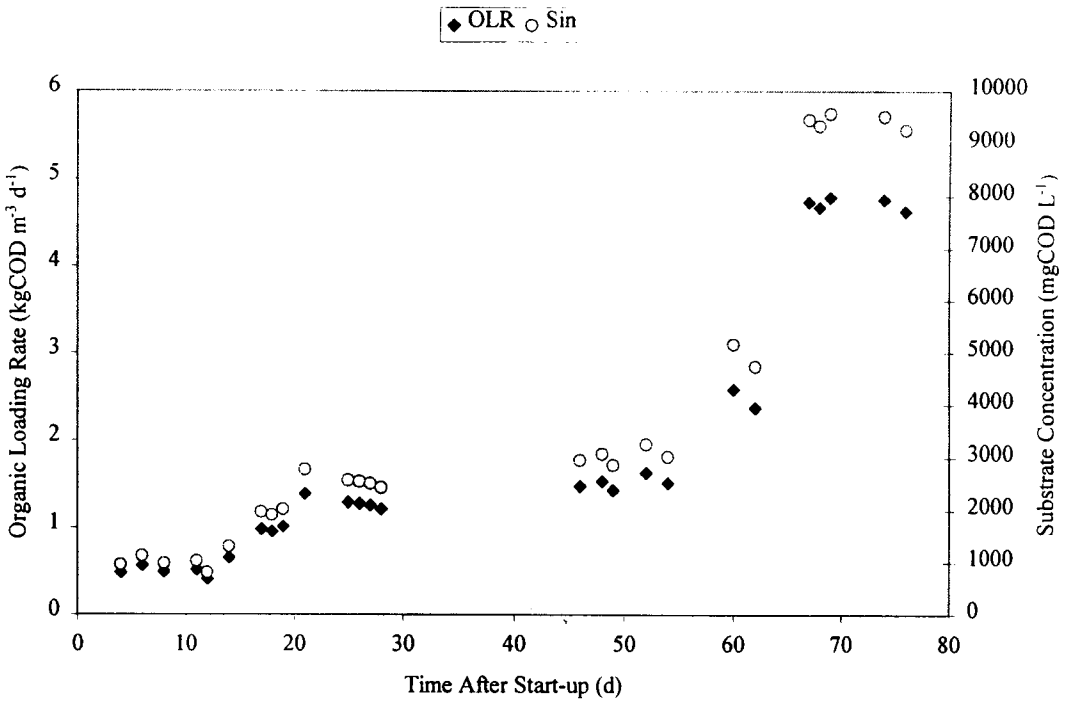


Figure 7. The OLR and substrate concentration in UASB-1A during the September 1997 start-up.

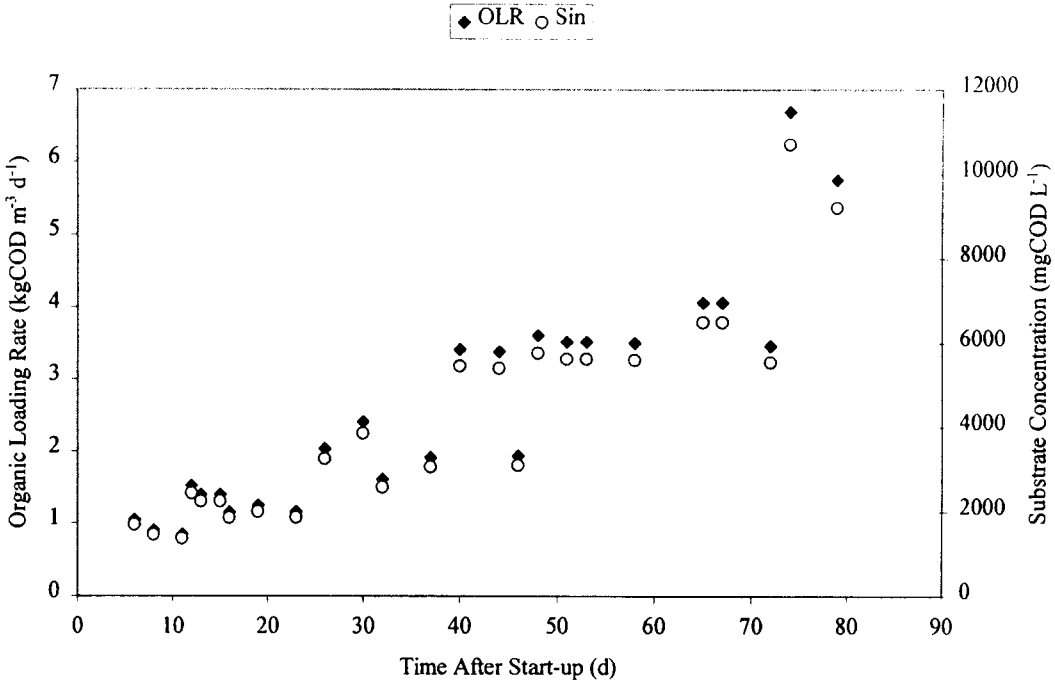


Figure 8. The OLR and substrate concentration in UASB-1A during the January 1998 start-up.

During the last start-up, sodium bicarbonate January 1998 start-up. was added to the reactor as needed when VFA concentrations exceeded 500 mg HAc L^{-1}.

System 1: DFAF-1B

The DFAF downstream of the UASB was started in June 1997. The reactor's initial fill was effluent from the UASB. Microorganisms in the UASB effluent provided the reactor's only inoculum. After two weeks in batch mode, the DFAF was continuously loaded with UASB effluent. Throughout the following months, the COD treatment efficiency increased to 60% (the DFAF performance was not well documented prior to September 1997; it was a summer intern project and not yet part of the study). However, due to its performance at ambient conditions as a secondary reactor, it became part of the study. During August 1997, the DFAF was not fed because the UASB was temporarily shut down. When the UASB was restarted with Speece's granules, the DFAF was restarted. Figure 9 represents the OLR in DFAF-1B during the September restart. It took three weeks for the treatment efficiency of the filter to return to 60%. When the UASB failed in mid-December, the loading to DFAF-1B increased suddenly. Except for the month after UASB failure, the DFAF OLR ranged from 0.3 – 4.0 kg COD m^{-3} d^{-1}. During the month after failure, OLR was 9 kg COD m^{-3} d^{-1}.

System 2: DFAF-2A

The primary DFAF in System 2 was started February 10, 1998. A decision was made to operate the reactor at 25°C based on the performance of the secondary reactor in System 1. Anaerobic microorganisms concentrated on dehydrated wheat germ *(Micro-Pro-AB)* from Microbe Masters (11800 Industriplex Blvd., Suite 4, Baton Rouge, LA 70809)

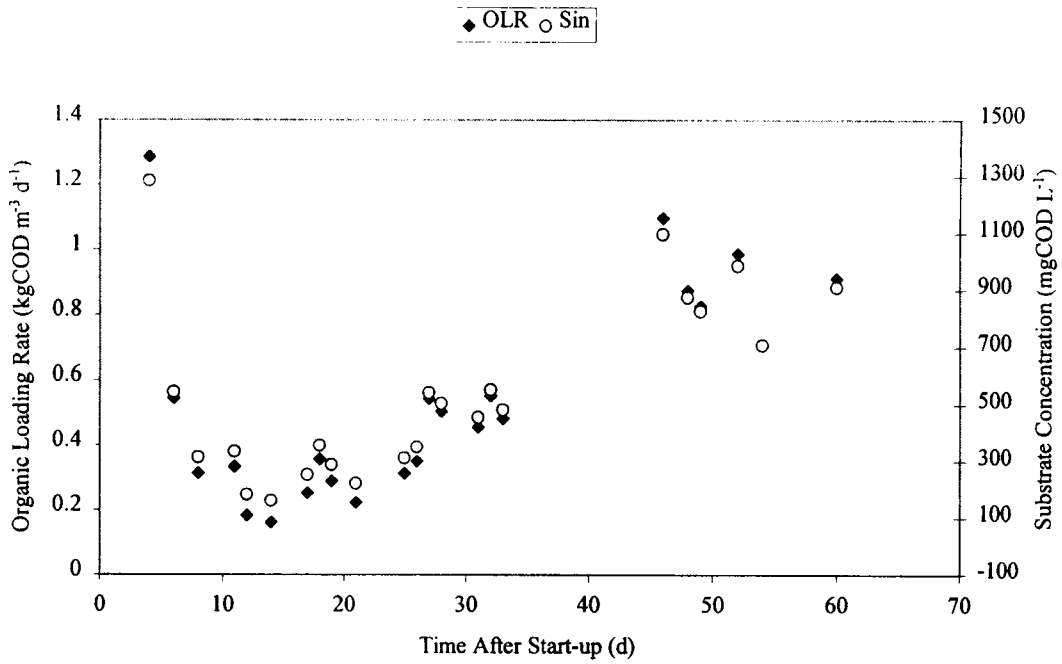


Figure 9. The OLR and substrate concentration in DFAF-1B during the September 1997 restart.

provided the inoculum for the reactor. A low strength waste with a concentration of 1000 mg COD L^{-1} was placed in the reactor. The batch load remained in the reactor for three weeks and underwent intermittent recirculation to disperse the bacteria throughout the system. Continuous loading into the reactor began on March 4, 1998 with a loading rate of 3 kg COD m^{-3} d^{-1}. The reactor recieved 30% step increases in OLR after each acclimatization period. Figure 10 shows the organic loading and substrate concentration increases during the February start-up. In mid-April, DFAF-2A failed. Two weeks prior to failure, the VFA concentration in the reactor increased above 1,000 mg acetic acid L^{-1} (HAc L^{-1}). As the concentrations increased above the level that could be handled by the system's alkalinity, failure occurred. Restart was begun immediately without the addition of new incoculum.

System 2:DFAF-2B

The secondary reactor in System 2 was started using the same method as was used for the secondary reactor in System 1. It was filled with effluent from the primary DFAF in March, 1998 and batch loaded for two weeks. When continuous loading began, the OLR was 2 kg COD m^{-3} d^{-1}. Figure 11 shows the organic loading and substrate concentration increases during the March start-up.

Reactor Monitoring

Sampling of reactor effluent generally occurred thrice weekly. When samples were collected, sufficient liquid volumes were drawn so that little head space remained in the sample container; the sample was immediately processed, or covered and refrigerated for storage. These precautions were performed to keep samples anaerobic and to prevent

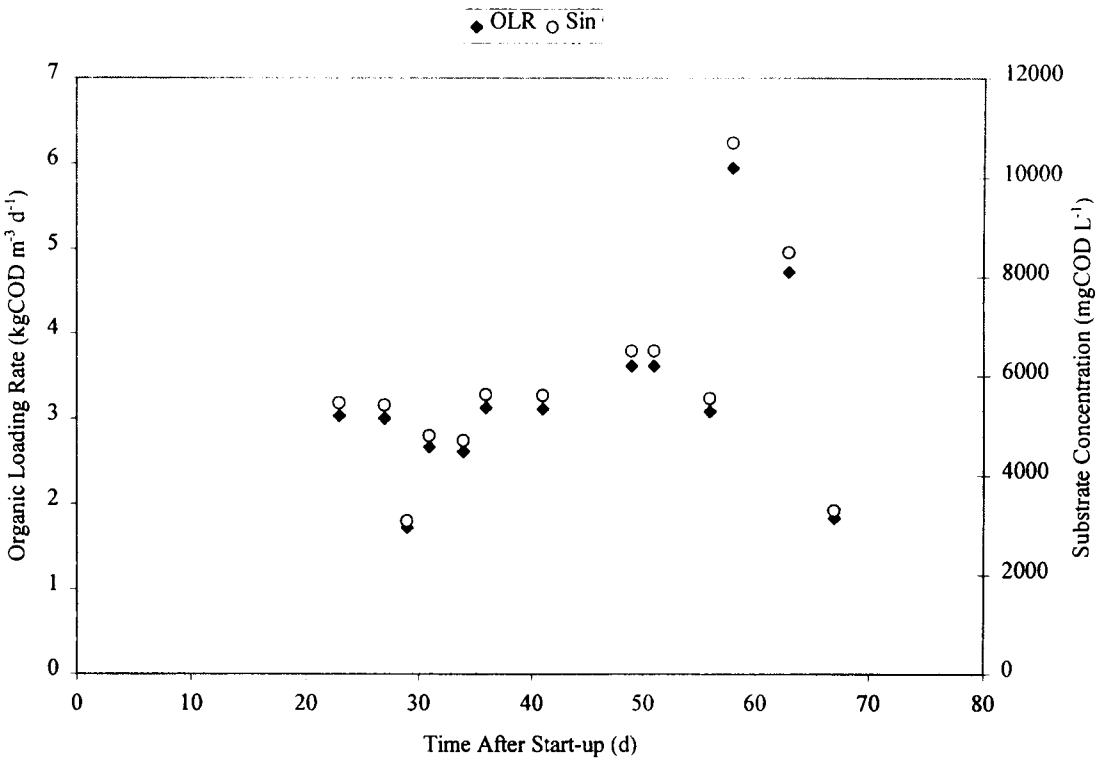


Figure 10. The OLR and substrate concentration in DFAF-2A during the February 1998 start-up.

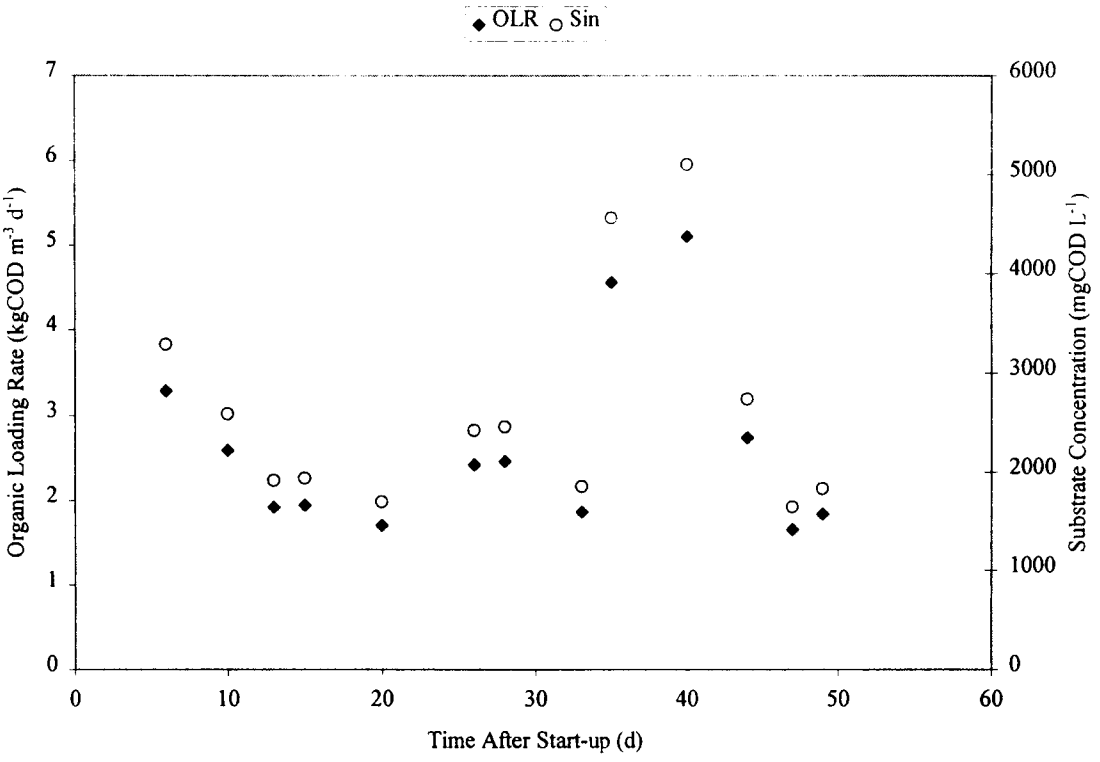


Figure 11. The OLR and substrate concentration in DFAF-2B during the March 1998 start-up.

further degradation. Prior to January 1998, before the first major UASB failure occurred, the following measurements on the reactor were made; pH, COD, and gas production. However, when the UASB was restarted for the final time, the following additional measurements were included; VFA concentration, total and bicarbonate alkalinity, and biogas composition. With the exception of biogas composition and production, all analysis methods were USEPA approved (Greenberg et al., 1992; Hach Company, 1996).

Before January 1998

When determining reactor pH, the sample was drawn and tested immediately without stirring to prevent loss of CO_2 from the sample. Loss of CO_2 from the sample yielded artificially elevated pH levels. The pH was determined via Standard Method 4500^{+}H–B (Greenberg et al., 1992). An Orion low maintenance pH triode (Model 9107BN) was used in conjunction with an Orion pH combination meter (Model 290A) for the measurement.

When samples were collected for determination of COD, they were allowed to settle for 5 minutes so as not to include granulated microorganisms in the assessment. Supernate off the settled sample was used for the test. The Hach high-rate COD testing procedure was used. In some cases, dilutions had to be performed to achieve the applicable range of 0 – 1500 mg COD L^{-1}. Samples were digested in the Hach COD Reactor and analyzed on the Hach DR/2000 Spectrophotometer (Hach Company, 1996).

Two tipping bucket gas meters, as shown in Figure 12, were constructed to measure continuous biogas production in the primary reactors. The design originated from a tipping bucket rain gauge. The design plans for the tipping bucket gas meter are

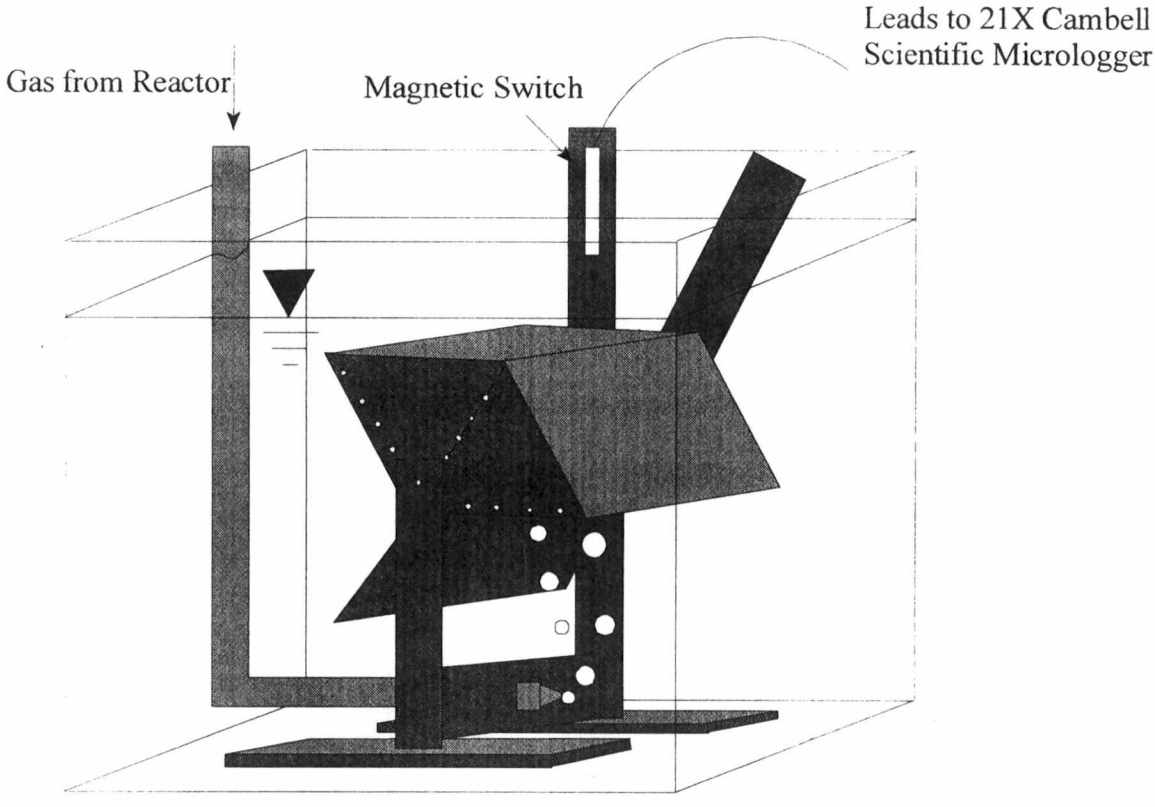


Figure 12. The tipping bucket gas meter used in the study.

included in Appendix C. In order to measure biogas, the tipping bucket was inverted and submerged under water. A magnetic switch located on an extension above the gas meter monitored the number of tips induced by biogas production. The 21X counted the tips. The gas meters were calibrated to measure 40 mL of biogas per tip. Calibration was performed by injecting a measured quantity of air into one side of the submerged bucket. The center of gravity of the bucket was adjusted until an adequate volume was obtained. The center of gravity of the tipping bucket must be above the pivot point so that each side of the inverted container can be filled.

After January 1998

Assessment of volatile fatty acids was performed via Standard Method 5560 – C, the distillation method (Greenberg et al., 1992). The distillation apparatus used in the measurement was calibrated with acetic acid at a concentration of 500 mg L^{-1}. The VFA concentration calculation was a means to account for the distillation apparatus' efficiency.

Alkalinity measurements were made on samples in conjunction with VFA evaluations. Monitoring of the total and bicarbonate alkalinity occurred. Standard Method 2320 – B: procedure b, the titration method, was used to determine total alkalinity which involved titrating a sample to an endpoint pH of 4.5 (Greenberg et al., 1992). Bicarbonate alkalinity determination involved titrating a sample to an endpoint pH of 5.75 (Ripley et al., 1986). The pH apparatus previously described measured the pH endpoints.

The amount of methane in the biogas was calculated via subtraction of the CO_2 volume from the production rate. Because no sulfur was available from the wastewater, it was assumed that minute amounts of hydrogen sulfide were present in the off-gas. Also, an assumption was made that 95% of the gas existed as either CO_2 or CH_4 (Stover, 1998). A Fyrite Gas Analyzer evaluated the amount of CO_2 in the biogas (Bacharach, 1994).

UASB Biomass Concentration

At the end of the study, samples were drawn from the UASB reactor to determine volatile suspended solids (VSS). Standard Method 2540-E determined the reactor's VSS (Greenberg et al., 1992). Because of the sludge bed in the lower portion of the reactor, samples were collected from the top and bottom reactor ports. The results from the two ports were weighted to account for differing proportions of biomass in the upper and lower portion of the reactor to determine overall biomass concentrations.

Combining biomass and substrate effluent concentrations allowed determination of the specific substrate utilization rate. Estimation of mass transfer effects also required determination of the reactor's biomass concentration. The specific substrate utilization rate was calculated with equation 3. The modified Damkohler equation described by Pavlostathis and Gomez (equation 7) was used to calculate mass transfer. The MatLab program used to determine the modified Da and the U value is provided in Appendix D.

CHAPTER IV

RESULTS AND DISCUSSION

The secondary DFAF in System 1 was the first anaerobic reactor to be brought on-line, despite several previous months of attempting to start-up the primary UASB. After the aquisition of granulated bacteria in September 1997, the UASB came on-line quickly. Except for the failure occurring in December 1997, System 1 has been in successful operation for ten months. The main focus of the study has been to increase the loading on System 1 to levels required by the confectionery plant. System 2, brought on-line late in the study, was more difficult to start-up, presumably because of the decreased (25°C) operating temperature in the primary reactor. Attempts were made to increase the loading on System 2, but the biochemistry in the primary DFAF never stabilized. The following sections describe the performance of all the reactors operated in the study; with the focus on System 1.

Performance of System 1

UASB-1A

By the end of the first week of UASB operation during September 1997, the reactor was achieving 60% COD removal at an OLR of 0.2 kg COD m^{-3} d^{-1}. Over a period of two months, OLR was increased from 0.5 to 3.0 kg COD m^{-3} d^{-1} in 30% increments. The ORR increased with the OLR and the treatment efficiency remained above 50%. The OLR, ORR, and TE during the September start-up are shown in Figure 13. During the

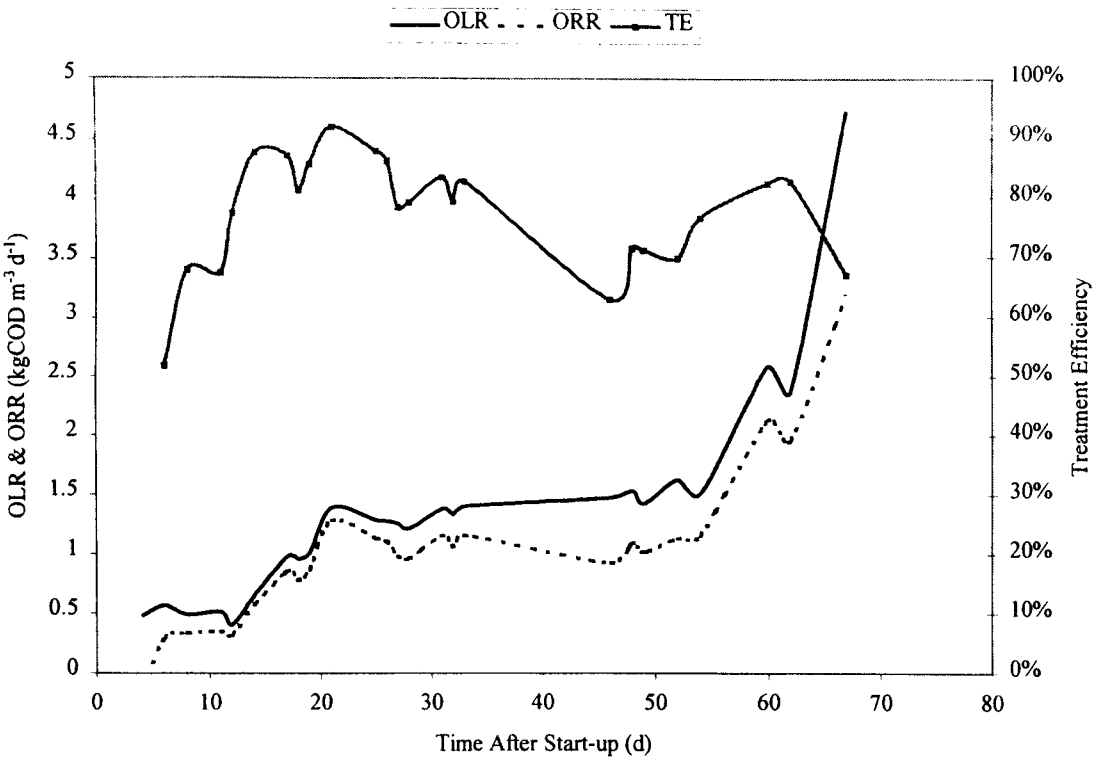


Figure 13. Performance of UASB-1A during the September start-up as OLR increased.

ninth week, the OLR was inadvertently increased by 55% (to 4.7 kg COD m^{-3} d^{-1}), which was greater than 30% as had been the rule.

The first indication of reactor failure occurred when the pH level dropped from 6.7 to 5.2 overnight. During failure, biogas production was reduced to 1 L CH_4 d^{-1}.At this time, the first VFA measurement on the reactor was made, and the VFA concentration was 1360 mg HAc L^{-1}. An attempt was made to prevent reactor deterioration by stopping the influent and adding sodium bicarbonate to raise the pH. The pH was raised to 6.9, but increased VFA concentrations persisted and no COD reduction resulted.

The overload on the system disrupted the stable microbial populations in the reactor. Hydrolyzing and fermenting bacteria have higher growth rates than the acetotrophic bacteria. The result was excess VFA concentrations that were not adequately buffered. The H^+ concentration increased leading to reactor failure via decreased methanogen productivity (Speece, 1996).

The January 1998 start-up was aided by the ability to measure VFA and alkalinity concentrations. Figure 14 represents the UASB reactor's second start-up. Again, by the end of the first week, the COD treatment efficiency was above 50%. Over a period of 2.5 months the OLR in the UASB was increased from 1 – 7 kg COD m^{-3} d^{-1}. As before, the ORR increased with the OLR. By the end of 2.5 months, the reactor steadily achieved 90% COD removal. The reactor's sludge bed was recieving a substrate concentration of 10 g COD L^{-1} when the OLR was 7 kg COD m^{-3} d^{-1}. At this point, a recycle was initiated to dilute the COD concentration applied to the reactor sludge bed.

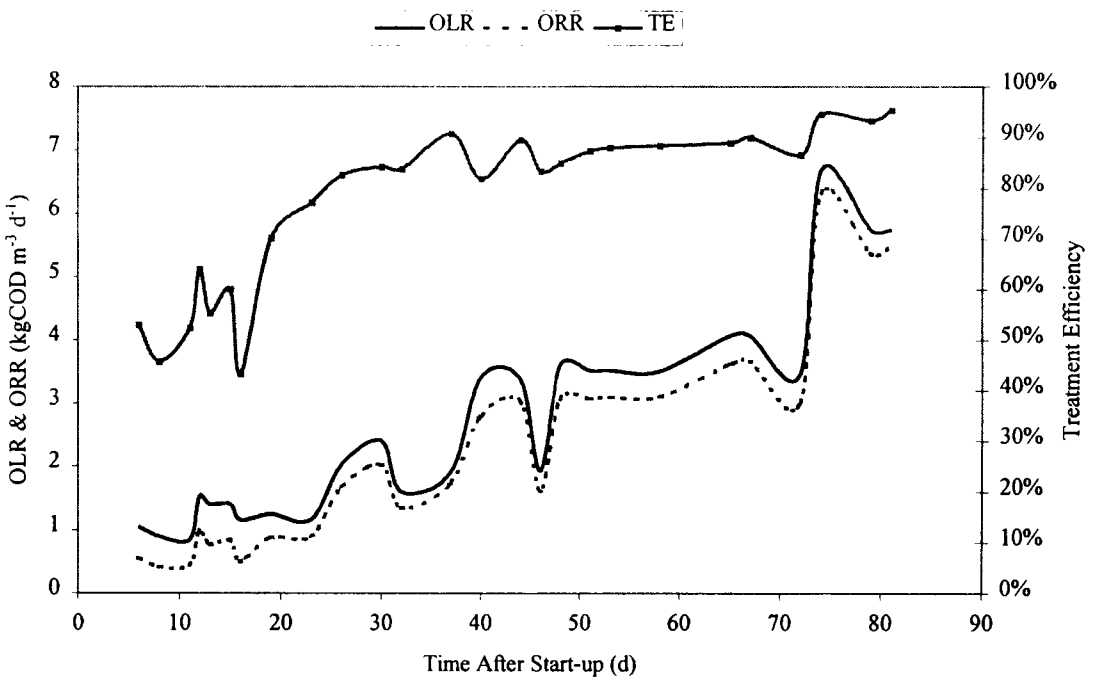


Figure 14. Performance of UASB-1A during the January start-up as OLR increased.

With recycle, OLR could be increased while the concentration entering the sludge bed remained below 10 g COD L^{-1}. At the beginning of recycle, the recycle line and the feed line pumped fluid at the same rate; this effectively doubled the rate at which waste entered the reactor. Figure 15 and 16 show the reactor's performance during recycle operation. At the start of recycle, the OLR was initially increased to 10 kg COD m^{-3} d^{-1} with a substrate concentration of 15 g COD L^{-1}; because of recycle, substrate concentration seen by the sludge bed was 7.8 g COD L^{-1}. The TE increased to 95%. During the next two weeks, OLR was increased two more times until it reached the design goal of 19 kg COD m^{-3} d^{-1} with a substrate concentration of 30 g COD L^{-1}. During this time recycle flow was doubled to increase recycle dilution. With the increased recycle, the sludge bed concentration was maintained below 10 g COD L^{-1}. During this period, COD TE remained between 90 – 95%. Lettinga (1978) found similar results on a UASB treating sugar beet waste. The reactor achieved 95% TE with an OLR between 15 – 30 kg COD m^{-3} d^{-1} at 25 – 35°C.

After recycle began, the ORR greatly increased, leading to a sharp increase in the biomass concentration. Prior to recycle, the sludge bed occupied the lower one-third of the reactor. By the end of the study, the sludge bed occupied the lower two-thirds of the reactor. Biomass quantities were not quantified until the final days of the study; changes in sludge bed size were based on visual observations. It was suspected that the large quantity of biomass helped maintain the high treatment efficiencies in the UASB during increased loading.

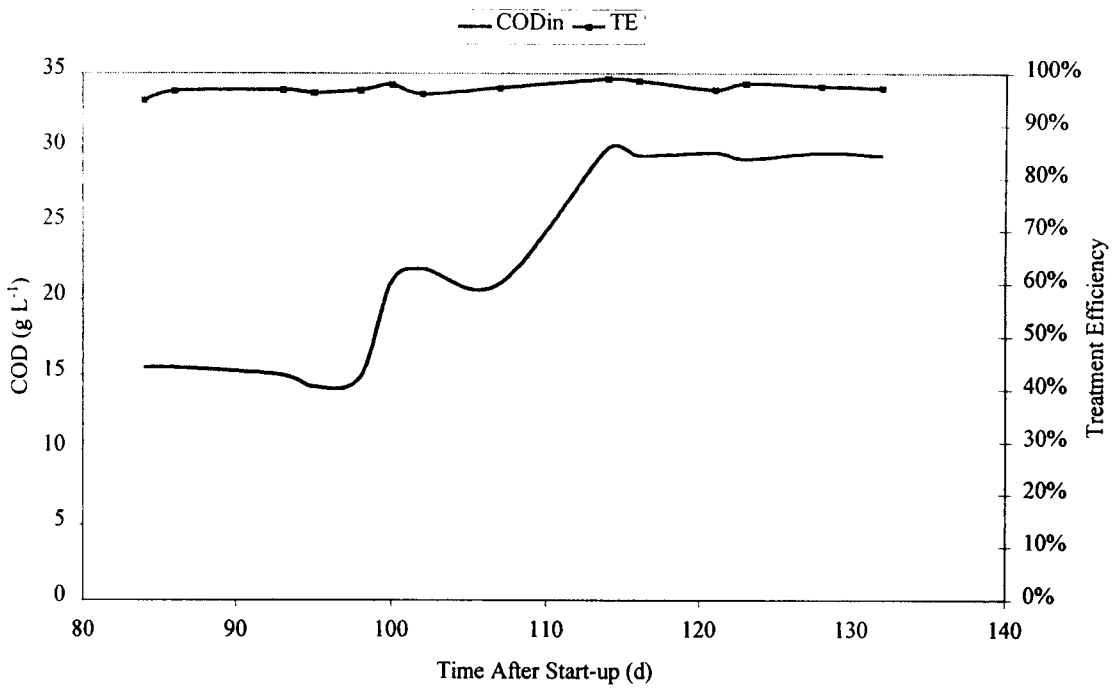


Figure 15. Performance of UASB-1A with increasing influent concentrations after the start of recycle.

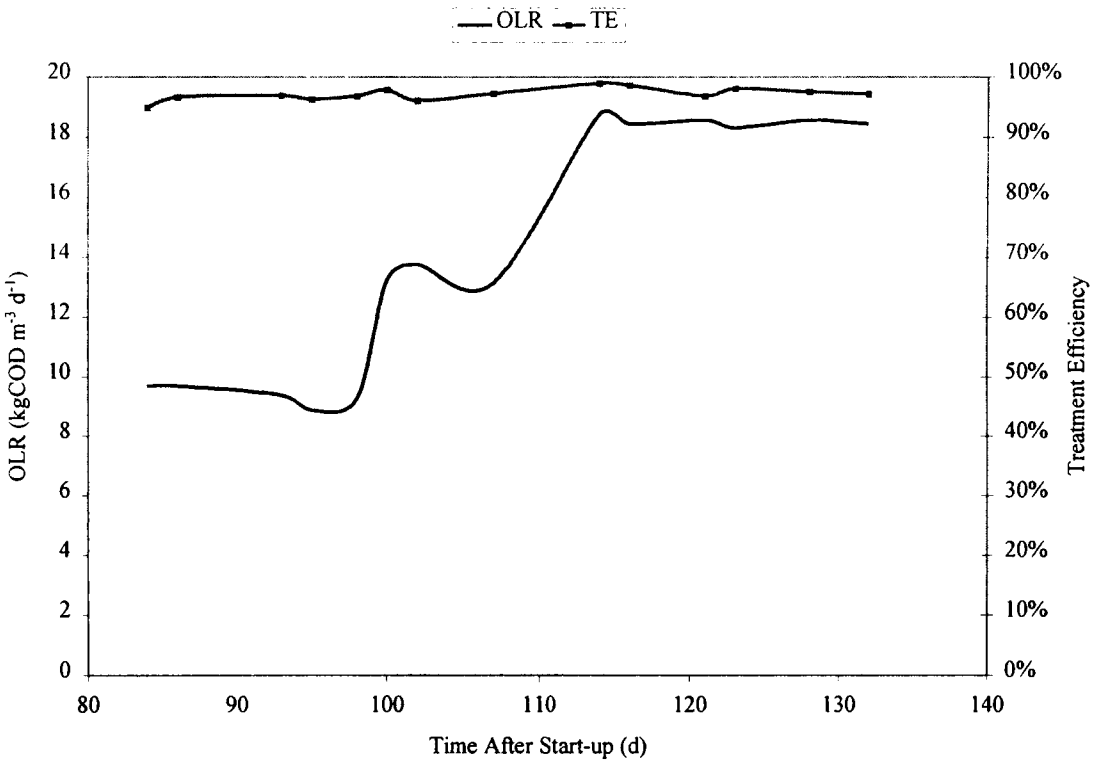


Figure 16. Performance of UASB-1A with increasing OLR during recycle.

Increased biomass quantities in the UASB decrease the reactor's working volume. Biomass in the reactor could be wasted to reduce these quantities. By wasting sludge, the VSS concentrations in the reactor will be reduced and if COD removal remained constant then the specific substrate utilization rate would be increased.

At the end of study, the UASB's biomass concentration was quantified. The sludge bed concentration was 44 g VSS L^{-1}, while the concentration of biomass above the sludge bed was 14 g VSS L^{-1}. Since the sludge bed made up two-thirds of the reactor volume, the average biomass concentration at the end of the study was 29 g VSS L^{-1}. These values are in agreement with literature reports. Speece (1996) stated that a successfully operating UASB retains biomass concentrations in the sludge bed around 30 – 80 g VSS L^{-1}. Stronach et al. (1986) claim the sludge bed should have a biomass concentration between 40 – 70 g VSS L^{-1}. They also stated that the average biomass concentration should be about 20 – 40 g VSS L^{-1}.

Increased velocities in the sludge bed may also have helped in the maintenance of high ORRs. Prior to recycle, the full sludge bed may not have been utilized. Fluid flows, before recycle, were smaller than with recycle possibly causing the lower portion of the sludge bed to achieve a higher percentage of substrate degradation than the upper portion of the sludge bed.

Due to UASB failure during December 1998, it was determined that VFA and alkalinity measurements were required to maintain a stable system. When the reactor was restarted in January 1998, VFA concentrations rose above 1000 mg HAc L^{-1}. The addition of sufficient $NaHCO_3$ to maintain the bicarbonate alkalinity above 1000 mg $CaCO_3$ L^{-1},

maintained the pH above 6.5 until the growth of acetotrophic bacteria reduced the VFA concentrations below 500 mg HAc L^{-1}, where they remained throughout the rest of the study (see Figure 17).

Initially, only total alkalinity was measured; in March bicarbonate alkalinity began to be measured. Because bicarbonate alkalinity only measures the alkalinity available to buffer the short-chain fatty acid and H^+ concentrations, it is more relevant than total alkalinity. Figure 18 shows VFA and alkalinity concentrations in UASB-1A. The reactor's bicarbonate alkalinity concentration never fell below 700 mg $CaCO_3$ L^{-1}. The pH was always maintained above 6.5, and reactor failure never occurred.

Though VFA concentrations never rose above 500 mg HAc L^{-1} after the first week of operation, they did oscillate as the OLR was increased to19 kg COD m^{-3} d^{-1}. In Figure 17, the peaks in the VFA concentration coincided with step increases in the OLR. After a few days of operation proceeding a step increase, the VFA content decreased on average to below 100 mg HAc L^{-1}. This trend was also seen by Cail and Barford (1985) treating cane juice stillage. It occurred because of the difference in growth rates of the hydrolyzing and fermenting bacteria described previously. Figure 19 shows the subsequent decreasing trend in the VFA concentration that occurred in the days following step load increases. In the days following step increases in organic loading, reactor VFA levels were elevated above the stablized concentration reached after each previous increase. The VFA concentration then decreased to its previously stable concentration.

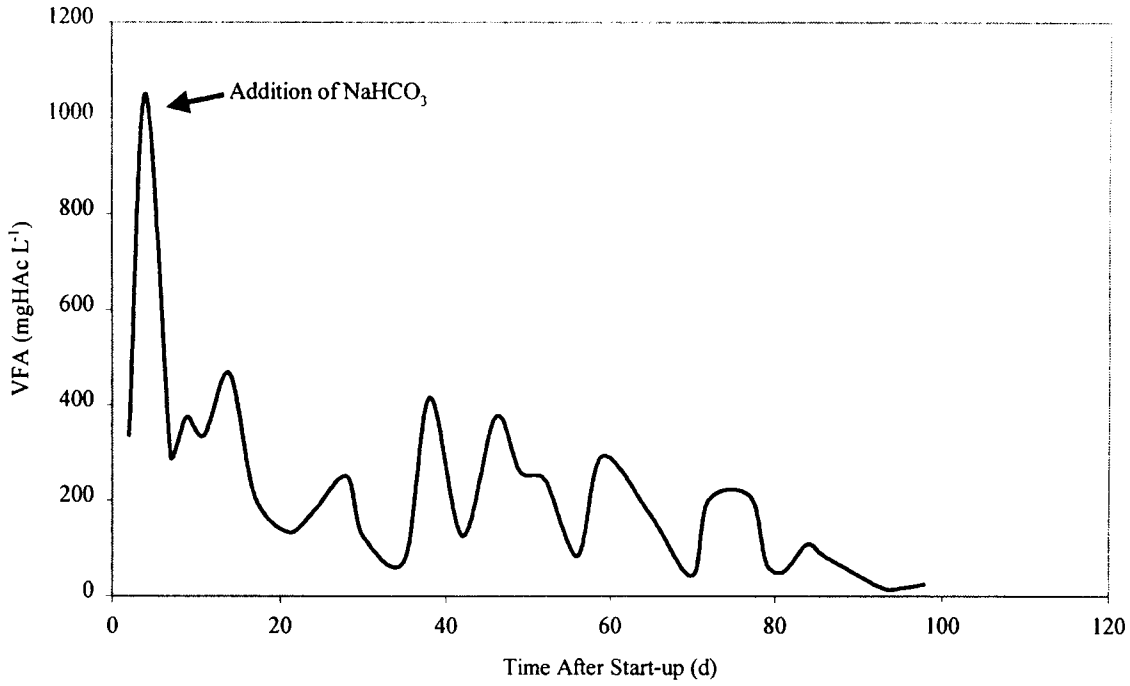


Figure 17. Variability in VFA concentration during the months following the UASB's January start-up.

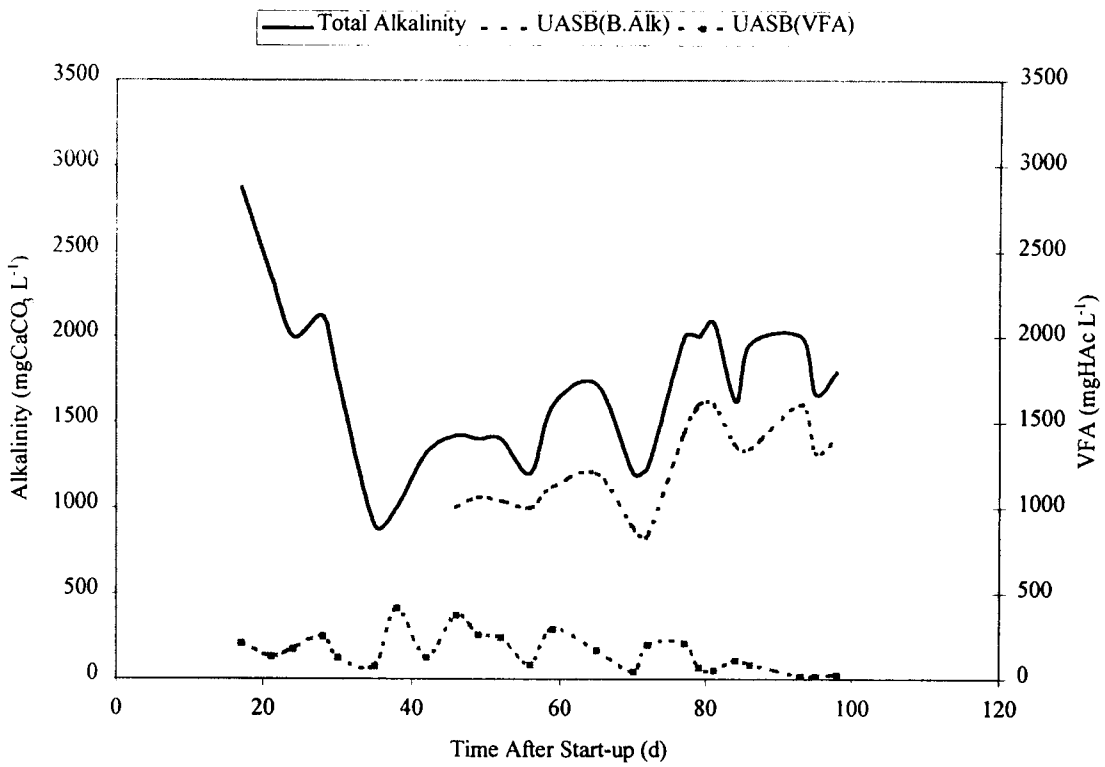


Figure 18. Relationship between total alkalinity, bicarbonate alkalinity, and VFA concentration in UASB-1A during the months following the January start-up.

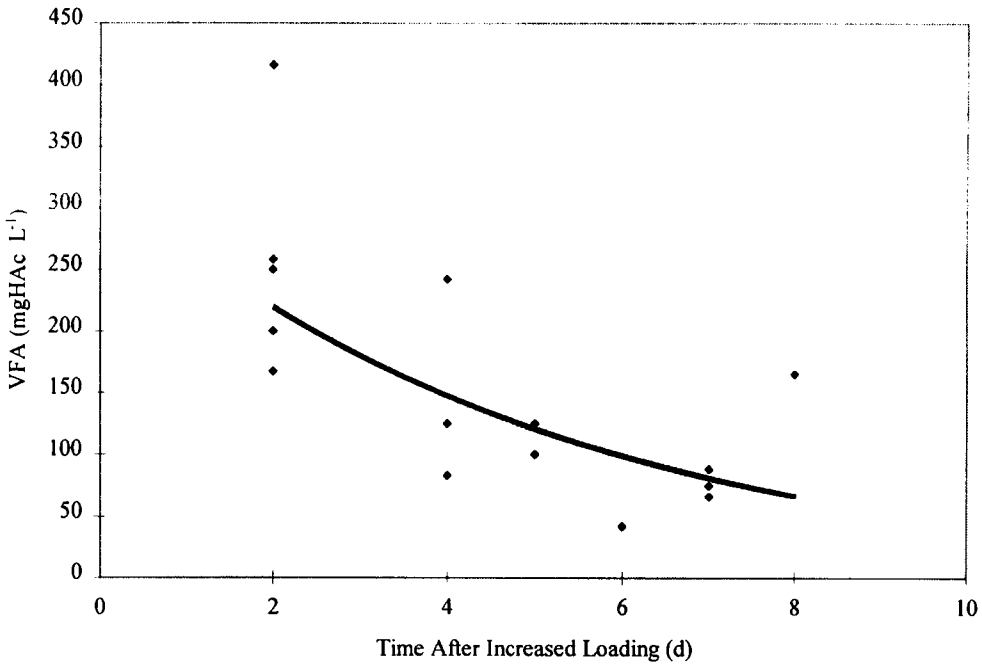


Figure 19. The decreasing trend in VFA concentrations in UASB-1A after elevated levels as a result of step increased OLRs.

Biogas production rate was measured throughout the study; however, biogas composition was only monitored after the January start-up. Figure 20 shows that methane production rates were closest to the stochiometric maximum methane yield rates when the mass COD removal was below 150 g d^{-1}. Mass COD removal below 150 g d^{-1} occurred while the OLR was below 6 kg COD m^{-3} d^{-1}. The average methane yield during this time was 0.28 L CH_4 (g COD removed)$^{-1}$. As the OLR and ORR increased, the methane yield appeared to decrease. However, this decrease reflected blockages that occurred in the gas line at the higher organic loading rates. The blockages were caused by increased concentrations of granulated bacteria in the upper portion of the reactor that clogged the gas lines. When the gas lines were plugged, the liquid level in the reactor dropped until it fell below the effluent port allowing gas to leave the reactor unmeasured.

Assuming the methane yield remained at 0.28 L CH_4 (g COD removed)$^{-1}$, the reactor would have produced around 100 L CH_4 d^{-1}, instead of the maximum measured production of 60 L CH_4 d^{-1}. This corresponds to a methane productivity of 5 L CH_4 (L-reactor d)$^{-1}$. The methane yield and productivity observed in this study concur with literature reports on UASB reactors treating carbohydrate rich wastewaters. A UASB treating diluted molasses at an OLR of 20.4 kg COD m^{-3} d^{-1} produced a methane yield of 0.23 L CH_4 (L-reactor d)$^{-1}$ (Jhung and Choi, 1995). The methane productivity of the UASB treating diluted molasses was 2 L CH_4 (g COD)$^{-1}$. Cail and Barford (1985) achieved a methane yield of 0.32 L CH_4 (g COD removed)$^{-1}$ at an OLR of 22.5 kg COD m^{-3} d^{-1} treating cane juice stillage. The methane productivity was 5 L CH_4 (L-reactor d)$^{-1}$.

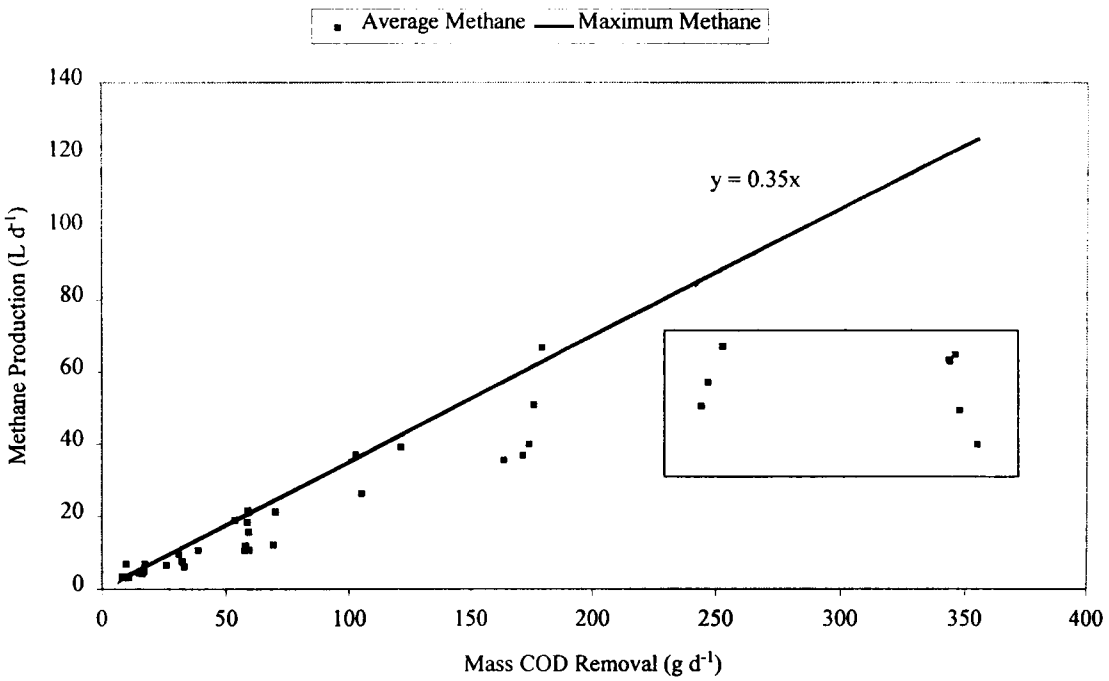


Figure 20. Methane production in UASB-1A as compared to the stochiometric maximum methane yield of 0.35 L CH_4 (g COD removed)$^{-1}$.

Figure 21 compares the relationship of the biogas production rate (which includes the CO_2 production rate) with the methane production rate, as the organic load increases. As the organic load increased, the percentage of CO_2 in the biogas increased from 22 - 44%. After recycle began, the composition of the biogas remained at 44% CO_2.

The methane produced during anaerobic digestion can be used to heat a reactor. The average temperature of the wastewater coming from the confectionery plant, as reported by plant personnel, was 30°C which was 5°C below the operating temperature of UASB-1A. The amount of energy required to raise the wastewater's temperature by 5°C was 0.25 MJ d^{-1}. Based on the energy content of methane and the adjusted methane production rate from the UASB, the reactor produced 3.1 MJ d^{-1} of biogas energy. This means excess energy was available from the process. A full scale UASB reactor at the confectionery plant achieving performance similar to the laboratory scale reactor would require 8,200 MJ d^{-1} to heat the reactor. The CH_4 yielded from the system would supply this demand and provide 90,000 MJ d^{-1} (25,000 kWh d^{-1}) in excess. The energy calculations are provided in Appendix E.

Overall, operation of the UASB during the final phase of the study was consistent. Stablization of the reactor's performance was due to VFA concentration and alkalinity adjustments that were made as a result of enhanced reactor monitoring. Further experiments should be performed on the UASB to determine the relationship between treatment performance and HRT.

Using the MatLab program in Appendix C, the modified Damkohler number was calculated (equation 7) to determine the mass transfer effects on the granules and

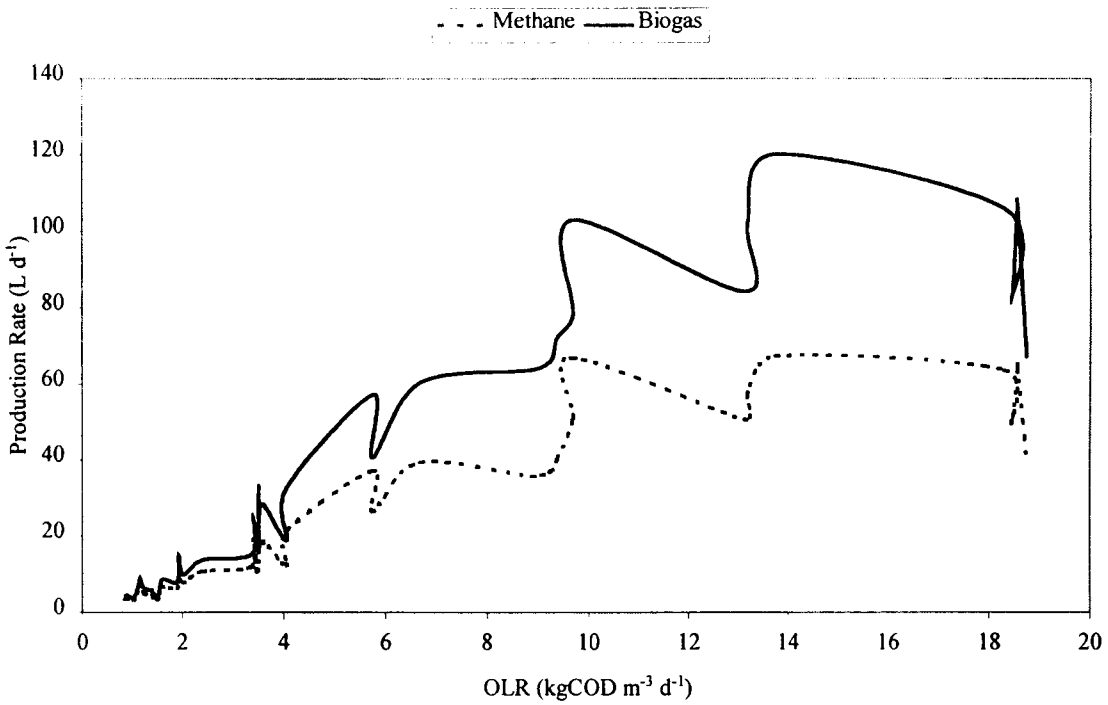


Figure 21. Comparison between biogas and methane production in UASB-1A as the OLR in the reactor increased.

substrates in the reactor. With two-thirds of the reactor volume filled with biomass, the Da number was 0.055 indicating a system with no diffusional limitations. When the sludge bed fraction was one-third the reactor's volume and with no recycle, the Da number was 0.092 which also implied no diffusional limitations. The primary role of recycle was therefore to decrease the sludge bed COD concentration; not to enhance mass transfer through stirring.

The specific substrate utilization rate was calculated using a MatLab script and the biomass concentration. The U value was determined to be 0.6 mg COD (mg VSS)$^{-1}$ d^{-1}. The calculated U value was slightly lower than the numbers determined in other studies on similar wastes. Cail and Barford (1985) reported the U value of a UASB treating cane juice stillage to be 0.77 mg COD (mg VSS)$^{-1}$ d^{-1}, and Driessen et al. (1995) reported a U value of 0.86 mg COD (mg VSS)$^{-1}$ d^{-1} for a UASB treating sugar cane molasses. The specific substrate utilization rate calculated for UASB-1A was slightly lower than the results presented for similar wastewaters. However, the sludge bed in UASB-1A filled two-thirds of the reactor volume, which was double the typical amount in a UASB. If the sludge bed volume was decreased via sludge wasting, the average biomass concentration in the reactor would decrease. Provided that reactor performance was maintained, the U value would be increased.

DFAF-1B

The three months of anaerobic filter performance prior to September 1997 was not well documented, as it was not yet part of the study. Though it existed one month without loading prior to becoming part of the study, it returned back on-line quickly due to the

anaerobic microorganisms already present in the system. Organic loading of the DFAF was fully dependent on the degree of treatment that occurred in UASB-1A. During the initial weeks of restart, the organic loading rate (1.25 kg COD m^{-3} d^{-1}) was greater than expected due to the low treatment occurring in the UASB-1A upstream of DFAF-1B, as seen in Figure 22. However, because the TE of UASB-1A increased, the OLR on DFAF-1B decreased to below 0.5 kg COD m^{-3} d^{-1} until the end of the first month..

By the end of the first month, as the organic loading rate increased to 1 kg COD m^{-3} d^{-1}, the TE increased to 80%. The rapid increase in the OLR after day 60 was caused by the failure of UASB-1A. Interestingly, though UASB-1A failed and consequently shock loaded DFAF-1B, DFAF-1B increased its ORR and TE with the higher organic loading. The response of DFAF-1B to the increased loading can be explained two ways. First, it is possible that the degradation process was operating in first order regime at the lower organic loading rate. Consequently, the microbes simply increased their degradation rates with increased loading. Alternatively, it is possible that only the upper portion of the filter was normally being utilized. In this scenario, the increased loading caused more of the filter bed to be utilized.

When the UASB was restarted in January, the DFAF initially received an organic loading rate around 1.2 kg COD m^{-3} d^{-1}. Figures 23 and 24 show the DFAF performance during start-up. The influent COD during this time was typically 1 g L^{-1}. Peaks in the OLR of DFAF-1B coincide with step increases in the OLR of UASB-1A. Each increased load on the DFAF was due to a decreased TE in the UASB. Effluent COD from DFAF-1B remained below 0.4 g L^{-1}.

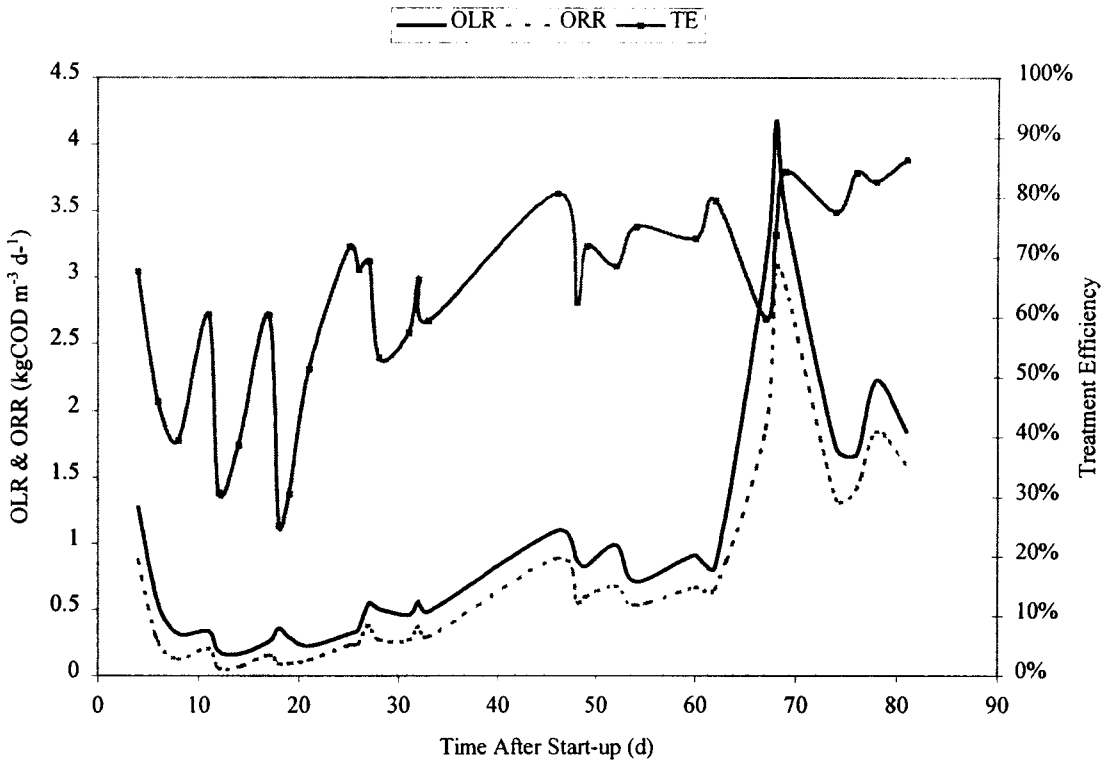


Figure 22. OLR and ORR in DFAF-1B after it was restarted in September 1997.

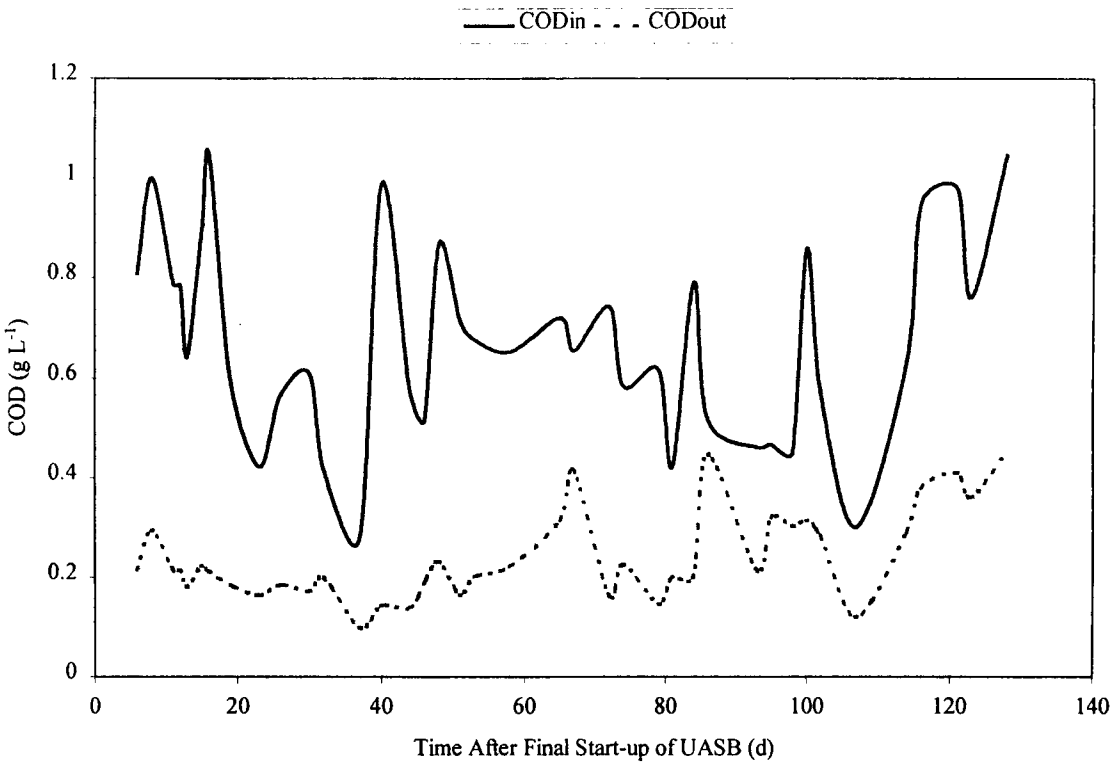


Figure 23. Performance of DFAF-1B with varying influent concentrations after the January start-up of UASB-1A.

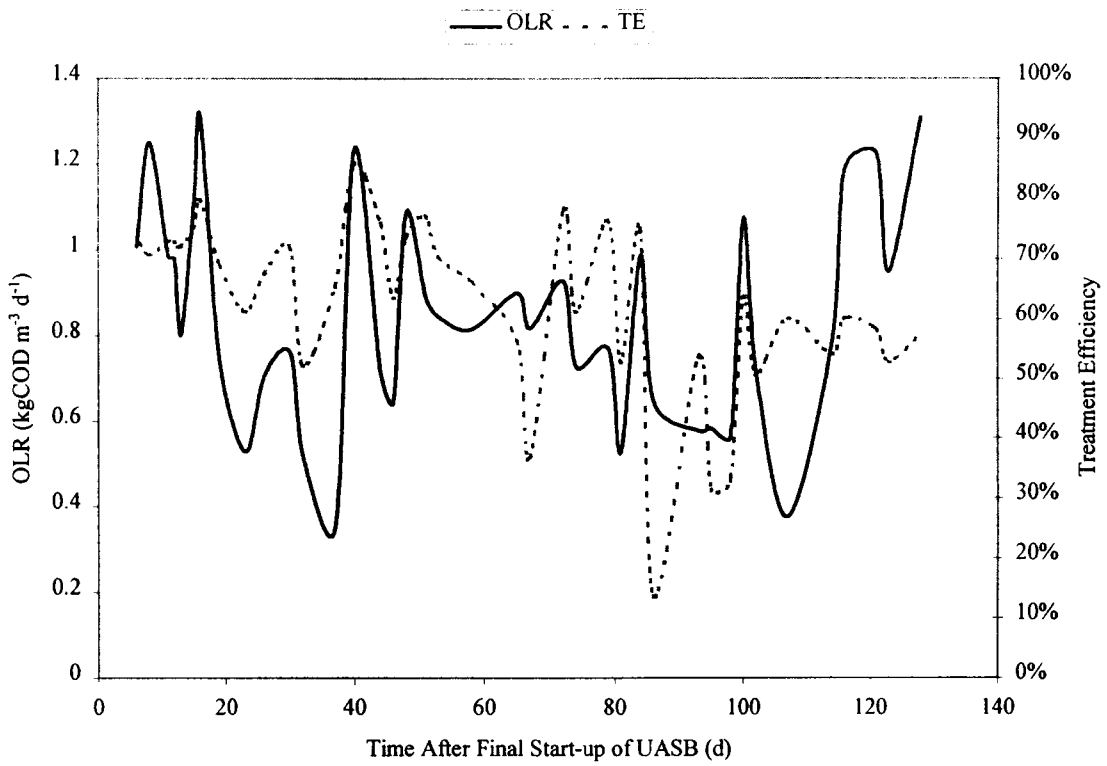


Figure 24. Performance of DFAF-1B with varying OLR after January start-up of the UASB-1A.

Biogas, VFA, and alkalinity data were not collected for this reactor. The pH of the reactor always remained greater than 7.

Composite Performance of System 1

When System 1 received its maximum OLR (at 12 kg COD m^{-3} d^{-1}), UASB-1A consistently achieved a COD treatment efficiency of 95%, while DFAF-1B consistently achieved a COD treatment efficiency above 50%. Therefore, overall TE of the composite system was 98% at a total system OLR of 12 kg COD m^{-3} d^{-1} and a total system HRT of 2.4 days. Figure 25 shows the substrate concentration as it moves through System 1 when it was loaded at 12 kg COD m^{-3} d^{-1}. On average, the sequential UASB - DFAF pair reduced from COD concentrations from 30 g COD L^{-1} to 0.4 g COD L^{-1}. While not yet dischargeable to surface waters, this 0.4 g COD L^{-1} waste was much more amenable to aerobic treatment than was the raw waste.

Performance of System 2

DFAF-2A

It is suspected that the wheat germ carrier for the anaerobic microorganisms used to inoculate the reactor underwent anaerobic degradation. This degradation resulted in greater effluent COD values than influent COD values. Figure 26 shows the performance of DFAF-2A during start-up. When continuous loading began, the OLR was 2 kg COD m^{-3} d^{-1}. As the loading rate increased, so did the TE. At 45 days after start-up, the OLR was 3.1 kg COD m^{-3} d^{-1} with a COD TE of 60%. During this time the ORR was greater than 2 kg COD m^{-3} d^{-1}.

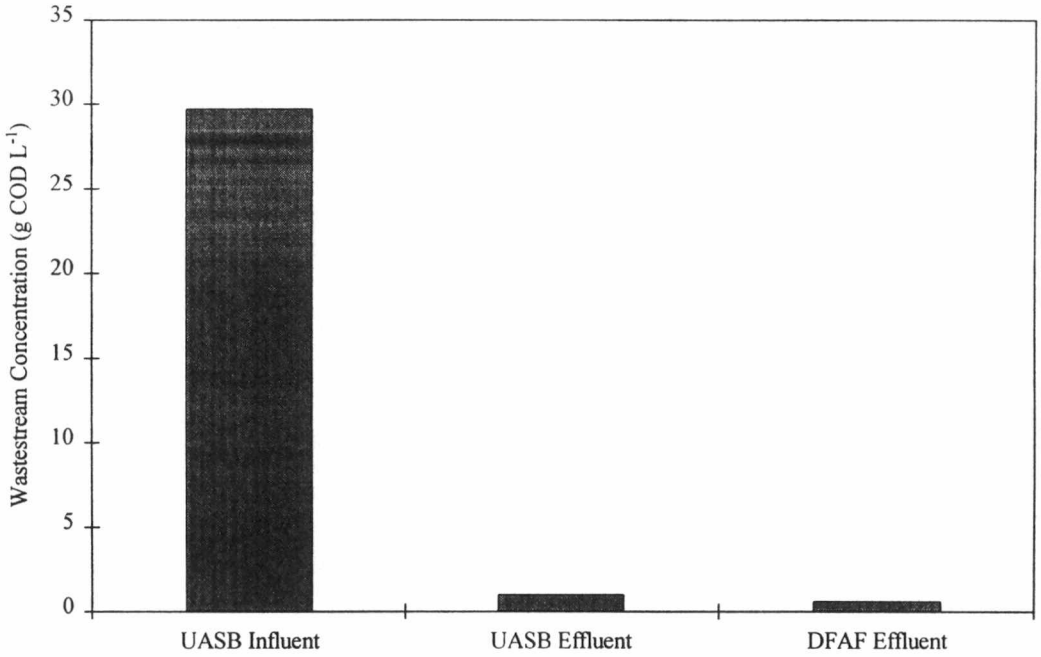


Figure 25. Influent COD reduction throughout System 1.

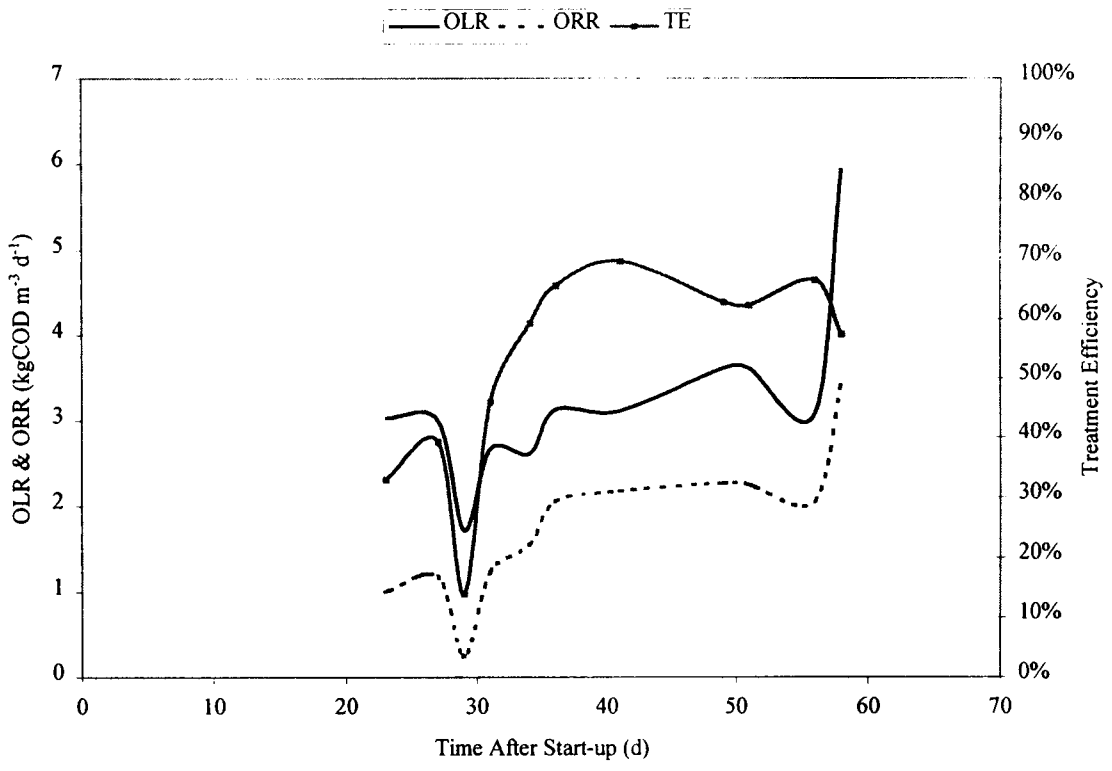


Figure 26. Performance of DFAF-2A after February start-up

Figure 27 shows the VFA and alkalinity concentrations in the reactor. Similar to UASB-1A, the VFA concentration was initially greater than 1000 mg HAc L^{-1}. By the third week, the VFA concentration dropped to 600 mg HAc L^{-1}, and continuous loading began. With the increased loading, the VFA levels increased. As the microbial populations adjusted, the VFA concentrations decreased. The organic load was increased on the 48th day, and the VFA concentration rose. However, because the reactor was not sufficiently buffered, the pH fell. One day later, the OLR was reduced to prevent failure, but it was inadvertently increased two days later due to an erroneous COD measurement. As seen in Figures 28 and 29, the treatment efficiency continued to decline until the loading was decreased. In looking a Figure 27, the failure should have been foreseen due to the low bicarbonate alkalinity to VFA ratio. The bicarbonate alkalinity concentration was in a range similar to those in the UASB; however, the VFA concentrations in DFAF-2A were much greater than those in the UASB.

After failure, an attempt was made to bring DFAF-2A back on-line. Late in the study, treatment efficiencies were increased at a low OLR (ca. 2 kg COD m^{-3} d^{-1}). However, the OLR was never increased due to the elevated VFA concentrations that persisted in the reactor. DFAF-2A was built due to the success of the secondary anaerobic filter in System 1. However, during operation, they did not perform equally. The composition of the wastewater seen by both reactors was similar. The probable cause for the poor performance of DFAF-2A was the increased OLR. On average, DFAF-2A received an OLR twice that of DFAF-1B.

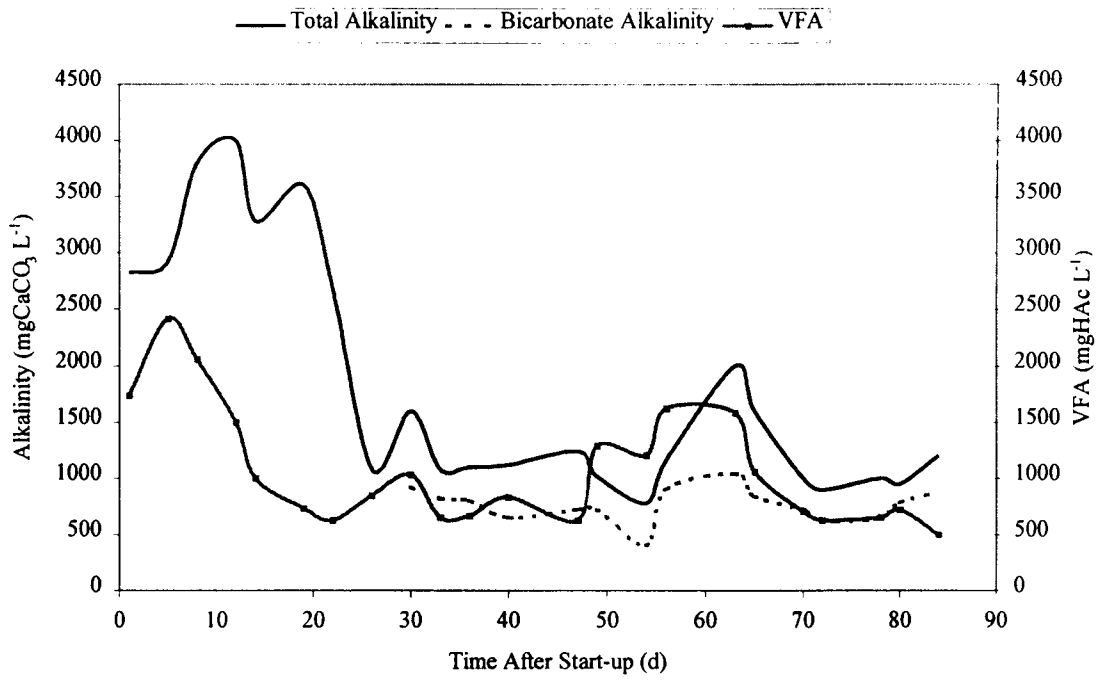


Figure 27. Relationship between total alkalinity, bicarbonate alkalinity, and VFA concentration during DFAF-2A operation.

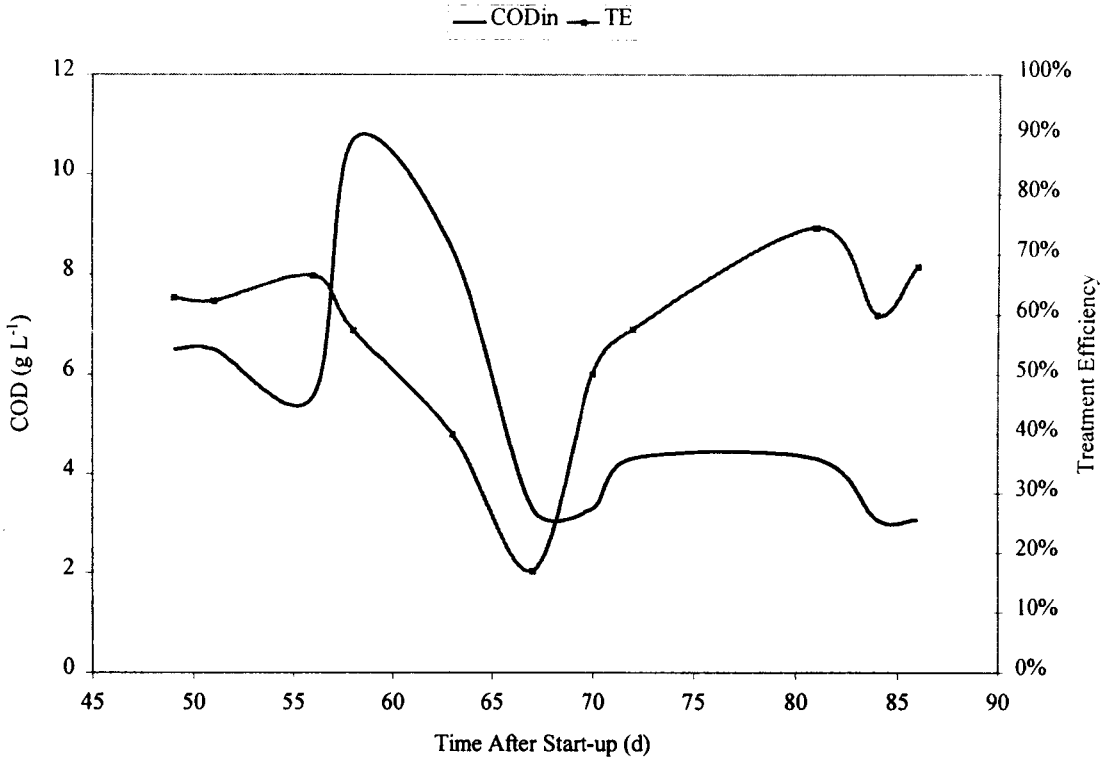


Figure 28. Performance of DFAF-2A with decreased substrate concentration after the failure occurred.

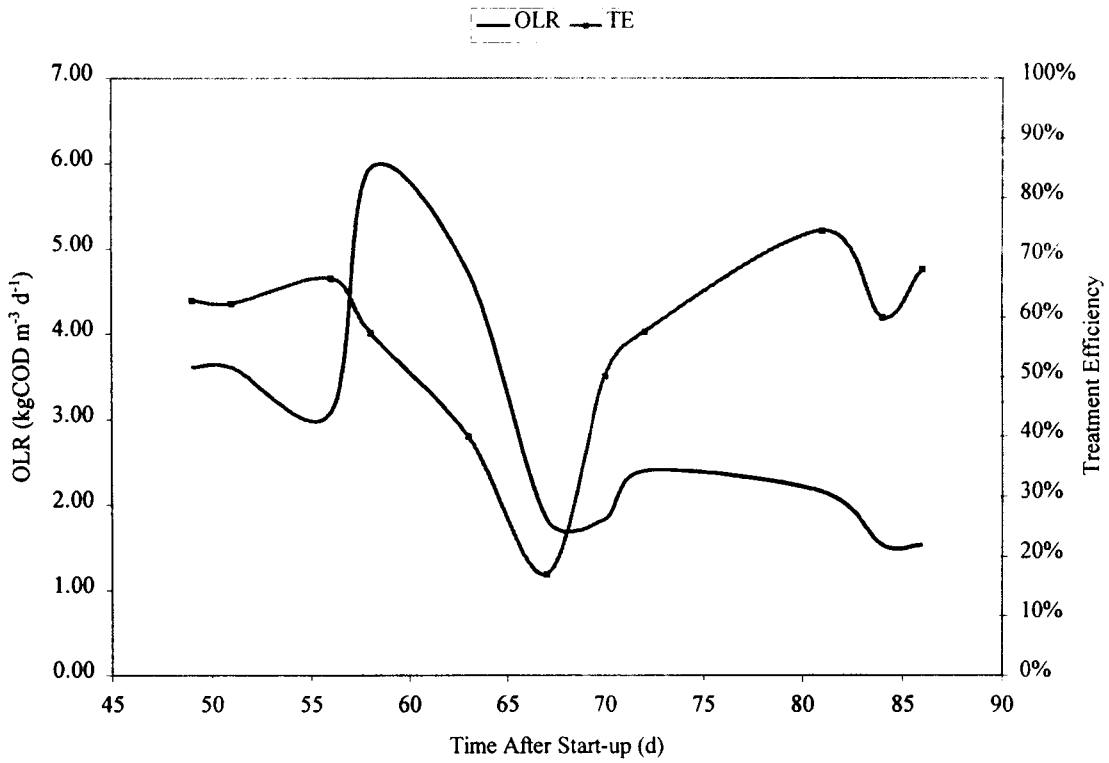


Figure 29. Performance of DFAF-2A with decreased OLR after failure occurred.

Biogas production in DFAF-2A was highly variable due to the variance in reactor performance. Prior to reactor failure, the organic loading rate was 3.1 kg COD m^{-3} d^{-1} and the methane yield was 0.28 L CH_4 (g COD removed)$^{-1}$.

A possible reason for the high volatile acid concentration and poor performance in DFAF-2A as compared to UASB-1A was the temperature difference between the two reactors. The lower temperature may have further slowed the growth of the acetotrophic bacteria.

DFAF-2B

The secondary anaerobic filter in System 2 was operated for the least amount of time. It was started similarly to the secondary anaerobic filter in System 1; its inoculum was bacteria in the upstream reactor's effluent. Also like DFAF-1B, its OLR was dependent on the influent and treatment efficiency of the upstream reactor (DFAF-2A). Figure 30 shows DFAF-2B start-up. The initial OLR was 3.5 kg COD m^{-3} d^{-1} as the upstream reactor began coming on-line. The treatment efficiency initially rose to 30%, but quickly fell to 10% where it remained for the next four weeks. The treatment efficiency increased again when DFAF-2A began to fail during the overload, peaking at 50% when the reactor was loaded at 5.5 kg COD m^{-3} d^{-1}. As the OLR was reduced to below 1 kg COD m^{-3} d^{-1} the COD TE dropped to below 10%.

The last drop in reactor efficiency may suggest that the filter should be loaded at increased organic loading rates to maintain effective treatment. However, this was not likely since the secondary anaerobic filter parallel to it performed adequately at similar loading rates. It was more likely that the increased loading caused an elevation in the

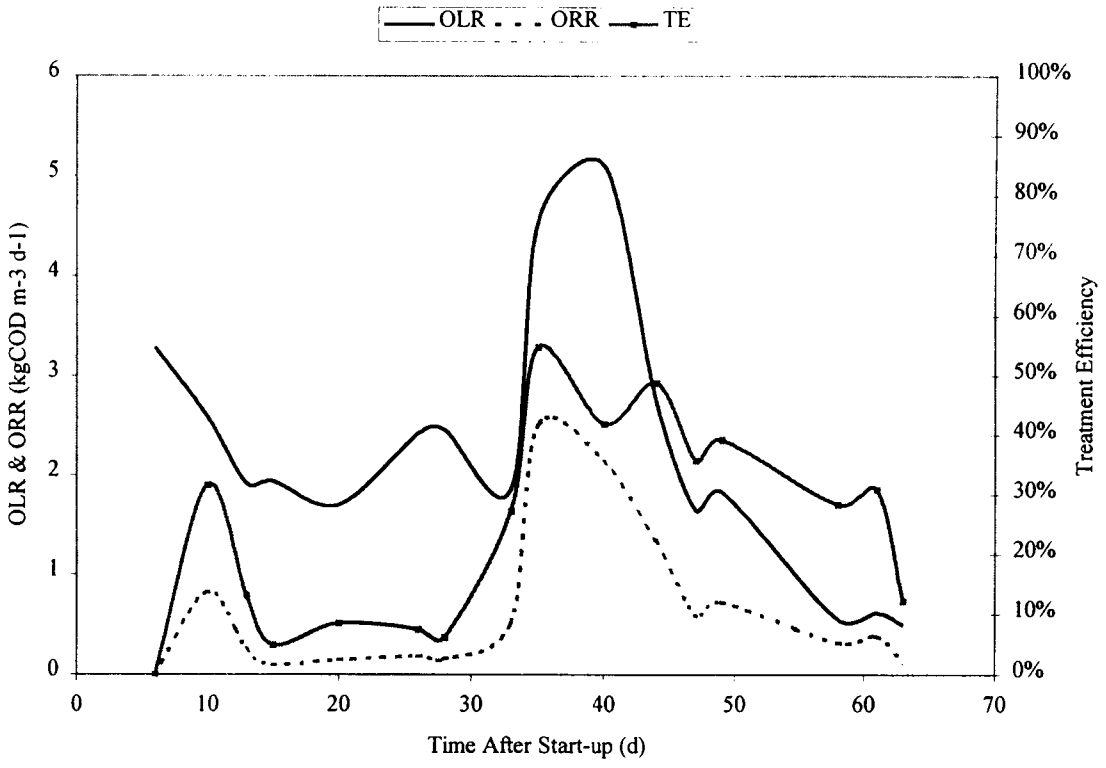


Figure 30. Performance of DFAF-2B during start-up.

VFA concentration that reduced the methanogen population, resulting in reactor failure. Volatile fatty acid measurements were not made in DFAF-2B.

Composite Performance of System 2

System 2 never achieved the performance maintained in System 1. The maximum ORR achieved by DFAF-2A for more than a week was 2.3 kg COD m^{-3} d^{-1} when the OLR was 3.5 kg COD m^{-3} d^{-1} (65% removal). The maximum ORR in DFAF-2B achieved for more than a week was 0.5 kg COD m^{-3} d^{-1} when the OLR was 1.2 kg COD m^{-3} d^{-1}(40% removal). However, the performance of both anaerobic filters in System 2 was never at steady state. The maximum overall COD removal efficiency in System 2 was 80%. This number was not representative due to the large variance in the system's results.

In retrospect, DFAF-2A should have been operated at 30°C, which would have enhanced its ORR potential, while not requiring supplemental reactor heat (at full scale). This might have enabled successful reactor operation, and would have then allowed a more meaningful comparison between System 1 and System 2.

CHAPTER V

CONCLUSIONS AND RECOMMENDATIONS

The study involved the operation of four continuously fed high-rate anaerobic reactors varying in volume from 10 to 24 L during a 12-month period. After overcoming hydraulic, microbial, and biochemical monitoring problems, an understanding of bench-reactor start-up and operation was achieved.

System 1 (UASB-1A and DFAF-1B) reduced the COD concentration in the confectionery plant's combined wastewater by 98%. Of the two systems studied, it was more feasible. Because the effluent COD concentration was 0.4 g COD L^{-1}, it could not be released to surface waters without further treatment. Table 4 provides the overall operation characteristics of System 1. UASB-1A performed the largest amount of COD reduction removing 95% of the original wastewater concentration. Further experiments should examine reactor performance under the following conditions: increasing the OLR above 19 kg COD m^{-3} d^{-1} by increasing the influent concentration, reducing the HRT, and wasting biomass to reduce the sludge bed to one-third of the reactor volume. DFAF-1B

Table 4. Overall System 1 achieved 98% COD TE with a UASB at 35°C a with HRT of 1.6 d and DFAF at 25°C with HRT of 0.8 d.

	System 1
Influent (g COD L^{-1})	30
Effluent (g COD L^{-1})	0.4
Nutrients Added	
Nitrogen (g NH_4NO_3 L^{-1})	2.5
Phosphorous (g DAP L^{-1})	0.5
Alkalinity (g $NaHCO_3$ L^{-1})	0.8
Methane Yield (L CH_4 (g COD removed)$^{-1}$)	0.28

performed consistently achieved above 50% COD TE; to increase the TE, lengthening the HRT and increasing of the operating temperature should be studied.

System 2 (DFAF-2A and DFAF-2B) did not treat the confectionery plant's wastewater as effectively and consistently as System 1. It did not appear to be a feasible treatment system. Though the composite system temporarily achieved 80% removal, operation of the two reactors was never stable. Also, the target OLR of 19 kg COD m^{-3} d^{-1} was never achieved; the highest OLR achieved was 6 kg COD m^{-3} d^{-1}. The performance of DFAF-2A could probably be enhanced by elevating the reactor operation temperature to increase microbial growth rates. The TE of DFAF-2B may increase with a more stable influent COD concentration during start-up, which could be achieved only if the upstream reactor is stable.

The UASB with the 35°C operating temperature performed better during start-up than the reactors operating at ambient temperatures. UASB-1A was able to start-up more quickly than the other reactors because of the increased microbial growth rates.

The secondary anaerobic filters (DFAF-1B and DFAF-2B) contained differing media types. DFAF-1B, containing brick packing, operated at steady state COD TE above 50%. DFAF-2B, containing the plastic packing achieved 40% removal at one time, but it never achieved steady state. The inability to bring DFAF-2B fully on-line prevents comparisons between the two media types.

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APPENDICES

APPENDIX A

21X REACTOR CONTROL PROGRAM

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Page 3   Table 2

08:  P          End Table 2

*     3         Table 3 Subroutines

01:  P          End Table 3

*     4         Mode 4 Output Options
  01: 0         (Tape OFF) (Printer OFF)
  02: 0         Printer 300 Baud

*     A         Mode 10 Memory Allocation
  01: 28        Input Locations
  02: 64        Intermediate Locations

*     C         Mode 12 Security (OSX-0)
  01: 00        Security Option
  02: 0000      Security Code
```

Page 2 Table 2

```
01:  P14        Thermocouple Temp (DIFF)
  01: 1         Rep
  02: 2         15 mV slow Range
  03: 3         IN Chan
  04: 1         Type T (Copper-Constantan)
  05: 2         Ref Temp Loc
  06: 3         Loc :
  07: 1         Mult
  08: 0         Offset

02:  P89        If X<=>F
  01: 3         X Loc
  02: 4         <
  03: 34.9      F
  04: 41        Set high Port 1

03:  P89        If X<=>F
  01: 3         X Loc
  02: 3         >=
  03: 36.1      F
  04: 51        Set low Port 1

04:  P14        Thermocouple Temp (DIFF)
  01: 1         Rep
  02: 2         15 mV slow Range
  03: 5         IN Chan
  04: 1         Type T (Copper-Constantan)
  05: 2         Ref Temp Loc
  06: 5         Loc :
  07: 1         Mult
  08: 0         Offset

05:  P89        If X<=>F
  01: 5         X Loc
  02: 4         <
  03: 3         F
  04: 52        Set low Port 2

06:  P89        If X<=>F
  01: 5         X Loc
  02: 3         >=
  03: 6         F
  04: 42        Set high Port 2

07:  P14        Thermocouple Temp (DIFF)
  01: 1         Rep
  02: 2         15 mV slow Range
  03: 4         IN Chan
  04: 1         Type T (Copper-Constantan)
  05: 2         Ref Temp Loc
  06: 4         Loc :
  07: 1         Mult
  08: 0         Offset
```

```
Program:
Flag Usage:
Input Channel Usage:
Excitation Channel Usage:
Continuous Analog Output Usage:
Control Port Usage:
Pulse Input Channel Usage:
Output Array Definitions:


*      1          Table 1 Programs
  01: 1200        Sec. Execution Interval

01:   P3          Pulse
  01: 1           Rep
  02: 1           Pulse Input Chan
  03: 2           Switch closure
  04: 0           Loc :
  05: 1           Mult
  06: 0           Offset

02:   P86         Do
  01: 10          Set high Flag 0 (output)

03:   P70         Sample
  01: 6           Reps
  02: 1           Loc

04:   P77         Real Time
  01: 110         Day,Hour-Minute

05:   P17         Panel Temperature
  01: 2           Loc :

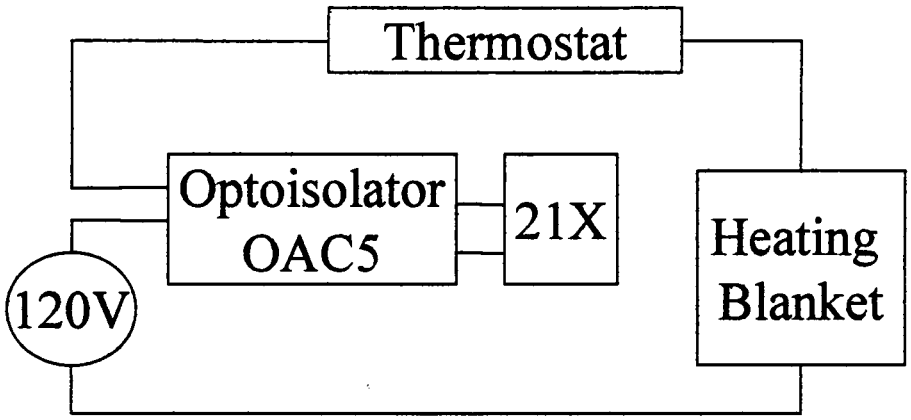
06:   P3          Pulse
  01: 1           Rep
  02: 3           Pulse Input Chan
  03: 2           Switch closure
  04: 6           Loc :
  05: 1           Mult
  06: 0           Offset

07:   P           End Table 1


*      2          Table 2 Programs
  01: 10          Sec. Execution Interval
```

APPENDIX B

CIRCUIT FOR CONTROLLING REACTOR HEATING PADS

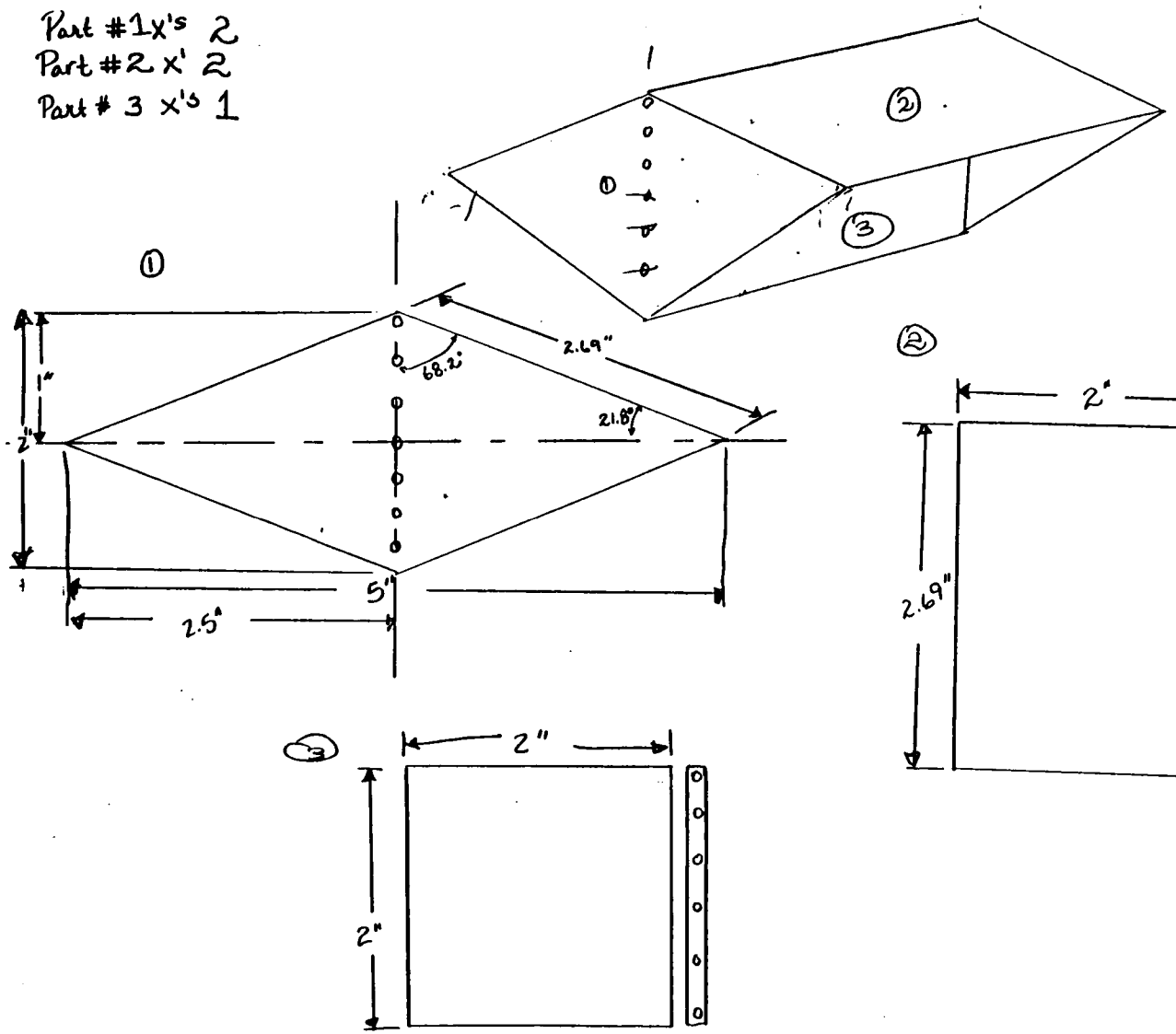


The optoisolator is controlled by the 21X. When the 21X senses the need for an increased reactor temperature, it switches the optoisolator on to allow the heating blankets to receive 120V. The opposite occurs when the 21X senses the need for a decreased reactor temperature.

APPENDIX C

DESIGN PLANS FOR TIPPING BUCKET GAS METER

Part #1 x's 2
Part #2 x' 2
Part # 3 x's 1



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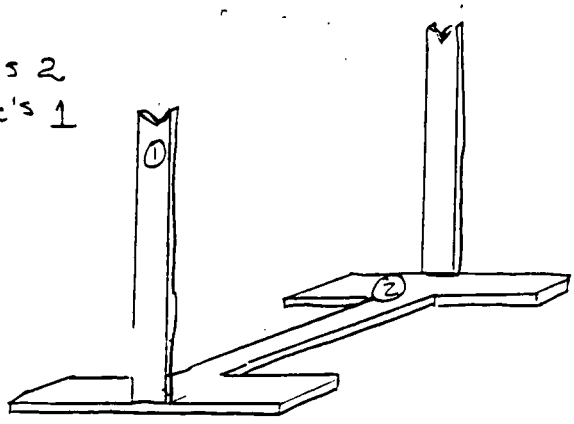
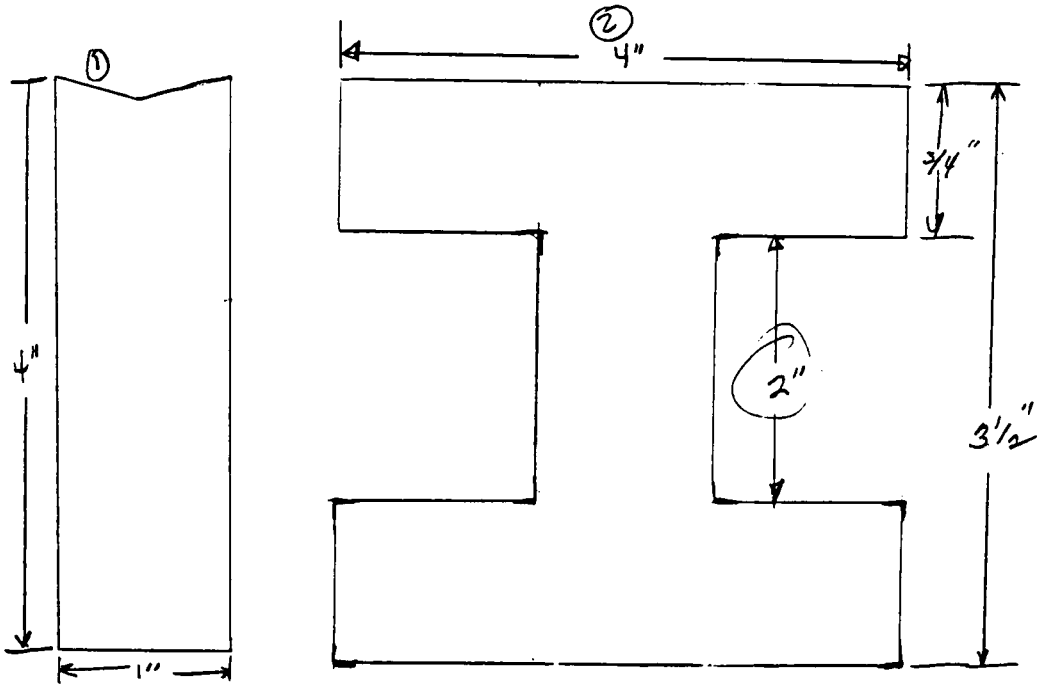
Part 1 x's 2
Part 2 x's 1





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2-3-97

APPENDIX D

MATLAB SCRIPT

```
%Mass transfer of substrate into the granules within the UASB reactor.

gsf=1                    %gas stirring factor to account for increased energy
                        %input to reactor due to biogas convection
cin=30000                %inlet concentration (mg/L)
cout=1000                %outlet concentration (mg/L)

deltaP=2;                %pressure drop (inches)
dPPa=deltaP*248;         %convert to Pa

HRT=1.6;                 %(day)

V=20e-3;                 %volume (m3)
sludge_fraction=0.3;     %fraction of reactor filled with granules

Q=V/(HRT*24*3600);       %liquid flow rate (m3 s-1)

Q=gsf*Q;                 %increase by gsf to account for gas-stirring
dPPa=dPPa*gsf;           %increase by gsf to account for gas-stirring

vl=Q*3600/(pi*0.1^2)     %liquid velocity (m/h) for comparison with T 18
m=V*1000;                %liquid mass (kg)
E=Q*dPPa/m               %energy dissipation rate per unit mass of liquid
viscosity_dynamic=0.7085e-3;     %dynamic viscosity (kg m-1 s-1)
kviscosity=viscosity_dynamic/1000; %kinematic viscosity (m2 s-1)
D=1.71e-10;              %for acetate (m2 s-1)
Sc=kviscosity/D          %Schmidt number
Dp=2e-3;                 %particle diameter (m)
Re=(E^(1/3)*Dp^(4/3)/kviscosity)^0.6  %modified Reynolds No.
Sh=2+0.51*Re*Sc^(1/3)    %Sherwood number

kl=Sh*D/Dp               %mass transfer coefficient (m s-1)

SAgranule=4*pi*(Dp/2)^2;     %Surface area of each granule (m2)
Vg=V*sludge_fraction;        %total volume of granules in the reactor (m3)
Vgranule=(4/3)*pi*(Dp/2)^3;  %volume of each granule (m3)
Ngranule=Vg/Vgranule;        %number of granules in the reactor
SAtotal=Ngranule*SAgranule;  %total surface area of granules in reactor (m2)
MRR=(cin-cout)*(V)/(1000*(HRT*24*3600))%Mass removal rate (kg s-1)
Vmax=MRR/SAtotal;            %maximum substrate uptake rate (kg m-2 s-1)

%calculate the overall specific substrate utilization rate in mg COD removed (mg VSS)-1 d-1
U=(cin-cout)/23000

Da_prime=Vmax/(kl*cout/1000)     %Damkoeler number for spherical biological particles
```

APPENDIX E

CALCULATIONS FOR ENERGY REQUIREMENTS AND YIELD

Calculations for Energy Requirements and Yield

Amount of energy required to raise wastewater temperature 5°C:

$$P = \Delta T \cdot C_p \cdot \dot{m}$$

where ΔT is the temperature change (K), C_p is the heat capacity for water at 30°C (J kg^{-1} K^{-1}), and $\dot{m}$ is the mass flow rate (kg d^{-1}).

$\Delta T = 5$ (K)
$C_p = 4.2$ (kJ kg^{-1} K^{-1})
$\dot{m} = 12$ (kg d^{-1})

$P = 0.252$ (MJ)

Amount of energy yielded from the reactor:

Energy in $CH_4 = 47 \times 10^3$ (kJ kg^{-1})
MW CH4 = 16 (g mol^{-1})

$$PV = nRT$$

where P is pressure (Pa), V is volume (m^3), n is the number of moles (mol), R is the gas constant (kJ mol^{-1} K^{-1}), and T is the temperature (K).

$P = 101.3$ (kPa)
$V = .100$ (m^3 d^{-1}) (amount of methane produced per day)
$R = 8.3$ (kJ mol^{-1} K^{-1})
$T = 298$ (K)

n = 4.1 (mol d^{-1})

Energy yielded = 3.1 (MJ d^{-1})

VITA

Lara Jacqueline Beal was born on May 1, 1974 in Atlanta, Georgia. After graduation from Powell Valley High School in Speedwell, Tennessee in 1992, she entered The University of Tennessee, Knoxville and received a B.S. degree in Agricultural Engineering in December, 1996. The following January she entered graduate school at The University of Tennessee, Knoxville. In August 1998, she received an M.S. degree in Biosystems Engineering.