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Combined Anaerobic/Aerobic Treatment for Municipal Wastewater

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COMBINED ANAEROBIC/AEROBIC TREATMENT FOR MUNICIPAL WASTEWATER

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ABSTRACT

Implementation of combined anaerobic/aerobic processes for wastewater treatment has been shown feasible, especially for industrial wastewaters with high concentration of organics. However, the utilization of this type of technology for treating wastewaters with low content of organic matter, such as domestic sewage is quite recent, and very limited information is available regarding the topic.

Recent investigations have demonstrated that it is feasible to utilize a combined technology composed of anaerobic pretreatment followed by an aerobic post-treatment to efficiently treat municipal wastewater. This research is a continuation of previous investigations about the feasibility of using an anaerobic fluidized bed reactor-aerated solids contact process to treat domestic wastewater. In the proposed system the excess sludge produced in the aerobic stage is recycled to the anaerobic unit. The proposed configuration is very attractive because the anaerobic fluidized bed reactor serves as pretreatment unit and a sludge digester at the same time.

The main objective of this research is to quantify the SS removal and accumulation rates in the AFBR, and determine the degree of stabilization of solids in the unit. All this to demonstrate the feasibility of avoiding separate sludge stabilization units.

An existing pilot plant unit, located in the Marrero Wastewater Treatment Plant, 6250 Lapalco Boulevard, Marrero, Louisiana 70072 served as experimental station for this project. The experimental program started on September 2002 and continued through December 2004.

Analysis of the data obtained shows that the anaerobic fluidized bed reactor is highly efficient in stabilizing the solids produced in the aerobic stage, and reducing the amount of sludge produced by the system. The results indicate that at a solids loading of 1.09 kg SS/m^3.d, 0.173 kg SS/m^3.d were consumed in the unit by the action of anaerobic bacteria, and 0.173 kg SS/m^3.d accumulated at the top of the fluidized bed of the unit. The results also show that the solids entrapped in the reactor are almost completely stabilized, and that due to their position at the top of the fluidized bed they can be removed without affecting the reactor operation.

1. INTRODUCTION

The anaerobic technology has been traditionally used for excess sludge digestion in wastewater treatment plants. In fact, it is one of the oldest processes used for the stabilization of solids and biosolids, and at present it remains as the dominant process for stabilizing sludges. In recent years anaerobic process has been used successfully for the pretreatment of industrial wastewaters and several studies have suggested its potential for the pretreatment of domestic wastewaters. Among the different anaerobic processes available, Anaerobic Fluidized Bed Reactors (AFBRs) emerge as a good alternative for the treatment of wastewaters with a low concentration of organics, such as domestic sewage. This is due to their unique characteristics, like the high concentration of active biomass they can maintain,

AFBRs have been successfully used for industrial wastewater treatment at real scale (Nicolella et al., 2000). However, little is known on the operation of full-scale system treating municipal wastewater. Some of the reasons for this imbalance are based on the advanced technology used on fluidized bed reactors and the fact that the process itself is difficult to control. This difficulty seems to increase with the size of the unit and numerous operational problems have been encountered in real scale AFBR. Another disadvantage associated with AFBR, like with other anaerobic systems, is that they usually produce a poor quality effluent. Consequently, post-treatment is usually required for removing residual BOD and TSS to acceptable levels.

Several post-treatment alternatives and process configurations have been suggested to improve the effluent of AFBRs. An important research effort has been carried out at the University of New Orleans UWMRC in this field. The research started in 2000, when Corzo (2001) studied the feasibility and efficiency of chemical and biological flocculation of the effluent of an anaerobic fluidized bed reactor treating municipal wastewater. Her study revealed that the AFBR/aeration chamber had an excellent potential for providing secondary treatment for municipal wastewater. Corzo also proved that it was feasible to recirculate the excess sludge from the aeration chamber to the AFBR, thus reducing the amount of surplus sludge produced. Corzo concluded that the AFBR/aeration chamber system is a very attractive alternative for municipal wastewater treatment because of its low operation and maintenance costs. Subsequently, Bustillos (2002) studied the same AFBR/AC system. In her research Bustillos investigated the effect of varying the hydraulic retention time (HRT) in the aerobic chamber to improve the quality of the final effluent. Bustillos concluded that the efficiencies of the anaerobic/aerobic process were higher at 100 min hydraulic detention time in the aerated solids contact chamber, producing an effluent with TSS concentration as low as 4 mg/L, a Total COD of 38 mg/L, and a Filtered COD of 13 mg/L. The present investigation is a continuation of the research started by Corzo (2001) in an effort to expand the understanding of the AFBR/AC system.

The investigation presented herein quantifies the SS removal and accumulation rates in the AFBR, and determines the degree of stabilization of solids in the unit, demonstrating the feasibility of avoiding separate sludge stabilization units.

1.1. Objectives and Scope

The main objective of this research is to determine the rates of accumulation and removal of suspended solids in the AFBR, in a combined anaerobic fluidized bed reactor / aeration chamber system in which the sludge generated during the process is recycled to the anaerobic unit.

The specific objectives of this project are the following:

- Set up the mass balance of solids in the anaerobic fluidized bed reactor.
- Establish the feasibility of pre-digestion of solids in the anaerobic fluidized bed reactor.
- Evaluate the AFBR effectiveness in removing total suspended solids, volatile suspended solids, and total chemical oxygen demand.

The experimental phase of this research was carried out at a large-scale pilot plant located at the Marrero Wastewater Treatment Plant in Marrero, LA. The influent of the pilot unit was municipal wastewater taken from Marrero's grit chamber splitter box which was treated by a rotating screen.

2. LITERATURE REVIEW

2.1. Combined Anaerobic Treatment of Wastewater and Sludge

Activated sludge system is by far the most commonly employed biological process used for treatment of industrial and municipal wastewaters (Dwight et al., 1997). This wastewater treatment system is highly versatile. However, it has the disadvantages of being energy intensive, and that it generates excess sludge whose treatment and disposal represents a major expenditure in wastewater treatment facilities. On the other hand, anaerobic digestion is among the oldest processes used for the stabilization of solids and biosolids. Anaerobic technology has been traditionally used for excess sludge digestion in wastewater treatment plants. Furthermore, because of the emphasis on energy conservation and recovery, anaerobic digestion continues to be the dominant process for stabilizing sludges (Metcalf and Eddy, 2003). In recent years anaerobic process has been used successfully for the pretreatment of industrial wastewaters and several studies have suggested its potential for the pretreatment domestic wastewaters. Furthermore, anaerobic pretreatment followed by aerobic post-treatment of wastewater is being used frequently (Jenicek et al., 1999).

Systems composed of anaerobic units followed by aerobic processes for effluent polishing utilize the benefits of both technologies. Series of reactors of anaerobic-

aerobic processes have been shown feasible for treating municipal wastewaters in warm climates resulting in lower energy requirements and less sludge production (Metcalf and Eddy, 2003). The combination of both anaerobic processes (excess sludge stabilization and wastewater pretreatment) in a single anaerobic unit would represent an enormous advantage for these combined (anaerobic pretreatment-aerobic polishing) technologies. However, little literature could be found regarding this topic.

Jenicek et al., (1999) stated that the main problems are the diverse character of wastes (wastewater and excess sludge), and the difference in the aims of both processes (wastewater pretreatment and sludge stabilization).

Adrianus et al., (1994) discuss the potential advantages of a combined anaerobic-aerobic treatment concept composed of a UASB reactor with complementary secondary treatment in an activated sludge process and stabilization of the excess active sludge in the UASB reactor.

According to Adrianus et al., (1994) the advantages of the anaerobic-aerobic treatment depicted in Figure 2-1 are:

- As a result of the removal organic material and suspended solids achieved in the UASB reactor, the sludge mass in the subsequent activated sludge process becomes relatively small and consequently the volume required for the activated sludge unit is reduced.
- The presence of the anaerobic reactor dispenses the need for a sludge stabilization unit. The excess activated sludge can be conveyed to the UASB reactor.

- The stabilized sludge production will be smaller in an anaerobic-aerobic system because of the comparably smaller sludge production in the anaerobic system. Additionally, the sludge stabilized in the anaerobic reactor has a high concentration. Therefore, the liquid-solid separation is simpler.
- By removing part of the organic load anaerobically, the oxygen demand of the aerobic stage is reduced. Consequently, less power is required for aeration. Moreover, depending on the efficiencies of methane production and collection part of the required power may be generated from the biogas produced.

Adrianus et al., (1994) also present a design example comparing the aerobic treatment of raw sewage, using a conventional activated sludge process, a sludge thickener and an anaerobic sludge digester versus the anaerobic-aerobic treatment system using a UASB reactor with secondary treatment in an activated sludge process, and stabilization of the excess active sludge in the UASB reactor. Figure 2-1 shows the configuration considered by Adrianus to do the comparison. The results of their calculations are very attractive. They concluded that the total volume required for the UASB-Activated sludge system would be 56 percent of the value needed for the conventional treatment option, and that the oxygenation requirements would be reduced by 56 percent or more. These results are promising. However, the authors' conclusions are based on a purely theoretical analysis and they did not report any experimental results to show the feasibility and efficiency of the proposed system.

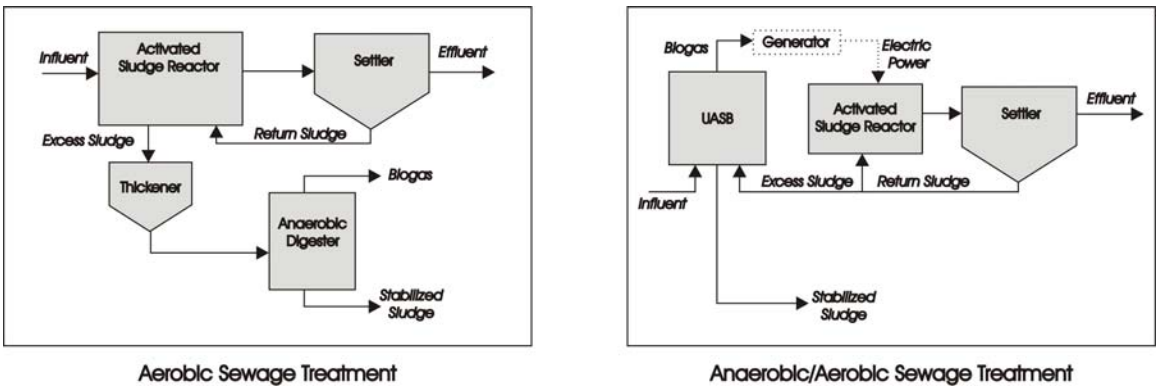

Figure 2-1 Configurations considered by Adrianus et al., to evaluate the advantages of the Anaerobic/Aerobic treatment over the typical Aerobic Treatment (Adapted from: Adrianus, et al., 1994)

In 1999 Jenicek et al., carried out a series of experiments in a laboratory scale upflow staged sludge bed (USSB) reactor with five compartments. The researchers operated the reactor within the mesophilic range (35°C) and fed it with artificially prepared glucose-based wastewater. At later stages of operation they added an aerobic biofilm reactor to study the denitrification phenomena. The reported volumes of the reactors were: USSB 4.0 L, aerobic biofilm reactor 4.0 L, settler 0.5 L. The height of the USSB reactor was 55.0 cm. The researches tested different alternatives trying to optimize the performance of the USSB reactor with respect to wastewater treatment and sludge stabilization. The researches concluded that in many cases the joint anaerobic treatment of wastewater with biological sludge could be an optimal technological solution from an ecological and economical point of view. They also concluded that with the anaerobic-aerobic treatment of wastewater, the presented technology is especially beneficial because of its simplicity and minimization of surplus sludge production. In

general, their experimental results show that the vertically compartmentalized USSB reactor is highly appropriate for the combination of sludge and wastewater treatment.

Corzo (2001) studied the feasibility and efficiency of biological flocculation of the effluent of an anaerobic fluidized bed reactor treating municipal wastewater. She studied a system composed of an anaerobic fluidized bed reactor (AFBR) followed by a small aeration chamber, with a hydraulic detention time of less than 1 hour. In this system the sludge collected at the bottom of the clarifier is recycled to both the aeration chamber and the anaerobic reactor. Her study revealed that the AFBR/aeration chamber had an excellent potential for providing secondary treatment for municipal wastewater. She also proved that it was feasible to recirculate the excess sludge from the aeration chamber to the AFBR as recommended by Adrianus (1994).

Continuing Corzo's research, Bustillos (2002) investigated the effect of varying the hydraulic retention time (HRT) in the aerobic solids contact chamber of the AFBR/aeration chamber system, in an effort to improve the quality of the final effluent. Bustillos concluded that the efficiencies of the anaerobic/aerobic process were higher at 100 min hydraulic detention time in the aerated solids contact chamber. She also concluded that the aforementioned system is highly efficient with 64% TCOD, 45% FCOD, and 92% TSS removal, and that the system reduces the amount of sludge produced.

The present study is a continuation of the extensive research project started by Corzo and Bustillos. The main objective of this investigation is to determine the rates of

accumulation and removal of suspended solids in the AFBR, and the degree of stabilization of the solids in the unit.

2.2. Anaerobic Fluidized Bed Reactors

An anaerobic fluidized bed reactor (AFBR) consists of a vertical vessel containing an inorganic media (e.g. rock, sand, activated carbon, anion and cation exchange resins…) (Metcalf and Eddy, 2003). The media serves as support for the development of a biofilm layer, which is retained by natural attachment of the microorganism to the solid substratum particles (Hidalgo and Garcia-Encina, 2001). The particles are fluidized by high upflow liquid velocities, generally produced by a combination of the influent and recirculation flow-rates (Iza et al., 1991). Depending on the type of media used, AFBRs can be operated at upflow liquid velocities as high as 20 m/h to provide about 100 percent bed expansion (Metcalf and Eddy, 2003). In the fluidized state the media provides a large specific surface for attached biological growth and allows biomass concentrations in the range 10-40 Kg/m^3 to develop (Nicolella et al., 2000). This large concentration of biomass in AFBRs results in smaller reactor volume (Hermanowicz et al., 1990). Moreover, process COD loading values of 10 to 20 Kg $COD/m^3.d$ are feasible for AFBRs with greater than 90 percent COD removal, depending on the type of wastewater (Metcalf and Eddy, 2003).

The advantages of the AFBR process include (Iza, 1991).:

- The ability to provide a high concentration of biomass, attached to a dense carrier, which cannot be easily washed out from the reactor.
- A very large surface area for biomass attachment.
- Initial dilution of the influent with effluent, which provides alkalinity and, thus, some neutralization (due to the CO_2 present in the effluent as a consequence of the anaerobic conversion), reduces substrate concentration (important for high COD wastes), and contributes to reduce the shock effect of toxicant spikes.
- High mass transfer properties.
- Low concentration gradients around the particles are possible, allowing the treatment of low strength wastes.
- No plugging, channeling or gas hold-up.
- Ability to control and optimize biofilm thickness.
- Biomass carrier can be tailored to a specific application to enhance performance

Some of the disadvantages of AFBRs that could be mentioned are lack of sufficiently comprehensive sets of reported experimental data for validation of proposed models (Fitzgerald, 1996); need of secondary treatment, due to inadequate effluent quality. Minimal solids capture due to the high turbulence and thin biofilms developed; pumping power required to fluidize the bed (Metcalf and Eddy, 2003); difficult and long

start-up for the development of enough mass inventories (Bustillos, 2002); bed height difficult to control; difficult reactor design and scale-up; little full-scale experience; and depending on the type of media used, the media cost may be a considered a disadvantage (Weiland et al., 1991).

2.2.1. Fluidization Phenomena

An important variable in fluidized bed reactors, which has a critical effect on their operation, is the fluidization and the percentage of bed expansion (Marin et al., 1999). The bed expansion establishes the organic matter residence time in the biocatalyst zone, and is directly related with the process pumping cost (Blanco et al., 1995)

Fluidization of low density solids such as glass, sand or GAC occurs when a liquid passes through a bed of particles producing the bed to expand and the particles to get suspended and free to move with respect to the others (Iza, 1991). The fluidization of a particulate bed occurs very smoothly, with a homogeneous expansion, if the particles are uniform, and with a high tendency to segregation, if particles are heterogeneous (Iza, 1991).

In order to get a fluidized bed, an increasing flow of liquid is applied through a settled bed of particles which form a fixed bed. During the progressive increase of flow, the bed starts to expand. At this moment, the equations that apply to fixed beds describe the head loss-velocity relationship. If the flowrate is increased, a transition occurs and particles start to move suspended on the upflow liquid separated from the other particles. At this very moment, fixed bed laws are still followed. If the flowrate is

further increased, particles separate more from each other and their hydrodynamic behavior resembles the behavior of particles settling. The limit of this phenomenon is called fluid transport, where particles are carried out of the bed by the liquid flow (Iza, 1991).

After passing the flow threshold which causes the fluidization, two different types of behavior can occur:

- Particulate fluidization, where the bed expands increasing the distances between particles.
- Aggregative fluidization, where the excess flow passes through the bed forming bubbles.

During fluidization, the bed itself behaves as a fluid with a new set of physical properties (density, viscosity), which follow hydrostatic and hydrodynamic fluid laws. These considerations and the improved characteristics of thermal and mass transfer are some of the reasons for the use of fluidized beds for biological processes (Iza, 1991).

2.2.1.1 Bed segregation:

The presence of particles with different shapes and sizes causes a segregated bed: the heavier particles move down to the bottom of the bed whereas the lighter ones rise to the top. In most cases there is a linear distribution by sizes from the bottom to the top. This phenomenon is caused by the interaction between segregation and diffusion (Iza, 1991). The size distribution of the particles used on a fluidized bed reactor is

usually very narrow. If broad ranges are used, the smallest particles are highly fluidized, even washed out from the reactor, whereas the bigger ones remain non-fluidized, forming a fixed bed (Iza, 1991). An important consideration concerning the biological nature of the process is that biofilm growth affects the size, overall density, shape and roughness of the particles, as well as its chemical and adsorptive characteristics. Therefore, along the operative life of a biological fluidized bed changes in the distribution of the bed particles are common.

The segregation phenomenon makes the AFBR attractive as a unit for the stabilization of the solids generated in the aeration chamber. Due to their lower sizes and densities it is expected that the entrapped organic solids remain at the top of the fluidized bed. This makes it easy to remove the excess solids without affecting the reactor biofilm coated media.

2.2.2. Media Selection

Types of particles used in fluidized bed include substances such as sand, coal, granular activated carbon (GAC), reticulated polyurethane foam, fired clay, porous glass beads, ion exchange media, and diatomaceous earth (Speece, 1966). The selection of a material for a fluidized bed reactor should consider many aspects of vital importance for the sizing of equipment, for the biological process itself and for the operation of the system (Iza, 1991). The carrier properties influence the reactor hydraulics and the biofilm thickness (Weiland et al., 1991). Consequently, both physical and chemical characteristics of the media should be considered prior its selection. The physical

characteristics that have to be considered are size, shape, particle density, hardness and surface area (Marin et al., 1999).

In AFBRs the biofilm formation is strongly influenced by the surface properties of the support media. Thus, porosity and roughness of the support surface play the major role during the first phase of start-up. Hence, for fast start-up, supports with a porous or rough surface are necessary or recommended (Weiland et al., 1991). As mentioned before, one of the advantages of fluidized bed reactors is the large surface area available for biofilm attachment, which allows higher concentration of biomass. Porous materials such as GAC, sepiolite, pumice, kaolinite, offer the advantage of internal pores which, depending on their size, can also be colonized, thus, increasing the amount of available surface area. This increment can be of two or three orders of magnitude (Iza, 1991).

The size of the particle influences the available surface for attachment as well as many characteristics of fluidization and consequently mass transfer (Iza, 1991). As Iza (1991) stated, “in order to reduce the diameter over the specific area and the operating costs the superficial fluidization characteristics velocity should be kept at low values, forcing the use of small size particles, which also provides greater surface area available for colonization”. The same author recommended sizes ranging between 0.1 and 0.7 mm (100-700 µm).

Shape is another factor that should be considered when selecting an appropriate media. Support media for AFBR should have a uniform size and shape, in order to achieve a uniform particle fluidization throughout the reactor height without diffusion limitations (Weiland et al., 1991). Another important aspect related to media particle

shape is the spatial distribution of the biofilm. One of the assumptions made when developing models is that the biofilm is uniformly distributed along the carrier, forming a layer of equal thickness. This is a very rough approach since visual evaluation shows biofilm accumulation filling the crevices and holes where shear forces are smaller and bald areas where exposure is greater (Iza, 1991).

The density of the material is another important variable for selecting a media. Density affects the hydrodynamics of the fluidized bed and has a direct relationship with power consumption and, thus, process economy. Iza (1991) reported that for particle density closer to the density of the fluidizing liquid, the superficial velocity for minimum fluidization and 20% expansion become close. Therefore, the hydrodynamic control of the bed is difficult.

The cost of the material is obviously another aspect to consider because it influences the economy of the system (Iza, 1991). Some artificial supports, like open-pores sintered glass, ceramics or plastics show excellent immobilization properties, but are usually extremely expensive. Therefore, the benefits of such media are therefore controversial, because the economical advantages due to the size reduction of the reactor are often overweighted by the high support costs. (Weiland et al. 1991)

Among the chemical properties to consider when selecting the media, the most important are chemical compatibility and adsorption (Iza, 1991). Table 2-1 summarizes the characteristics of an ideal carrier for a fluidized bed.

Table 2-1 Beneficial characteristics of a fluidized bed media (Speece, 1996).

• Withstands physical abrasion • Provides maximum cumulative pore surface and volume area available for colonization by bacteria • Minimizes required fluidization velocity • Enhances non-limiting diffusion/mass transfer • Provides a shielded irregular surface to protect biomass from abrasion

2.2.3. Activated Carbon as Support Media

Granular activated carbon provides an excellent surface for microbial attachment in expanded-bed anaerobic bioreactors. It has an exterior roughness which renders it superior to most other media in microbial sheltering and attachment, GAC media also has the capacity to store substrate until the biomass develops sufficient capacity to metabolize it (Speece, 1996).

The adsorptive properties of activated carbon increase the concentration of soluble organic matter at the interface, thus stimulate biological growth and assimilation (Speece, 1996). Fox, Suidan, and Bandy (1990) reported that more than 20 times as much biomass accumulated on GAC vs. sand when equal-sized particles of both media were employed for microbial attachment in side-by-side comparisons of two expanded-bed reactor treating a prepared solution of 5,000 mg/L acetic acid. Additionally, the steady state data from both reactors revealed that the effluent concentrations of volatile

suspended solids and acetic acid from the sand reactor were 350-700 mg/L compared with 7-40 mg/L from the GAC reactor.

Fox et al., (1990) reported that the majority of the biomass in the GAC and anthracite reactors was attached and these reactors had consistent effluent quality, whereas a sludge blanket on top of the sand reactors was critical to reactor performance and reactor performance deteriorated if the sludge blanket dispersed.

Suidan et al. (1988) suggested that GAC has an adsorptive capacity which accommodates the retention of inhibitory or less biodegradable compounds. Suidan et al. (1991) compared the operational impact of the fluidized bed carriers anthracite and GAC. They reported that the capacity of GAC to adsorb pulse overloads of phenol resulted in no increases in the effluent concentration when compared to the anthracite, with its negligible adsorption capacity.

2.2.4. Biological Solids Yield and Sludge Wasting in AFBR

One of the main problems in fluidized bed reactors is the control of the biomass growth (Iza, 1991). Biofilm development brings about changes in particle size, density and hydraulic drag coefficient (Hermanowicz and Ganczarczyk, 1983). Therefore, biomass growth directly and significantly affects reactor hydrodynamics changing important parameters such as bed porosity and bed height (Hermanowicz et al,. 1990).

Additionally, due to high liquid upflow velocity, non-attached biomass usually leaves the reactor with the effluent; this phenomenon could deteriorate the effluent

quality (Iza, 1991). The biomass growth and formation of thicker biofilms may have two side-effects (Iza, 1991 b):

- Thicker biofilms are not as well attached to the support carrier as thinner ones. Collisions can cause major damage and detachment of big portions of the biofilm, which in turn, can be washed out from the reactor or can promote the formation of granules without carrier particle, depending on the operating conditions.
- Bioparticles with different biofilm thicknesses have different physical properties (volume, density, cross-sectional area), and consequently different fluidization properties (terminal velocity, minimum fluidization velocity, hydraulic drag coefficient, etc.). These differences can lead to bed segregation or mixing, which can affect the system performance. Also, the differences produce a non-homogeneity on the fluidization of the bed which can cause particle washout or bed compaction and stagnant areas. Neither of these conditions is desirable for good operation.

2.3. Solids in Wastewater

One of the most important wastewater characteristics with reference to reactor design, operation, and performance is the presence of suspended solids. As mentioned before, in the proposed system the excess sludge produced in the aeration chamber is recirculated to the AFBR. Therefore, it is important to know not only the ability of the AFBR reactor to degrade such solids, but also the likely negative effects of solids entering, and remaining inside the reactor, on its long term operation. In most

wastewater especially in domestic wastewater organic solids are for the biggest part biodegradable. Therefore, it is the rate of hydrolysis and the solids retention time what determine if there will be an accumulation of biodegradable organic suspended solids (Iza, 1991).

In anaerobic systems, it is expected that entrapped organic solids inside the reactor would become soluble organics as a result of hydrolysis and that the soluble substances would then be converted into methane gas at the end of the anaerobic reactions (Morris and Jewell, 1981).

Research and full-scale experience to date has provided some useful information on the effect of suspended solids on biomass performance and has given some insight on the tolerance of different rector designs to suspended solids influx (Iza, Garcia, Sanz, Hernando, and Fdz-Polanco, 1988).

1. Contact reactor also referred as the anaerobic activated sludge process. The wastewater S.S. concentration tolerated depends primarily on: (I) the type of separation device being used (i.e. membrane; gravity settler, etc.) and (II) the efficiency of the internal mixing system. In general contact reactors can achieve substantial degradation of biodegradable suspended solids.
2. Anaerobic filter (AF) and hybrid anaerobic filter. These reactors can tolerate medium concentration of S.S. Some possible adverse effects are a reduction in specific sludge activity and possible blockages within the packing material. The latter effect is obviously less of a problem in hybrid

AF reactors. Both fully packed and hybrid units can achieve conversion of biodegradable S.S.

3. Downflow stationary fixed film (DSFF) reactor. This reactor design can tolerate “high” S.S. concentrations. However, the bulk of the S.S. pass through the reactor untreated because of the channeled nature of the support and the downflow mode of operation. Some solids may accumulate at the base of the channels and may undergo some degradation. The degradation of these suspended solids in the reactor is influenced by the rate of recycle.
4. Upflow anaerobic sludge blanket (UASB) reactor. In the case of granular UASB reactors, a low S.S. concentration in the influent is preferred and conversion of such low concentrations can be achieved if the S.S. are biodegradable. However, at higher concentrations, influent S.S. can cause granular sludge deterioration. UASB reactors containing flocculent sludge can accommodate higher S.S. concentrations, although the specific sludge activity may be reduced.
5. Fluidized bed (FB) / Expanded bed (EB) reactors. These reactors can tolerate high S.S. concentration in the influent. But even if the S.S. are biodegradable, under certain conditions they could exit the reactor untreated.

2.3.1. Suspended Solids in Anaerobic Fluidized Bed reactors

Compared with other high-rate anaerobic treatment systems the fluidized bed reactors are insensitive to higher suspended solids concentration (Weiland et al., 1991). Saravanane et al. (2001) treated a wastewater from sago mills using a bench scale AFBR. The wastewater had total suspended solids and total volatile solids concentrations of 1410 and 1350 mg/L respectively. These researches obtained an effluent with 222.6 mg/L of volatile suspended solids at organic loading rates of 66 kg/m^3.d.

Yoda, Haittori, and Miyaji (1985) treated primary settled domestic wastewater using an bench scale ANFBR and obtained that the actual gas productions were always smaller than the theoretical gas productions from the total COD removed, but exceeded the potentials from soluble COD removed. Thus, they deduced that methane was derived not only from soluble organics removed but also from organic solids entrapped, which underwent hydrolysis while detained in the reactor. They observed that organic suspended solids introduced into the reactor were entrapped in the upper portion of the fluidized bed to form granular pellets. Later, through a mass balance of organic material the researches also confirmed that organic solids detained in the pellets underwent hydrolysis to soluble organics, which later were converted to methane.

In general, the influx or precipitation of a certain amount of solids within an ANFBR reactor does not affect the global yield, since the bed expands further to hold the extra solids (Marin, Alkalay, Guerrero, Chamy and Schiappacasse, 1999). However, this accumulation of inert solids in the reactor lowers the specific methanogenic activity

of the sludge. Also particulate refractory material will remain in the reactor producing a solids build-up and occupying reactor volume. However, Since such solids, if organic, contribute to volatile suspended solids measurements, they interfere with the common operational practice of equating reactor VSS with the microbial biomass (Iza 1991).

2.4. Aerated Solids Contact Process

The solids contact process involves the production of a mass of microorganisms capable of stabilizing a wastewater under aerobic conditions. The system has the capability of converting the finely divided and dissolved organic matter in wastewater into floc particles, ranging in size from 50 to 200 µm, which can be removed by gravity settling, leaving a relatively clear liquid as the treated effluent. Most activated-sludge processes receive wastewaters that are pretreated by primary sedimentation. Primary sedimentation is most efficient at removing settleable solids, while the biological process is excellent for removing soluble, colloidal and suspended organic substances (Metcalf and Eddy, 2003).

By definition, the basic solids contact process consists of three basic components: an aerator, a liquid-solid separation unit, and a recycle system (Metcalf and Eddy, 2003). In the aerator or aeration tank, contact time is provided for aerating and mixing influent wastewater with the microbial suspension, generally referred as the mixed liquor. Two streams enter the aeration tank. One is the untreated wastewater and the other is the concentrated slurry of microorganism which is being recycled from the

secondary clarifier. The concentration of microorganism in the recycle stream depends on several factors such as, the concentration entering the settler and the rate of the recycle flow in proportion to the raw waste flow rate (Leslie et al., 1980). On the other hand, the concentrations of soluble and particulate organic matter in the wastewater stream depend on the characteristics and nature of the wastewater and its pretreatment. Research performed at the University of New Orleans experimental station at Marrero, Louisiana, USA, demonstrated that most of the total organic material from many municipal wastewaters is in the form of organic particulate material. In the case Jefferson Parish, Louisiana, more than 80% of the TCOD is in the form of organic particles, while only 20% is truly dissolved organic material (Jimenez, 2002).

The solids contact system relies on three basic processes to yield a satisfactory clear final effluent: the synthesis of live microorganisms from the organic matter contained in sewage, the rapid aggregation of particulate matter into settleable solids, and the solid-liquid separation needed to get a clarified final effluent. The first process results from the solubilization of biodegradable organic particulates and the consumption of dissolved organic molecules originally present and/or produced in the solubilization step. Once the dissolved substrate has been depleted, if sufficient dissolved oxygen is provided, bacteria trap food particles (colloidal and suspended particles) through a process of biological flocculation. The particulate and colloidal matter physically entrapped in the floculent biomass is attacked by exocellular enzymes and solubilized to make it available for assimilation by the microorganism. Biological flocculation is the first step in building the compact, readily settleable floc necessary to

optimize the settling characteristics of the mixed liquor. Therefore, this process is the most important in the development of high-quality effluent (Jimenez, 2002). The third process is the final sedimentation of the flocculated particles in a sedimentation unit.

The microorganisms responsible for the degradation of organic matter are aerobic and facultative heterotrophic bacteria. Due to the selective pressure exerted by sludge settling the culture is flocculent with most organisms growing in large clumps of flocs. Some higher organisms, such as protozoa abound in the sludge, feeding upon bacteria. Fungi can also be present sometimes. However, they are considered a nuisance due to its filamentous morphology which prevents the formation of dense floc, thus reducing the settling velocity (Leslie et al., 1980).

An important feature of the solids contact process is the short hydraulic retention time (HRT) in the aeration tank. The HRT can as short as one hour or even less. However, the detention time in the aerator depends on several factors especially on the wastewater characteristics and the effluent requirements. Jimenez (2000) operated a pilot plant comprised of: a trickling filter, an aerated solid contact tank, and a secondary clarifier treating domestic sewage. The researcher reported that the minimum ASCC hydraulic residence time in which bioflocculation occurs satisfactorily as to produce final effluent SS concentrations of less than 20 mg/L is 15 min. The same author stated that in order to have a more stable operation the minimum hydraulic detention time recommended is 20 min. Later, in 2002 Jimenez performed some modifications to its experimental unit. The most important modification was the addition of a rotating fine

screen and bypassing the trickling filter unit to operate the aeration chamber as an activated sludge reactor with a short hydraulic detention time. In this research Jimenez (2002) varied the HRT in the aeration chamber from as low as 5 minutes up to 60 minutes, and obtained a final effluent with less than 30 mg/L of suspended solids using an HRT as low as 10 minutes, and 88% removal efficiency of SS in 30 minutes of flocculation. The researcher also obtained 50% and 86% removal efficiencies of colloidal and particulate COD respectively, with a HRT of 30 minutes in the ASCC.

As mentioned before, oxygen is utilized by aerobic microorganisms to oxidize the organic matter present in the wastewater. Based on the previous statement it is easy to understand that proper aeration is essential for optimum operation of the system. Air can be introduced to the aeration tank by diffusers or nozzles located near the bottom of the tank or by mechanical mixers which entrain air to the system by producing turbulence at the air-liquid interface. Factors affecting the oxygen transfer are bubble size, diffuser air rate, diffuser placement, velocity of the surrounding medium, and impeller speed and size (Syed et al., 1994). The rate at which oxygen is consumed by microorganisms in the biological reactor is called the oxygen utilization rate. The oxygen utilization rate is a function of the both the wastewater and the reactor characteristics (Peavy, Rowe, and Tchobanoglous, 1985). Jimenez (2000) obtained excellent final effluent quality with dissolved oxygen levels as low as 0.5 mg/L; however, for ASCC design, DO levels between 1.0 and 1.4 mg/L were recommended to produce a very good final effluent.

As explained before, the mixed-liquor suspended solids from the aeration tank must be settled in a sedimentation basin to produce a well-clarified effluent. The solids-liquid separation unit is also referred to as the secondary clarifier. The secondary clarifier in general must perform two basic functions: provide clarification to produce a high-quality effluent and provide thickening of settled solids. To fully accomplish its functions the sedimentation tank must have enough depth, so that the solids are not lost in the effluent and, at the same time, there is storage for the settled solids for thickening and maintaining an adequate sludge blanket. If sufficient sludge blanket is not maintained, unthickened sludge will be returned to the aerator and excessive sludge will have to be handled and treated (Syed et al., 1994). Part of the settled sludge is returned from the clarifier to the aeration tank to maintain the desired food-to-microorganism ratio. The most common operational return flow range is 20-30 percent of the average inlet flow. However, the return flow requirement is determined from settling tests of the MLSS. The excess sludge is wasted either from the effluent line of the aeration tank of from the return sludge line (Syed et al., 1994).

2.4.1. Excess Sludge

The major byproduct from colloidal particle flocculation and from the aerobic degradation of soluble organic matter is excess sludge, commonly referred to as secondary sludge. In activated sludge systems this secondary sludge contains appreciable amounts of insoluble organic matter that have been flocculated by the action of bacteria (Syed et al., 1994).

So much sludge is produced in aerobic systems that its disposal represents a major expenditure in wastewater treatment plants. Part of the expenditure arises from the need to stabilize and dewater the sludge prior to disposal. The main purposes of stabilization are to reduce pathogens, eliminate offensive odors and control the potential for putrefaction of organic matter. Sludge dewatering is necessary to remove moisture so that the sludge cake can be transported and can be composted or disposed by landfilling or incineration. In general, the problems involved with handling and disposing the sludge are complex. According to Peavy et al. (1985) sludge disposal facilities usually represent 40 to 60 percent of the construction cost of wastewater-treatment plants, accounts for as much as 50 percent of the operating cost, and are the cause of a disproportionate share of operating difficulties. The problem of sludge generation is especially important in developing countries that lack the technology and expertise needed. Therefore, it is important to find new technologies or system configurations that help to reduce the amount of sludge to be treated and/or disposed. This need is one of the justification for this research project.

3. EXPERIMENTAL SETUP AND DESIGN

This study is part of an extended experimental program aimed to determine the feasibility and efficiency of combined anaerobic/aerobic treatment of domestic wastewaters. The project started with Corzo (2001). She studied the feasibility and efficiency of chemical and biological flocculation of the effluent of an anaerobic fluidized bed reactor (AFBR) treating municipal wastewater. Subsequently Bustillos (2002) investigated the effect of varying the hydraulic retention time (HRT) in the aerobic solids contact chamber (ASCC) to improve the quality of the final effluent of the AFBR/ASCC system. The present study has the main objective of evaluating the efficiency of pre-digestion of solids in the anaerobic fluidized bed reactor. The experimental program will continue, and at present, Mr. Eudomar Silva is studying the efficiency of the system using an UASB reactor instead of an AFBR.

The present research, like its predecessors, was developed and carried out utilizing the pilot scale plant located within the University of New Orleans facility at the Marrero Municipal Wastewater Treatment Plant, 6250 Lapalco Boulevard, Marrero, Louisiana 70072.

For the present study the AFBR reactor was operated and monitored from December 2002 to December 2003. Total COD, total suspended solids (TSS), volatile

suspended Solids (VSS), biogas generation, CH_4 generation, Nitrates concentration were parameters measured during the experimental program.

3.1. Pilot Plant Description

The pilot plant is a combined aerobic/anaerobic system. The major components of the system are: a rotating screen or rotational strainer, an anaerobic fluidized bed reactor, an aerated solid contact chamber, and a secondary clarifier. Figure 3-1 shows a diagram of the pilot plant

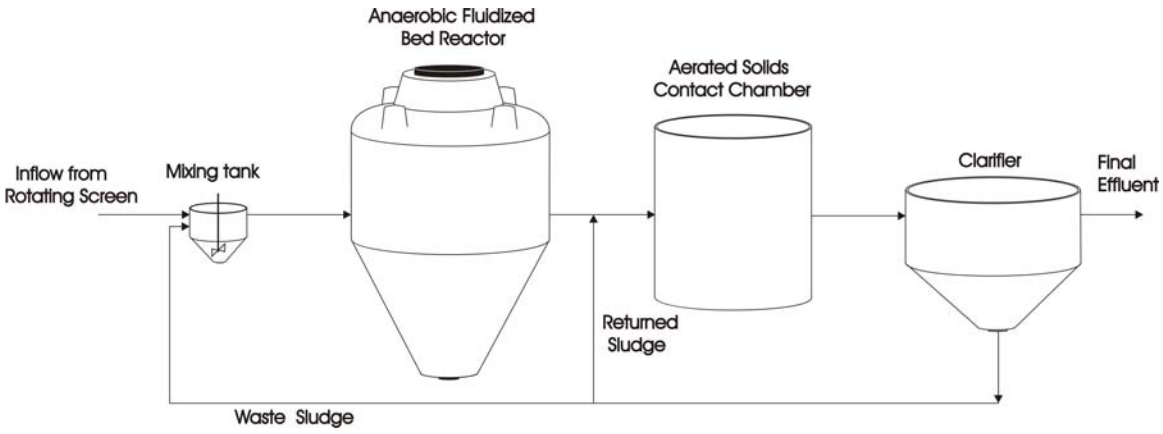

Figure 3-1 Pilot plant diagram

3.1.1. Feeding System

Wastewater is pumped from the grit chamber splitter box by a 372.5 W (½ hp) centrifugal pump (No. 1). (Specifications of each piece of equipment, with its respective number are presented in Table 3-3). The inlet of the pump No. 1 is connected to a straining device to remove coarse material. The straining device consists of a perforated

0.91 m (3 ft) section of 10.2 cm (4 in) diameter PVC pipe. The orifices are 5.08 cm (2 in) in diameter. The perforated pipe is wrapped in “chicken wire” of 9.5 mm (3/8 in) mesh size. The whole straining device is protected by a 20.32 cm (8 in) diameter and 3 m long PVC encasement pipe. Figure 3-2 illustrates the straining device.

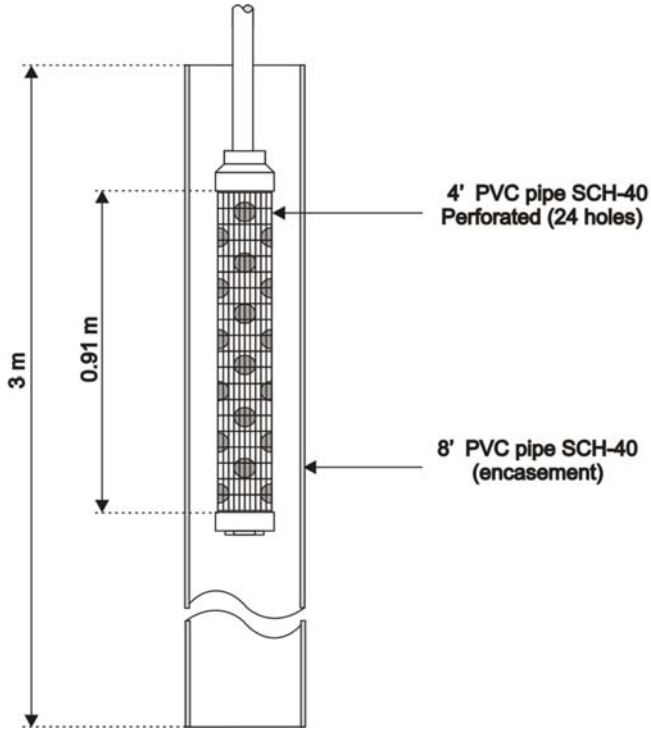

Figure 3-2 Strainer device

Pump No. 1 delivers the wastewater through 50 m (164 ft) of 2.54 cm (1 in) diameter PVC pipe to a rotational strainer (No. 2). The flow handled by pump No. 1 is about 3.7 L/s (3,500 GPH). This flow is higher than the operational flow of the pilot plant which is about 0.4 L/s (133 GPH). Consequently, excess wastewater is wasted from the rotational strainer to one of the full-scale plant’s primary clarifiers.

The rotational strainer removes solids larger than 0.5 mm. The unit has a blade assembly located along its frontal part in such a way that the blade is free to move and conform to the contour of the rotating screen cylinder. This blade, which is the full length of the cylinder, rides in contact with the surface of the rotating screen cylinder removing the solids and channeling them to a collection tank for future disposal.

The rotating screen's effluent is pumped with a 74.6 W (1/10 hp) centrifugal pump (No. 3) to a 120-L distribution tank located on the roof of the pilot plant. An electric 186.4 W (¼ hp) drum mixer (No. 4) continuously stirs the contents of the distribution tank in order to prevent the sedimentation of solids and obtain a homogeneous wastewater.

Wastewater flows by gravity from the distribution tank to a 57 L (15 gal) conical bottom tank (mixing tank), where it is mixed with the sludge wasted from the secondary clarifier. These two streams are mixed by a 14.9 W (1/50 hp) submersible pump (No. 5) located inside the tank. In addition to mixing the streams, the turbulence created by the submersible pump No. 5 prevents solids sedimentation in the mixing tank. The flow of screened wastewater to the mixing tank is controlled by a float valve, which is set to have a constant volume of 37.85 L (10 gal) in the mixing tank. This volume allows a proper blending of the wasted sludge (which has high concentration of solids) with the screened wastewater (which has low concentration of solids), thus minimizing solids shock loads to the AFBR.

The wastewater and wasted sludge mixture is fed to the AFBR by a diaphragm pump (No. 6). The flow rate delivered to the AFBR is controlled by adjusting the pump

settings. The flowrate fed to the AFBR was maintained at 125 L/h (33 GPH) throughout the experimental phase. It is important to highlight that not all the wastewater and sludge mixture was fed to the AFBR. Part of this mixture was used to feed first an anaerobic upflow packed filter (AUPF), which was later replaced by an upflow anaerobic sludge blanket (UASB) reactor. The flowrate to these anaerobic units (AUPF and UASB) was maintained at 80 L/h (21.1 GPH). Therefore, only 61 % of the wastewater-sludge mixture in the mixing tank was fed to the AFBR, the rest (39%) was fed to the alternative reactors (AUPF and UASB). The effluent of these alternative units was sent to the pilot plant final effluent discharge line.

3.1.2. Granular Activated Carbon Anaerobic Fluidized Bed Reactor

An anaerobic fluidized bed reactor with granular activated carbon as support media was used throughout the course of this study. The anaerobic reactor is a cylindrical tank with a 60-degree conical bottom. The tank is made of medium density polyethylene. It has a nominal capacity of 400 L (110 gal.), a diameter of 0.86 m (33.85 in) and a height of 1.16 m (45.67 in). For monitoring purposes several sampling ports are arranged along the experimental AFBR allowing media and solids samples to be taken. The gas produced in the reactor was collected and stored in a collection-measurement tank. Figure 3-3 shows the AFBR with its sampling ports.

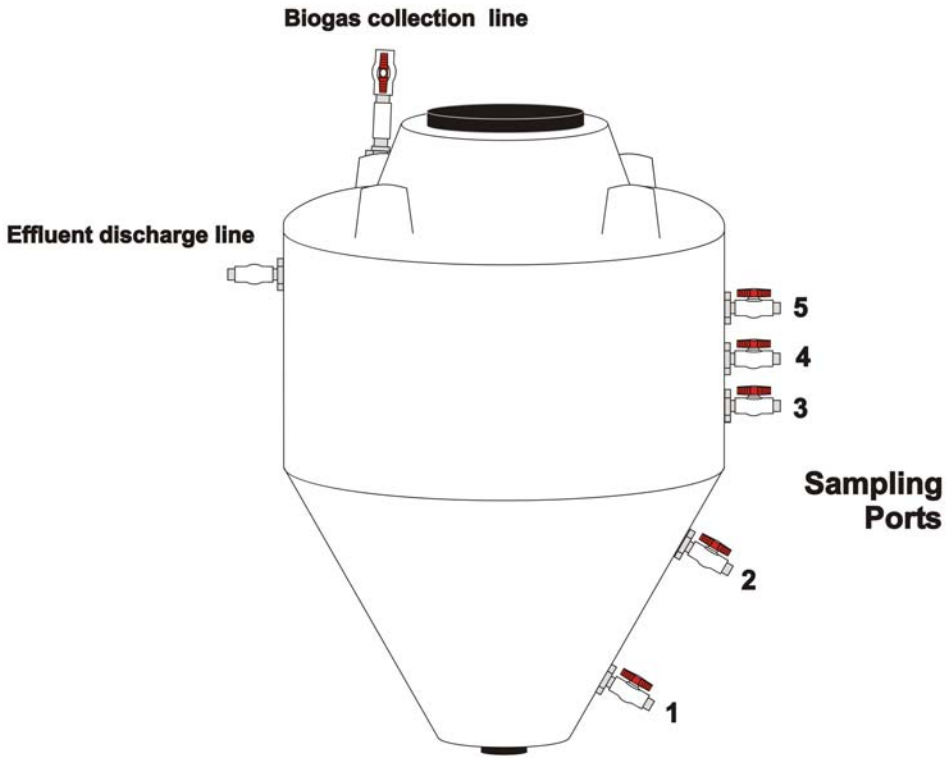

Figure 3-3 Anaerobic Fluidized Bed Reactor

The AFBR has an internal recirculation system that withdraws reactor effluent from near the top and pumps it to the bottom of the reactor. The purpose of the recirculation, as explained on the literature review, is to increase the upflow velocity to fluidize the bed. The recirculation is achieved utilizing a 372.5 W (½ hp) centrifugal pump (No. 7). The wastewater and wasted sludge mixture is fed from the mixing tank to the recycle line of the AFBR by the diaphragm pump No. 6. This configuration assures that the influent is introduced to the reactor through its bottom. The AFBR effluent is discharged near the top of the reactor. Figure 3-4 is a schematic representation of the AFBR showing the recirculation system.

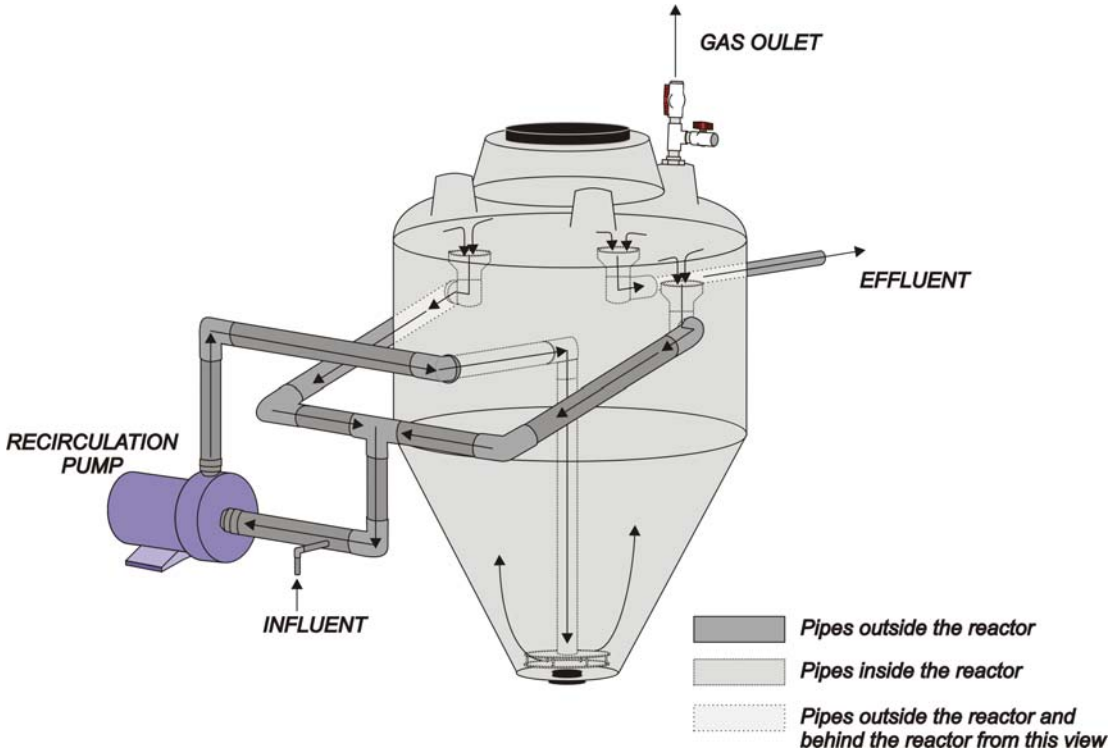

Figure 3-4 Schematic representation of the AFBR

Table 3-1 Characteristics of the activated carbon used as support media (Corzo, 2001)

• Density: 0.48 g/cm^3
• Mesh: 40/80
• Surface Area: 1150 m^2/g

3.1.3. Aerated Contact Chamber

The effluent of the AFBR is fed directly to the aerated contact chamber (ACC) by gravity. Therefore, the ACC influent flowrate is the same as that of the AFBR. Two different aeration contact chambers (ACC) were used to provide two different hydraulic retention times in the aerated solids contact process. The ACCs consist of polyethylene cylindrical tanks one of 202 L (53 gal) and the other of 114L (30gal). The aeration chambers are equipped with a fine bubble diffuser system fed by a 559.3 W (¾ hp)

compressor (No. 8). The compressor provides air to maintain the required dissolved oxygen levels in the aeration chamber and the velocity gradient for uniform mixing. The air injected to the ACC was regulated by a valved rotameter. Table 3-2 shows the specification of the air diffuser system.

Table 3-2 Specifications of the air diffuser system (Bustillos, 2002)

- Type of diffusers: heat bonded silica fine-pore diffusers
- Number of diffusers: 6
- Shape of diffusers: rectangular
- Length of diffusers: 15 cm
- Width of diffusers: 4 cm
- Max. pore size: 80 µm
- Estimated bubble size: 0.5-2.0 cm

The effluent from the AFBR and the sludge recycled from the clarifier are fed to bottom of the ACC where they are mixed with the reactor contents as a result of the turbulence created by the air injected. The mixture inside the ACC known as mixed liquor leaves the ACC through a center well as depicted in Figure 3-5.

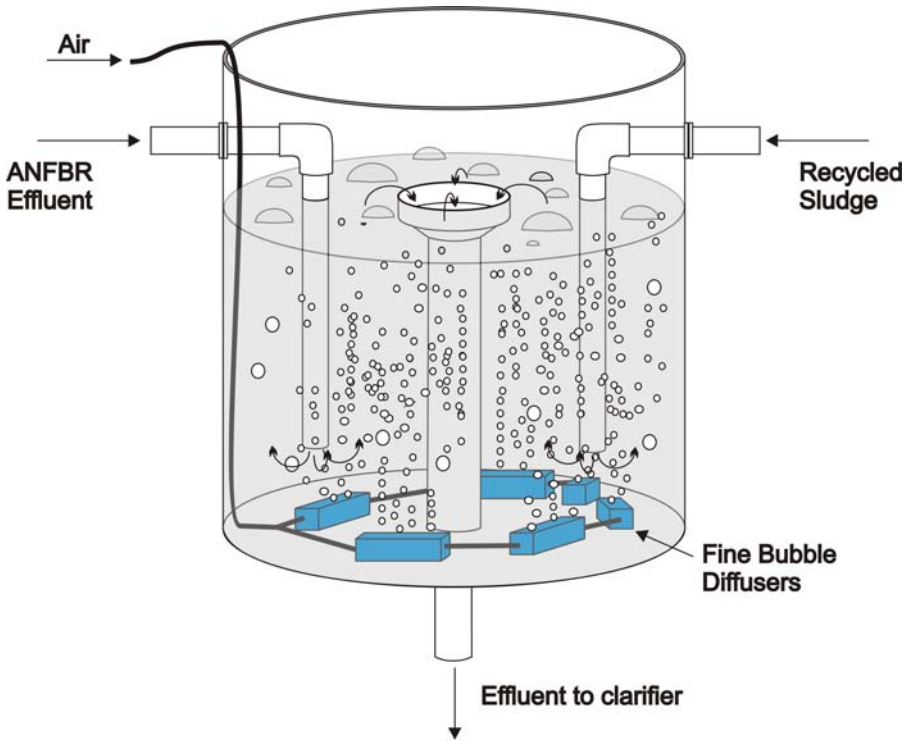

Figure 3-5 Graphical Representation of the Aerated Solids Contact unit

3.1.4. Clarifier

The secondary clarifier consists of a 280 L (70 gal) polyethylene tank with a conical bottom section. The clarifier receives the water from the aerated contact chamber through a 3.81 cm (1 ½ inch) PVC pipe. The water is introduced to the clarifier tangentially in an 8 20.3 cm (in) diameter center well to reduce the inflow energy and enhances flocculation. The unit also has a rotary arm or scrapper which is connected to a 1 rpm gear motor (No. 9). The function of the arm is to scrape the conical section of the tank. The clarified effluent is collected by three 38 mm (1 ½ in) PVC pipes located radially on top of the clarifier, and is finally discharged into the plant's final effluent line. As indicated on Figure 3-6, a portion of the sludge in the clarifier is recycled to the aeration contact chamber; this is done using a 14.9 W (1/50 hp) submersible pump (No.

10) which activation is regulated by a timer (No. 11). Another portion of the sludge is sent to the mixing tank, where it is mixed with the screened wastewater and fed to the AFBR. Sludge is pumped from the clarifier to the mixing tank using a 14.9 W (1/50 hp) submersible pump (No. 12) which activation is also regulated by a timer (No. 13).

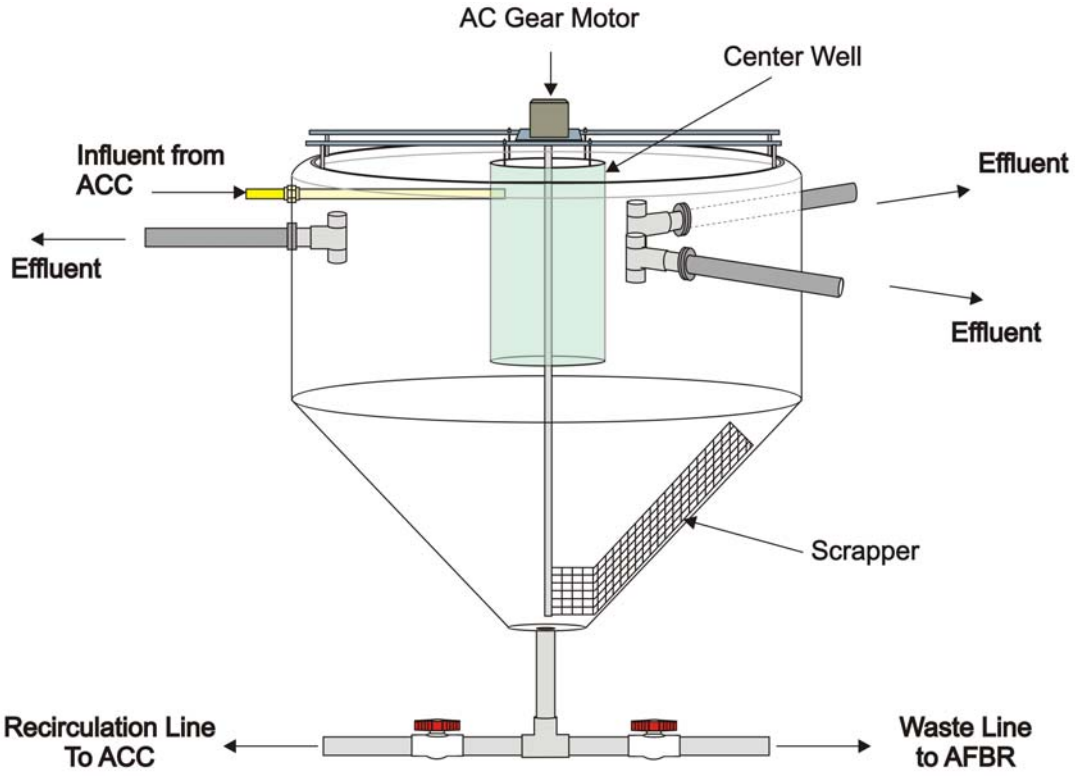

Figure 3-6 Graphical Representation of the Clarifier

Table 3-3 Description of the electric equipment used at the pilot plant

Equipment number and name	Manufacturer / Model	Characteristics
(1) Centrifugal pump	TEEL / 3P551	Self-priming, ½ HP, 115/230 volts, 58 GPM at 10 ft of head
(2) Rotational strainer	WaterLink Rotostrainer® / RSA2512UBCR	Rotating cylinder screen, 1/3 HP, 120 volts
(3) Centrifugal pump	TEEL / 1P809	Submersible, 1/10 HP, 115 volts, 900 GPH at 1 ft of head
(4) Open drum mixer	Neptune mixer company / B-10	¼ HP, 115/220 volts, 316 Stainless Steel Shaft and 3 Blade Propeller
(5) Centrifugal pump	TEEL / 1P808	Open air/submersible 1/50 HP, 115 volts,400 GPH at 1 ft of head
(6) Diaphragm pump	Cole-Parmer / 76302-50	Single head, 115 volts, 16.5 GPH of maximum flow 125 strokes / minute
(7) Centrifugal pump	TEEL / 2P390	Self-priming, ½ HP, 115/230 volts, 2280 GPH at 10 ft of head
(8) Air compressor	GAST / 4F742	¾ HP, 115/230 volts, free air flow at 10 Inches Vacuum 6.3 CFM
(9) Gear motor	Dayton® / 2Z804	AC Parallel shaft gearmotor, 115 volts
(10) Centrifugal pump	TEEL / 1P808	Open air/submersible 1/50 HP, 115 volts,400 GPH at 1 ft of head
(11) Timer	OMRON® / 2A179	Repeat cycle timer independent on/off times 120/240 volts
(12) Centrifugal pump	TEEL / 1P808	Open air/submersible, 1/50 HP, 115 volts,400 GPH at 1 ft of head
(13) Timer	OMRON® / 2A179	Repeat cycle timer independent on/off times 120/240 volts
(14) Submersible pump	LittleGIANT / 2P352	Submersible, 1/6 HP, 110 volts, 20 GPM at 1 ft of total head

It is important to highlight that the sludge wasted from the clarifier is discharged to the mixing tank intermittently. The frequency and duration of the sludge discharge cycle depends on the ACC. Therefore, it varied during the whole experimental phase. A typical value is 1.5 liters of sludge discharged to the mixing tank every 2 hours.

3.1.5. Biogas Collection System

Following the recommendations of Metcalf and Eddy (1972) a system of tanks filled with a retaining fluid were used to measure the volume of biogas produced in the AFBR. The retaining fluid used was a saturated sodium chloride solution containing 5% H_2SO_4 and methyl orange for color (Metcalf and Eddy, 1972).

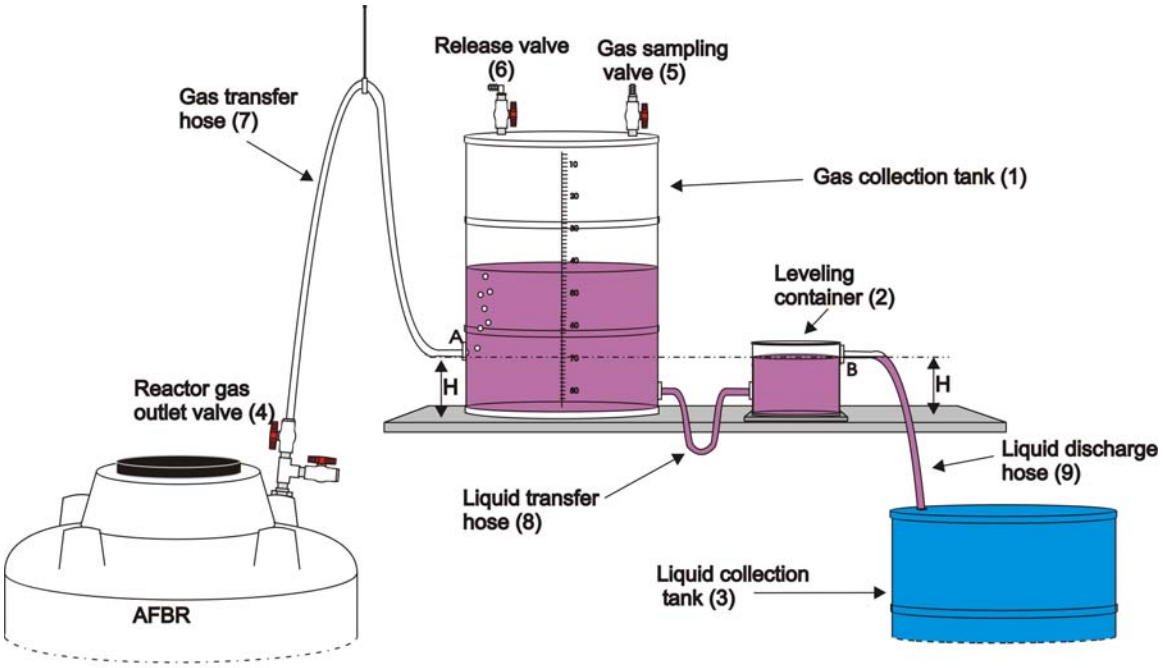

Figure 3-7 Schematic representation of the biogas collection system

As depicted in Figure 3-7 the biogas collection system consists of:

- A gas collection tank (1) which is a 114-L (30 gal) translucent closed-head drum made of low density polyethylene. To facilitate level readings, the tank is graduated every 2 liter. The tank is closed to the atmosphere, but has two ball valves. The gas release valve (6) is used to release the collected gas, while the gas sampling valve (5) is used as a collection port for taking gas samples when needed, and filling the tank with the retaining fluid. As shown in Figure 3-7, the gas collection tank is connected in its lower portion (point A) to the reactor's gas outlet through a transparent flexible hose (7). Additionally, the tank is connected to the leveling container (2) through another transparent flexible hose.
- A leveling container (2) which is an 8-L (2.1 gal) polyethylene container open to the atmosphere. As mentioned previously, the leveling container (2) is connected to the gas collection tank (1) through a liquid transfer hose (8). The leveling container also has an overload connection (B) to discharge the excess liquid and maintain a constant level of retaining fluid. The overload connection is joined to the liquid collection tank (3) through the liquid discharge hose (9).
- A liquid collection tank (3) which is a 200-L (55 gal) low density polyethylene drum open to the atmosphere (An explanation of how this gas collection system works is presented in Section 3.23)

3.2. Sampling

The sampling phase was initiated in January 2003 and lasted through December 2003. Sampling was done as often as possible depending on external factors such as weather and plant operating conditions.

3.2.1. Water Samples

Water samples were taken at the mixing tank (AFBR influent), the AFBR effluent discharge line, the clarifier effluent discharge line, and the sludge recycle line.

Since one of the main objectives of this research is to setup a mass balance on solids in the AFBR, it was important to collect truly representative samples of the AFBR influent and effluent. Consequently, it was decided to work with 24-hour composite samples in order to overcome the variations of the streams due to the intermittent discharges of sludge and the hourly variations of the pilot plant influent. Three automatic composite wastewater samplers from Global Water, model WS300, were used to collect the samples.

As mentioned before, the excess sludge from the clarifier is discharged to the mixing tank (AFBR influent) intermittently. Once in the mixing tank the sludge is rapidly mixed with the effluent of the rotational strainer (plant influent). The typical sludge discharge cycle has a length of 2-4 seconds and is repeated every 2-4 hours. Therefore, the samplers were set up to pump 50 milliliters of sample every 10 minutes during a 24-hour period. For sample preservation, sulfuric acid was added to the samplers collection tanks at the beginning of the collection cycle to ensure that the pH of the final samples was below 2, as recommended in the Standard Methods (AWWA,

1995). After being collected, the samples were taken to the Environmental Engineering Laboratory and stored in a refrigerator at a temperature between 4 and 6°C for future analysis. The samples were stored in glass bottles of approximately 120 ml each.

For the AFBR effluent, the blended or raw samples and their supernatants or settled samples were stored and analyze separately. The supernatant corresponds to the sample decanted after several minutes of settling, e.g. the sample taken directly from the wastewater sampler tank without agitation. The blended sample corresponds to the homogenized sample, which is taken after agitating vigorously the wastewater sampler tank.

3.2.2. Parameters Measured

Different water quality analyses were performed at the Environmental Laboratory in the Center for Energy Resources Management (CERM). The parameters tested to each sample of water collected were total suspended solids (TSS), volatile suspended solids (VSS), and total chemical oxygen demand (TCOD). Additionally, some nitrate determinations were performed in order to establish the occurrence of denitrification inside the AFBR.

2.2.2.1 Total Chemical oxygen demand

COD is defined as the amount of a specified oxidant that reacts with the sample under controlled conditions (AWWA, 1999). The amount of oxidant consumed is expressed in terms of its oxygen equivalent. Therefore, the COD expresses the amount of oxygen necessary to chemically oxidize the organic matter present in the sample.

Since the COD is an easy and quick test, it is often used to represent the amount of pollutants presents in wastewaters. To determine the chemical oxygen demand the wastewater samples were first homogenized by mixing using a Stir-Pak general purpose mixer model 04554-00. Then the samples were analyzed according to the method 5220D of the Standard Methods (APHA, 1999).

2.2.2.2 Total and volatile suspended solids

The total suspended solids (TSS) analysis is used to quantify the amount of suspended matter (organic and inorganic) present in the sample. TSS were determined following method 2040D of the Standard Methods (APHA, 1999).

The volatile suspended solids (VSS) test is used to quantify the amount of total organic solids (biodegradable and non-biodegradable) present in a sample. It also quantifies the amount of fixed or inorganic solids when combined with the TSS test. The VSS determination was done following method 2540E of the Standard Methods (APHA, 1999). Like for the COD test, before the analysis the samples were homogenized by mixing using a Stir-Pak general purpose mixer model 04554-00.

2.2.2.3 Nitrate concentration:

Nitrate concentration was measured in the AFBR influent and effluent four times. These analyses were done in order to determine if denitrification was occurring inside the unit. The nitrate concentrations were determined according to method 4500-NO_3-B of the Standard Methods (APHA, 1999).

3.2.3. Biogas Collection and Analysis

Since it is an indicator of the anaerobic activity, the methane concentration in the biogas was monitored throughout the whole experimental phase. The methane concentration was measured utilizing a portable landfill gas analyzer model LMS manufactured by CEA Instruments, Inc.

Additionally, during the last three months of the experimental phase, biogas produced from the AFBR was collected and quantified. Biogas sampling was done on a 24-hour basis. Usually, the wastewater samplers and the gas collection system were set up at the same time to collect the wastewater samples and the corresponding biogas produced after a 20-24 hours period.

The procedure followed to set-up the gas collection system for collecting and measuring the biogas produced by the AFBR in the next paragraph, where all the numbers and symbols refer to Figure 3-7

a. The reactor's gas outlet valve (4) was closed.
b. The leveling container (2) was raised until the level of liquid in it was equal to the top surface of the gas collection tank (1).
c. The gas release and gas sampling valves (5, 6) were open.
d. The retaining liquid was pumped to the gas collection tank (1) from the liquid collection tank (3). The liquid was pumped through a flexible hose connected to the gas sampling valve (5) by using a 1/6 hp submersible pump (14 in Table 3-3).

e. Once the gas collection tank (1) was completely filled with the retaining fluid, the gas sampling and gas release valves (5,6) were closed.

f. The leveling container was lowered until its overload connection (B) was at the same height as the connection of the gas transfer hose (7) to the gas collection tank (point A).

g. Finally the reactor gas outlet valve (4) was open to let the biogas enter the gas collection tank.

It is important to highlight that since the gas collection tank (1) is interconnected to the a leveling container (2), and point A is at the same height H as the surface of the liquid in the leveling container (point B), both points (A and B) are at the same pressure, which is atmospheric pressure (the leveling container is open to the atmosphere). Therefore, the biogas leaves the AFBR at atmospheric pressure.

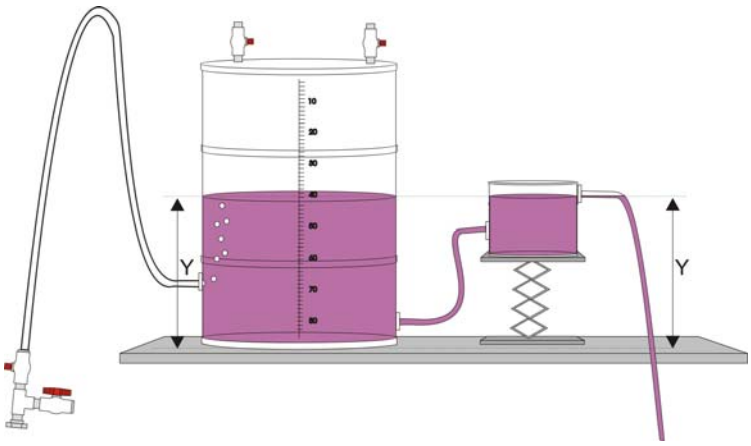

Figure 3-8 Schematic representation of the biogas measuring procedure

As the biogas produced enters the gas collection tank (1) it displaces the retaining liquid which is collected in the liquid collection tank. Once the collection cycle is ended the volume of gas produced is measured by closing the reactor's gas outlet valve (4) and raising the leveling container until the level of liquid in it and the level of liquid in the gas collection tank are at the same height (Figure 3-8). After measuring and recording the volume of biogas a thermometer is introduced through the gas sampling valve (5) to record the corresponding temperature of the gas.

3.2.4. AFBR Solids

In order to monitor the accumulation and distribution of solids inside the AFBR, TSS and VSS concentrations along the reactor height were measured in three opportunities. Additionally, since the behavior and fate of entrapped solids in the AFBR is a major objective of this research, an experiment was run to determine the potential for digestion of the organic matter present in the solids entrapped in the reactor. The procedure followed is indicated below.

First, samples (sludge, media and water) were taken from each of the five reactor ports. Then, equal volumes of the five samples were mixed to obtain a volumetric composite sample. This composite sample was prepared in order to obtain a representative sample of the reactor contents. Next, an aliquot of about 80 ml of sludge was taken from the composite sample and placed in a 250 ml Erlenmeyer flask. TSS and VSS were measured in the sludge aliquot. After that, the flask was sealed and incubated at a temperature of 37°C under continuous stirring. A magnetic stirrer was used to mix the contents of the flask continuously. In order to eliminate the potential

inhibitory effect of the air present in the overhead space of the flask, methane gas was blowed in to the flask to displace the air. Figure 3-9 shows a schematic representation of the test configuration. The flask was kept in the incubator for a period of 32 days. After that time, the sample was removed from the reactor and subjected to TSS and VSS determinations in order to determine the amount of solids degraded during the incubation time.

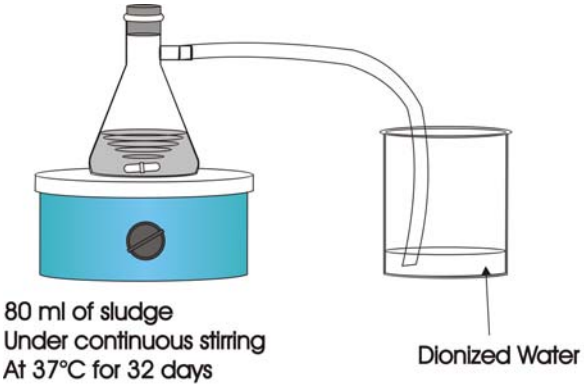

Figure 3-9 Schematic representation of the solids digestion test

3.3. Field Measurements

3.3.1. Plant Flow Rate

The flow rate to the AFBR and the alternative units (AUPF and UASB) was measured everyday to ensure a constant flow rate through the units along the whole experimental phase.

3.3.2. pH and Redox Potential

These two parameters where measured twice a week using a WTW pH meter, model 330. The parameters were measured to a sample collected from the internal recirculation line of the AFBR.

4. RESULTS AND DISCUSSION

Water analyses were made on the 24-hour composite samples of the influent and effluent of the AFBR reactor. Based on the settling nature of the particles in the effluent, it was decided to collect and analyze separately the mixed and settled effluent samples.

4.1.1. AFBR Influent

As explained before, during the experimental phase the AFBR unit was fed with a mixture of screened wastewater from the rotary strainer and excess sludge wasted from the bottom of the secondary clarifier. The characteristics of this influent are presented in Table 4-1.

Table 4-1 Characteristics of the AFBR influent

Parameter	Value
Total COD, mg/L	301
Total suspended solids, mg/L	144
Volatile suspended solids, mg/L	126

Figure 4-1 shows that there is a linear relationship between TCOD and TSS in the AFBR influent. A linear regression analysis generated the following equation:

$$TCOD = 2.0848 \times TSS \qquad \textbf{Eq. 4-1}$$

Where: TCOD and TSS are in mg/L

The coefficient of determination (R^2) for the data is 0.72 which means that 72% of the variability of the data could be explained by equation Eq. 4-1. This equation gives a ratio of 0.48 g of TSS per g of TCOD.

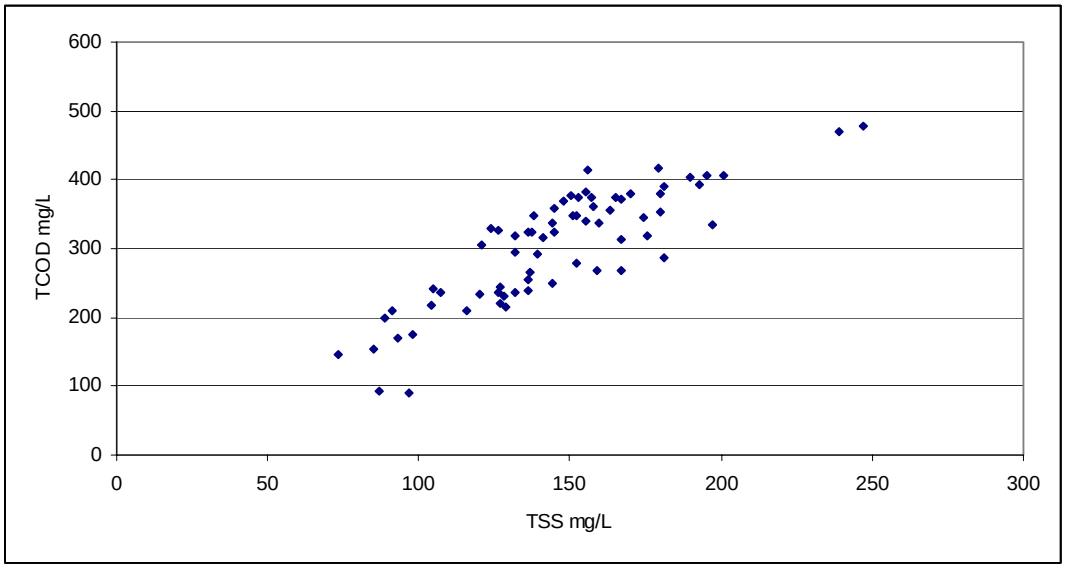

Figure 4-1 AFBR influent TCOD vs. AFBR influent TSS

In the same fashion, Figure 4-2 shows the relationship between TCOD and VSS in the AFBR influent. There is also a clear correlation between the TCOD and VSS. A linear regression analysis of the TCOD and VSS data generated the following equation:

$$TCOD = 2.2738 \times \mathrm{VSS} \qquad \textbf{Eq. 4-2}$$

With R^2 =0.75.

In this equation TCOD and VSS are in mg/L. The equation gives a ratio of 0.44g of VSS per 1g of TCOD.

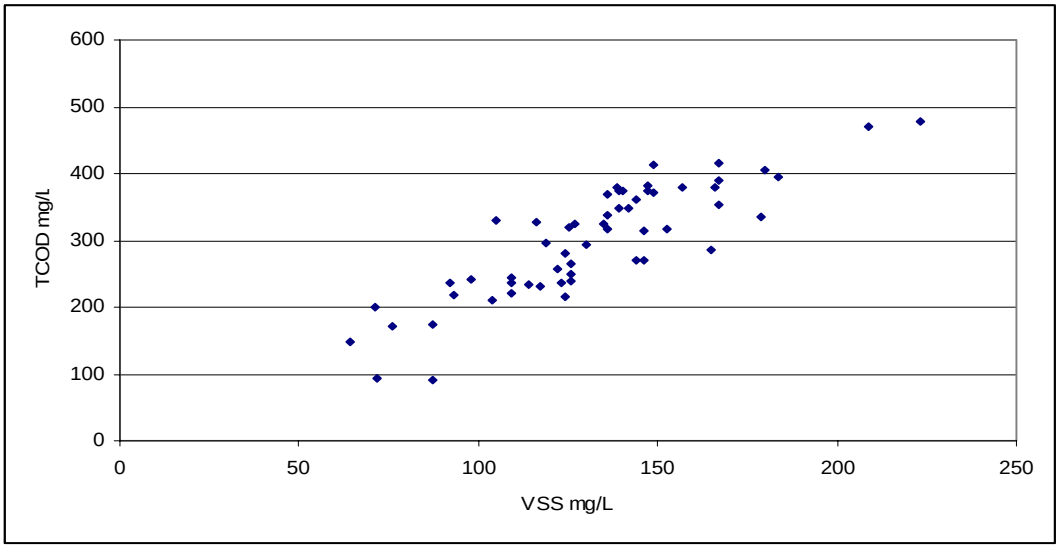

Figure 4-2 AFBR influent TCOD vs. AFBR influent VSS

4.2. Performance of the Anaerobic Fluidized Bed Reactor

Through the whole experimental phase the flowrate delivered to the AFBR was maintained at around 125 L/h (33 GPH). Therefore, the hydraulic retention time of the reactor was 3.4 h, the average organic load applied was 2.12 kg TCOD/m^3.d, and the average solids load was 1.01 kg TSS/m^3.d.

Table 4-2 AFBR raw effluent average removal efficiencies

Measured Parameter	Removal
Total chemical oxygen demand (TCOD)	23 %
Total suspended solids (TSS)	31.2 %
Volatile suspended solids (VSS)	32 %

Table 4-2 shows the average percent removals of TCOD, TSS and VSS obtained in the AFBR based on the mixed effluent. Comparing these results with the typical values obtained with anaerobic fluidized bed reactors, we can conclude that the reactor was not working at optimum conditions. However, no major changes in the AFBR operational parameters such as HRT were done because it would have affected the HRT in the ACC. The ACC and final clarifier were operated and controlled by Miss. Jacqueline Luque. She was studying the effect of the different hydraulic retention times in the ACC-clarifier system. Moreover, the main objective of this research was to study the performance of the AFBR in the system not to find its optimum operational conditions.

Table 4-3 shows the average percent removals of TCOD, TSS and VSS obtained in the AFBR when comparing the influent to the settled effluent.

Table 4-3 AFBR settled ffluent average removal efficiency

Measured Parameter	Removal
Total chemical oxygen demand (TCOD)	45 %
Total suspended solids (TSS)	60.9 %
Volatile suspended solids (VSS)	61.6 %

These results can be better analyzed using the definition given by Adrianus et al., (1994). The authors defined the removed load as the load of organic material that is either converted into sludge or into methane and therefore is removed from the liquid

phase. They also defined the digested load as the load that is actually converted into methane. Following the previous definition the removed load corresponds to the removal obtained considering the settled effluent while the degraded load corresponds to the removal considering the raw effluent.

From Table 4-2 and Table 4-3, it can be concluded that on the average 45% of the TCOD fed to the reactor was removed. Only 23% was degraded (converted to methane), however. A similar analysis can be done regarding TSS and VSS. Table 4-4 shows the average percent removals and degradation in the AFBR.

Table 4-4 Average performance of the AFBR

Parameter	Percent Degraded	Percent removed
TCOD	23%	45 %
TSS	31.2 %	60.9 %
VSS	32 %	61.6%

Figure 4-3 presents the performance of the anaerobic unit, regarding TCOD. The AFBR mixed effluent TCOD is shown on the vertical axis and the influent TCOD is on the horizontal axis.

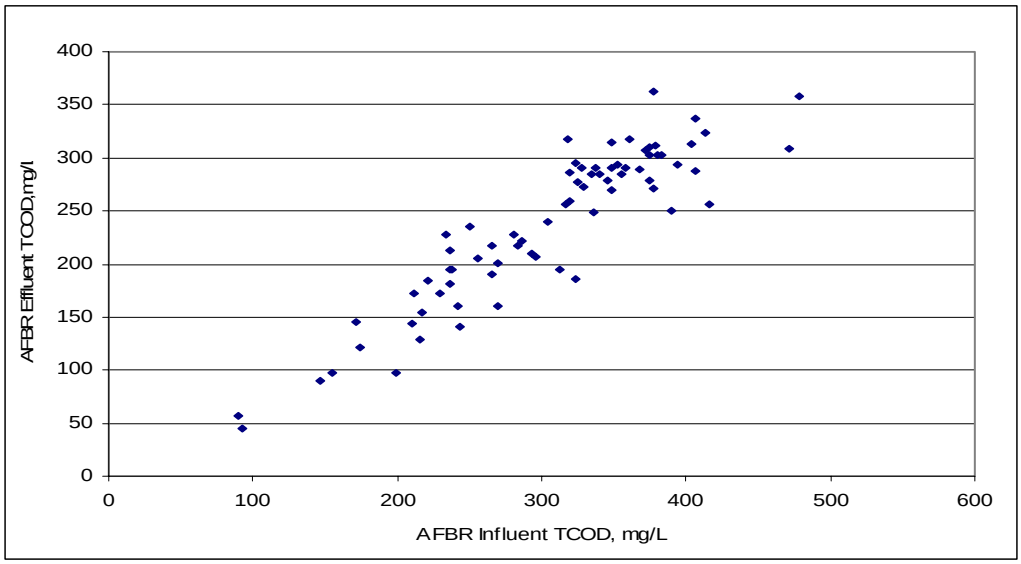

Figure 4-3 AFBR Effluent TCOD vs. AFBR influent TCOD

It can be seen that there is a linear correlation between the influent and effluent TCOD. A linear regression analysis yielded a coefficient of determination (R^2) of 0.84, and the following regression equation

$$TCOD_{Mixed\ Effluent} = 0.781 \times TCOD_{Influent} \qquad \textbf{Eq. 4-3}$$

This equation yields a removal of 22%, which is similar to the actual average value obtained with the experimental data (23%).

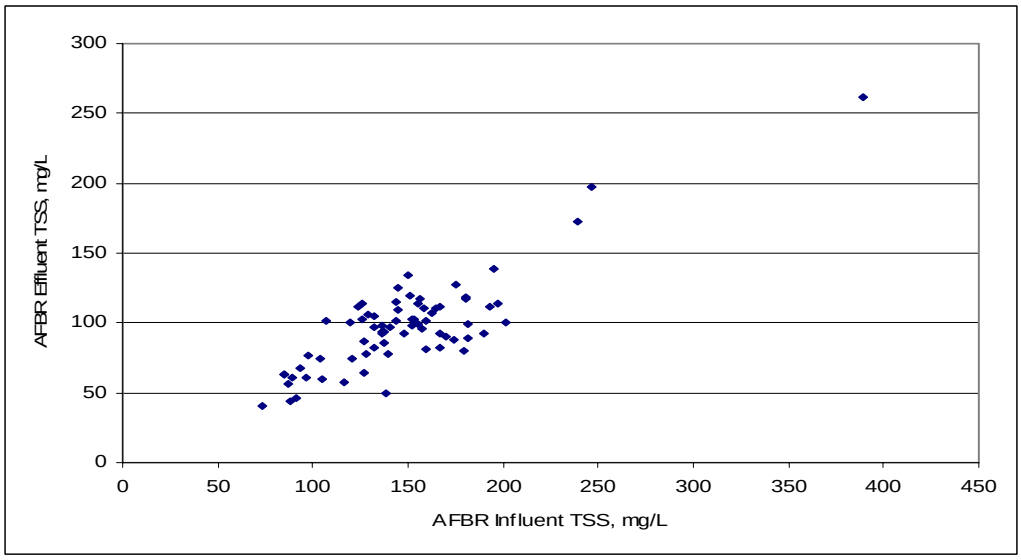

Figure 4-4 AFBR Effluent TSS vs. AFBR influent TSS

To analyze the performance of the AFBR regarding suspended solids Figure 4-4 and Figure 4-5 were prepared. Figure 4-4 shows the relationship between TSS in the influent and mixed effluent of the AFBR. Like in the case of TCOD, a glance to the graph suggests a linear relationship between the two variables. Therefore, the experimental data was fitted using a linear model. A coefficient of determination (R^2) equal to 0.72 was obtained. These results show that 72% of the variability of the data could be explained using the following equation:

$$TSS_{Effluent} = 0.656 \times TSS_{Influent} \qquad \textbf{Eq. 4-4}$$

This equation yields a removal of 34%, which is close to the average value obtained with the experimental data (31.2%).

In Figure 4-5 AFBR mixed effluent and Influent VSS concentration are shown on the vertical and horizontal axis respectively.

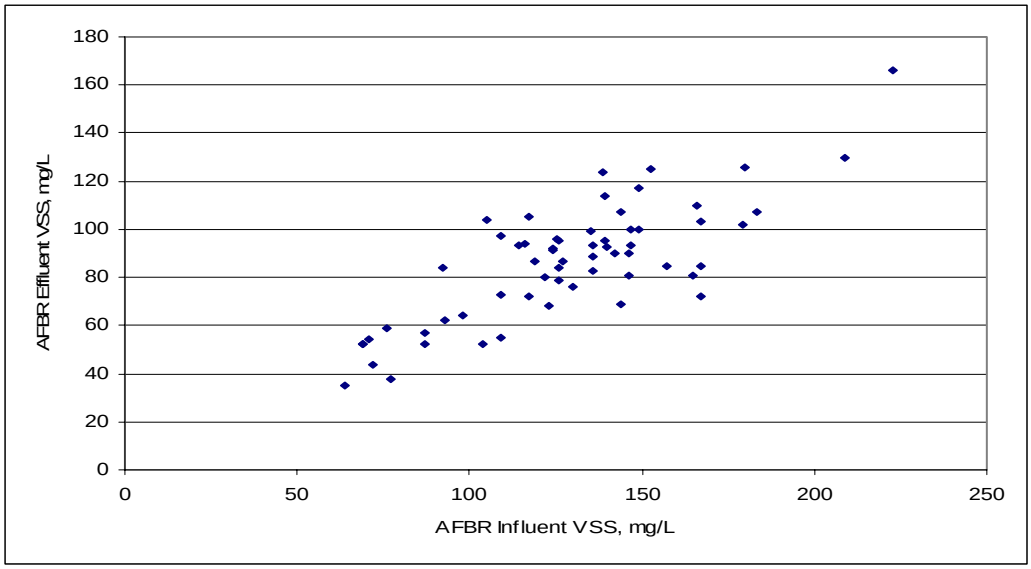

Figure 4-5 AFBR Effluent VSS vs. AFBR Influent VSS

Although the graph suggests a linear relationship, the coefficient of determination (R^2) obtained for this data was low (0.62). The line of best fit is the following:

$$TSS_{Effluent} = 0.663 \times TSS_{Influent} \qquad \textbf{Eq. 4-5}$$

This equation yields a 34% removal. This value is close to the average value obtained with the experimental data (32%)

4.2.1. Biogas Production

Yoda et al., (1985) reported that the biogas produced in an AFBR treating municipal wastewater had an average composition of 57 % methane, 40% nitrogen, 3% carbon dioxide, and traces of hydrogen sulfide. Lettinga et al., (1983) reported similar results treating municipal wastewater using a granular bed UASB reactor. The authors claimed that the nitrogen was initially dissolved in the influent and was stripped from the liquid by the methane gas produced.

In the present research, the methane content in the biogas produced ranged from 17 to 86 percent, and the carbon dioxide ranged from 3.9 to 6.5 percent. Based on the results reported by the aforementioned authors, this researcher assumed that the rest of the gas was nitrogen and traces of hydrogen sulfide. Table 4-5 shows the average concentration of methane and carbon dioxide in the biogas produced in the AFBR.

Table 4-5 Average biogas composition

Component	Methane	Carbon dioxide
Content	54 %	5.41%

As reported by Yoda et al., (1985) and Lettinga et al., (1983) the anaerobic treatment process for municipal sewage is required to operate at relatively high hydraulic loading as compared with that for high-strength wastes. Therefore, a significant part of methane leaves the reactor in a dissolved phase. According to Yoda et al., (1985), given the partial pressure of methane in the overlaying gas phase, the amount of methane dissolved in the effluent can be calculated using Henry's law.

Table 4-6 Methane production in the AFBR

Date	Observed CH_4 (ml gas/L sewage)	Dissolved CH_4 (ml gas/L sewage) Using Henry's Law	Total CH_4 (ml gas/L sewage)
9/28/2003	4.17	18.14	22
10/1/2003	2.05	20.23	22
10/3/2003	4.77	11.23	16
10/4/2003	4.78	9.88	15
10/6/2003	5.57	11.24	17
10/10/2003	5.20	11.58	17
10/12/2003	4.21	9.68	14
10/14/2003	3.80	17.86	22
10/15/2003	4.82	19.18	24
10/23/2003	5.76	14.74	20
10/24/2003	4.99	19.51	25
10/25/2003	5.58	19.46	25
10/27/2003	5.88	16.16	22
10/28/2003	4.82	10.77	16
10/30/2003	4.87	11.63	17
11/6/2003	7.36	18.52	26
11/7/2003	8.09	18.06	26

It was difficult to collect and analyze the biogas produced in the AFBR unit. Consequently, only a few points could be recorded. Table 4-6 shows the production of CH_4 observed, and an estimation of the total production according to Henry's law. Details of this calculation are presented in APPENDIX B. In Table 4-6 all volumes are reported at 25ºC and one atmosphere of pressure.

4.3. Sludge Concentration and Accumulation in the Reactor

In the AFBR the sampling points are situated at P_1=0.055 m, P_2=0.45 m, P_3=0.58 m, P_4=0.68 m, and P_5=0.92 m above the reactor bottom, as indicated in Figure 4-6.

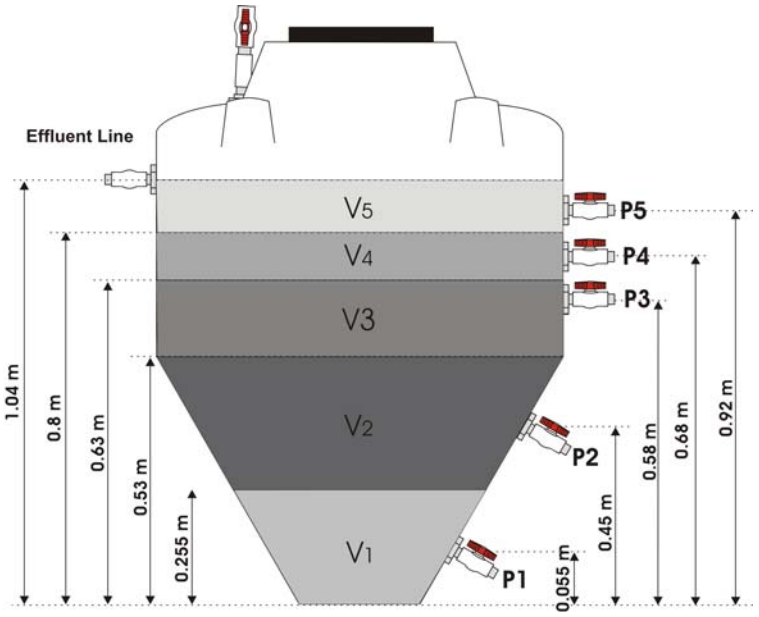

Figure 4-6 Schematic representation of the imaginary sections used to determine the sludge hold-up in the reactor

The results of the sludge concentration profile determinations are presented graphically in Figure 4-7 and Figure 4-8. As can be seen in Figure 4-7, the distribution of TSS changed between the tests. However, a common result of the three tests is that the concentrations of TSS in the first port (P_1) were very low. P_1 is within 0.05 m of the outlet of the internal recirculation system, and almost at the bottom of the conical section where the upflow velocities reach its highest value. These high upflow velocities are too high to fluidize the particles. Instead of been fluidized, the particles are

transported to higher positions, where the upflow velocities are lower because of the increase in the cross section of the reactor.

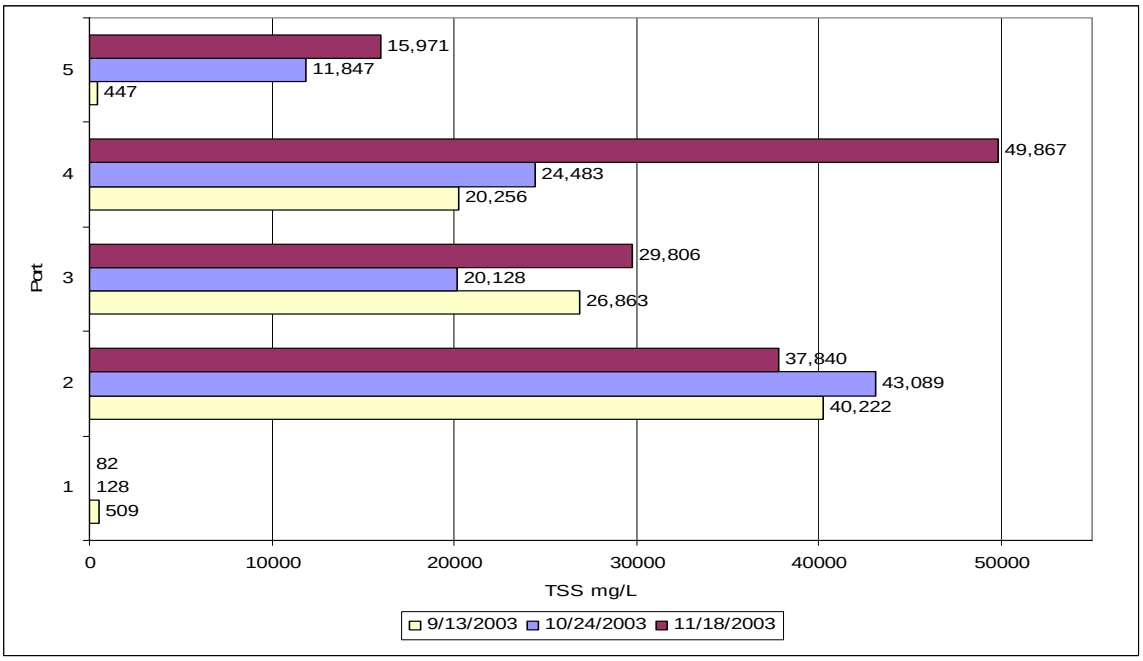

Figure 4-7 Evolution of TSS concentration

When analyzing the results of the first TSS profile test (09/13/2003), it can be noticed that the concentration of particles in the bed decreased gradually from P_2 to P_5. It can also be noticed that the concentration in P_5 was very low (445 mg/L). This results indicate that the boundary of the main sludge bed was somewhere between P_4 and P_5. The results of the second test (42 days later) show a different distribution of concentrations, showing an accumulation of solids in P_4, and a relatively high concentration in P_5 (11,847 mg/L). This results seem to indicate that the height of the bed was increasing and at that time it was somewhere between P_4 and P_5. The results of the last test (67 days after the first one) are similar to those of the second one, but

this time the highest concentration was found in P_4. Therefore, it is obvious that an accumulation of solids near P_4 took place between the first and the last test.

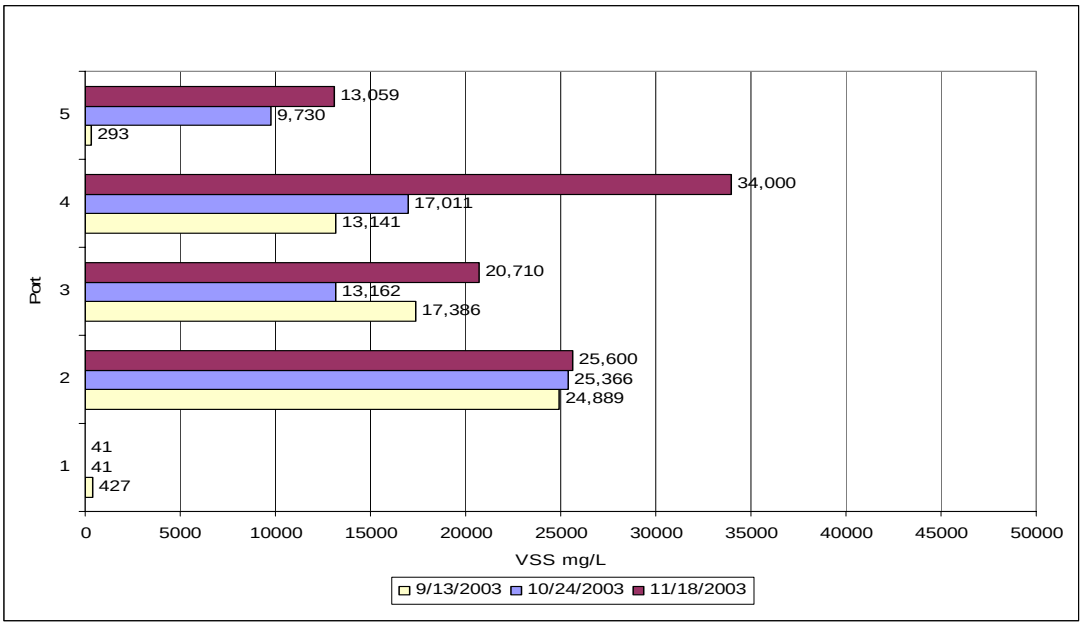

Figure 4-8 Evolution of VSS concentration

The sludge concentration profiles shown on Figure 4-7 and Figure 4-8 were used to estimate the sludge hold-up of the reactor. For this purpose the sludge profile was linearized, i.e. the reactor volume was divided in various imaginary sections. The first section V_1 was from 0 to 0.255 m above the reactor bottom (Figure 4-6), i.e. from the bottom of the reactor to the middle point between ports P_1 and P_2. The sludge concentration in V_1 was assumed to be equal to the concentration found at P_1. The concentration in the other sections was estimated similarly, e.g. the concentration found in P_2 was indicative of the concentration of the section between 0.255 m and 0.53 m

above the bottom of the reactor. Table 4-7 shows information about the imaginary sections considered for estimating the sludge hold-up.

Table 4-7 Characteristics of the imaginary sections considered to determine the reactor sludge hold-up

Section Volume	Lower limit (Meters from the bottom)	Upper limit (Meters from the bottom)	Volume, m^3
V_1	0.0	0.255	0.023
V_2	0.255	0.53	0.106
V_3	0.53	0.63	0.058
V_4	0.63	0.8	0.099
V_5	0.8	1.04	0.139

The corresponding sludge hold-up was calculated following the following equations.

$$Solids_i = V_i \times Conc._i \qquad \textbf{Eq. 4-6}$$

Where:

- $Solids_i$= mass of TSS or VSS in section V_i, g
- V_i= volume of section V_i, m^3
- $Conc._i$= concentration of TSS or VSS in section $_i$, g/ m^3

$$Total\ mass = \frac{\sum V_i \times Conc_i}{\sum V_i} \qquad \textbf{Eq. 4-7}$$

Where:

- Total mass= mass of solids (TSS or VSS, g) in the reactor

Table 4-8 shows the sludge build-up in the AFBR. The results indicate that between the first and the last solids profile tests (65 days), about 4,990 g of TSS and 4,099 g of VSS accumulated in the reactor. The results also show that the ratio VSS/TSS increased from 0.63 to 0.71, this result is logical because during the 66 days between the tests, almost all the solids accumulated in the reactor were organic solids.

Table 4-8 Sludge build-up in the AFBR

Date	TSS, g	VSS, g	Ratio TSS/VSS
09/13/2003	7,916	5,007	0.63
10/24/2003	9,827	6,501	0.66
11/18/2003	12,909	9,105	0.71

Figure 4-9 shows the solids build-up in the AFBR. These results combined with physical observation during the tests confirm that: The solids retained inside the AFBR (retained biomass) tend to accumulate in the upper portion of the reactor (region between V_4 and V_5), while the media and the biomass attached to it remains in the lower portion of the reactor (V_2 and V_3). These results are similar to the obtained by Yoda et

al., (1985), who reported that the SS introduced to their AFBR were entrapped in the upper portion of the fluidized bed to form granular pellets. Additionally, it was observed that almost no solids remain in the bottom of the reactor (V_1).

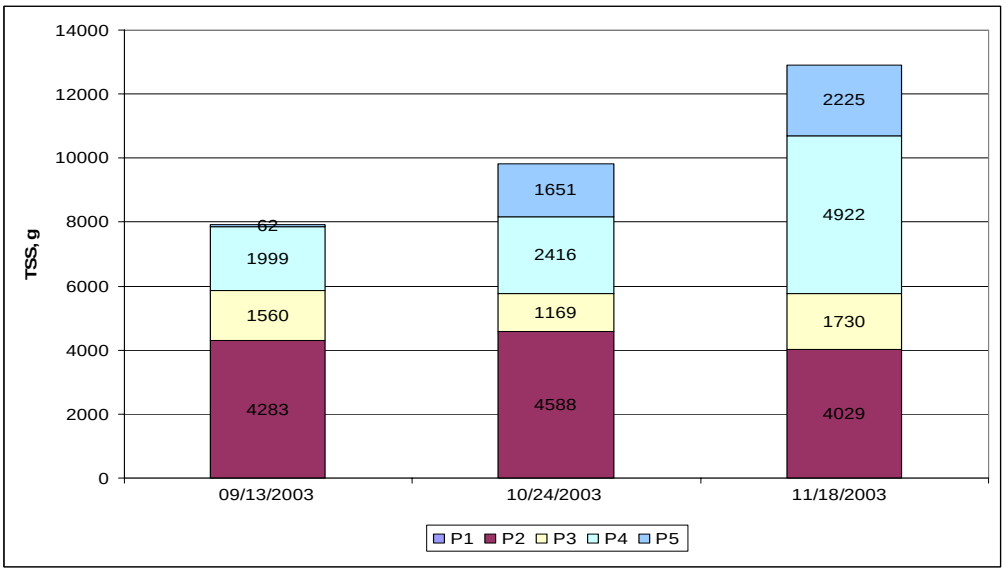

Figure 4-9 Solids build-up in the AFBR

The significant segregation of the bed is due to the difference in sizes and densities between the particles inside the reactor, i.e., the smaller particles (retained solids) are maintained fluidized at the upper portion of the reactor where the upflow velocities are lower, while the bigger particles (media) remain fluidized in the lower part of the reactor, where the upflow velocities are higher.

4.4. Mass Balance on Solids in the AFBR

To determine the amount of solids degraded inside the AFBR and establish a consumption rate it was necessary to perform a mass balance on the unit. The information available to perform the mass balance is presented in Figure 4-10.

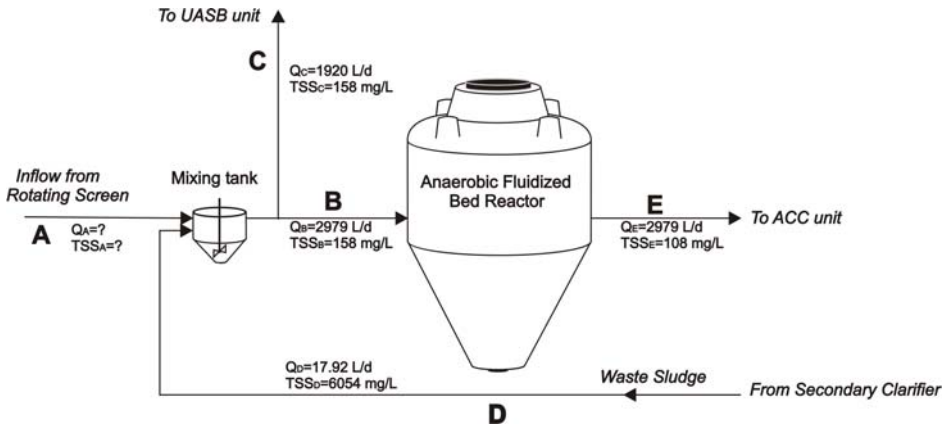

Figure 4-10 Information used to set-up the mass balance

The values presented in Figure 4-10 are the average values of the readings taken between the first and last solids profile tests (September 11 to November 18, 2003). As reported, the flow rate and TSS concentration of streams B, C, and E were directly measured by this researcher. Information about stream D was provided by Miss Jackeline Luque.

In order to know the flow rate and composition of stream A, a mass balance on the mixing tank was performed

General mass balance on the mixing tank:

$$Q_A\rho_A + Q_D\rho_D = Q_B\rho_B + Q_C\rho_C$$

Assuming constant density we have:

$$Q_A + Q_D = Q_B + Q_C$$

$$Q_A = Q_B + Q_C - Q_D$$

$$Q_A = 2{,}979\,L/d + 1{,}920\,L/d - 17.92\,L/d$$

$$\boxed{Q_A = 4{,}881.4\,L/d}$$

TSS balance on the mixing tank:

$$Q_A \times TSS_A + Q_D \times TSS_D = Q_B \times TSS_B + Q_C \times TSS_C$$

Then:
$$TSS_A = \frac{Q_B \times TSS_B + Q_C \times TSS_C - Q_D \times TSS_D}{Q_A}$$

$$TSS_A = \frac{2{,}979\,L/d \times 158\,mg/L + 1{,}920\,L/d \times 158\,mg/L - 17.92\,L/d \times 6{,}054\,mg/L}{4{,}881.4\ mg/L}$$

$$\boxed{TSS_A = 134\,mg/L}$$

TSS balance on the AFBR:

$$TSS_{feed} = TSS_{wasted} + TSS\ Accumulated + TSS\ Consumed.$$

$$TSS\ Consumed = TSS_{feed} - TSS_{wasted} - TSS\ Accumulated$$

From the solids profile tests it was determined that 4,990 g of TSS accumulated in the reactor in 66 days. Therefore, assuming a 66 days base for the mass balance we have:

$$TSS\ Consumed\ = 66\,d\left[Q_B\left(L/d\right)\times TSS_B\ - Q_E\left(L/d\right)\times TSS_E\right] - TSS\ Accumulate\,d$$

$$TSS\ Consumed\ = 66\,d * 2{,}979\left(L/d\right) * \left(158\,mg/L - 108\,mg/L\right) - 4{,}990{,}000\ \text{mg}$$

$$TSS\,Consumed = 4{,}841{,}690\,\text{mg} = 4{,}842\,\text{g}$$

Table 4-9 Mass balances results

- TSS fed to the AFBR in 66 days: 30,675 g
- TSS accumulated in the AFBR in 66 days: 4,990 g
- TSS degraded in the AFBR in 66 days: 4,842 g
- TSS recycled from the clarifier: 7,160 g
- TSS from the clarifier wasted through G (UASB unit): 2806 g
- TSS fed to the AFBR from the clarifier: 4,354 g

The results of the mass balance are presented in Table 4-9. According to these results, 16.3% of the TSS fed were removed by simple accumulation in the unit and 15.8% were degraded by the action of microorganisms. This yields an accumulation rate of 76.78 g/d and degradation rate of 74.50 g/d. Therefore, at the applied solids load of 1.09 kg/m^3.d, 0.173 kg/m^3.d are consumed, and 0.173 kg/m^3.d are accumulated in the unit.

4.4.1. Solids Digestion Test

One of the mayor applications of anaerobic digestion is the stabilization of concentrated sludges from the treatment of municipal wastewater. The degree of stabilization obtained in these systems is often measured by the reduction in volatile

solids (Metcalf and Eddy, 2003). Therefore, in order to determine the sludge stabilization attained in the AFBR, a composite sample of sludge was digested for 32 days a 37ºC and the VSS consumption was determined.

According to Metcalf and Eddy (2003), the typical volatile solids removal in an anaerobic digester treating raw sludge under similar conditions (temperature and residence time) is around 65%. The results of the solids digestion test are shown in Table 4-10. The results show that only 11% of VSS were removed during the test. Based on these we can conclude that the sludge was almost completely stabilized in the AFBR.

Table 4-10 Results obtained in the solids digestion test

Time, days	TSS, mg/L	TSS removal %	VSS, mg/L	VSS removal %
0	9600	----	5930	----
32	8947	6.8	5280	11.0

4.4.2. Nitrate Determinations

Nitrate concentrations were measured in the influent and effluent of the AFBR. These tests were done to determine whether or not denitrification was taking place inside the AFBR. The tests were done 4 times, and since no significant concentrations of nitrate were found on the influent of the AFBR, the test was not repeated and the theory of denitrification inside the unit was abandoned. These results support the hypothesis proposed by Yoda et al., (1985) and Lettinga et al., (1983). Table 4-11 shows the results of the nitrate tests.

Table 4-11 Results of the nitrate determination tests

Date	Influent NO_3^- Concentration mg/L	Effluent NO_3^- Concentration mg/L
11/14/2003	4.45	1.35
11/18/2003	4.2	0
11/21/2003	0.15	0
11/26/2003	0.35	0

5. CONCLUSIONS AND RECOMENDATIONS

The following conclusion can be drawn from this research project:

- The AFBR has a TSS removal efficiency of about 32%. Of the solids removed by the unit, 15.8% were degraded by the action of microorganisms, and the remaining 16.3% built up in the unit.
- At the applied solids load (1.09 kg SS/m^3.d) an accumulation rate of 76.78 g SS/d and degradation rate of 74.50 g SS/d was obtained in the unit. Therefore, at the applied solids load of 1.09 kg SS/m^3.d, 0.173 kg ss/m^3.d were consumed, and 0.173 kg SS/m^3.d accumulated in the bed and eventually would need to be removed.
- The AFBR stabilizes almost completely the entrapped solids, so there is no need for an additional solids digester.
- The proposed anaerobic/aerobic process is not only effective for providing efficient wastewater treatment, but also for minimizing the surplus sludge production and producing a well stabilized sludge. Therefore, it eliminates the need for an independent sludge digester thus reducing the high construction, and operation and maintenance costs associated with anaerobic sludge digestion.
- The concentration of solids in the lower half of the AFBR did not change significantly between the solids profile tests (65 days). It indicates that the

solids introduced are not entrapped in this region, and that apparently there is equilibrium between biomass growth and washout in this area.

- Because of the difference in density and size with respect to the AFBR media, organics solids and light inorganic solids introduced to the AFBR were entrapped in the upper portion of the fluidized bed. Consequently, they can be easily removed without shutting down the system and/or affecting the reactor operation.
- No significant denitrification occurs in the AFBR. Therefore, the nitrogen present in the biogas must be nitrogen that was initially dissolved in the influent wastewater and is stripped out by the methane generated during the anaerobic digestion.
- The AFBR continually produces methane gas at an average rate of 5.1 ml of CH_4 per liter of sewage treated. This could represent an additional saving, since power could be generated from the biogas produced.
- The AFBR/SC process is very efficient from the point of view of energy conservation.

Based on the experience of this research project, and the results obtained, the following items are suggested for further investigation.

- Improve the performance of the anaerobic fluidized bed reactor, (upflow velocities, bed expansion, HRT….)

- Perform a detailed mass balance on solids in the whole system (AFBR-ASCC-Clarifier.)
- Analyze the possibility of using the anaerobic unit (either the UASB or the AFBR) exclusively as a biological digestion unit. The treatment train would consist of rotating screen, aeration chamber, settling tank, and sludge digestion in the AFBR or UASB.

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7. APPENDIX A

Picture A-1 Rotating screen

Picture A-2 Distribution tank

Picture A-3 Mixing tank

Picture A-4 Anaerobic fluidized bed reactor

Picture A-5 Gas collection and measuring system

Picture A-6 Aeration chamber and secondary clarifier

8. APPENDIX B

Details of the Calculations Used to Estimate the Total Production of Methane According to Henry's Law.

From September 27,2003 at 8:00 a.m. to September 28,2003 at 8:00 a.m. (24 hours), the volume of biogas produced was 22 L, measured at 28 °C. The biogas had a methane concentration of 57 % V/V. During the 24 hour period the reactor was feed at a constant flow rate of 124.14 L/h

First the volume of gas (28°C) was converted to 25°C, following the ideal gas law.

$$P \times V = n \times R \times T \qquad \Rightarrow \qquad n = \frac{P \times V}{R \times T}$$

$$n_1 = n_2 \qquad \Rightarrow \qquad \frac{P_1 \times V_1}{R \times T_1} = \frac{P_2 \times V_2}{R \times T_2} \qquad \Rightarrow \qquad V_2 = \frac{V_1 \times T_2}{T_1}$$

$$V_2 = \frac{V_1 \times T_1}{T_2} = \frac{22\,L \times (25 + 273.15)K}{(28 + 273.15)K} = 21.78\,L$$

Therefore, the total volume of methane produced is:

$$Vol_{CH_4} = Vol_{gas} \times Concentration_{CH_4}$$

$$Vol_{CH_4} = 21.78\,L \times 0.57 \qquad \Rightarrow \qquad \boxed{Vol_{CH_4} = 12.41L = 12{,}410\,ml}$$

The volume of sewage feed to the reactor during the 24 h period is

$$Vol_{sewage} = Q_{sewage} \times t = 124.14\,L/h \times 24\,h = 2979.36\ \text{L}$$

Therefore, the volume of methane produced per volume of sewage feed to the reactor is: $\frac{12{,}410\,ml}{2{,}979.36\,L} = 4.17\,ml\,CH_4/L\,sewage$

Hence, the observed production of methane is: $\boxed{4.17\,ml\,CH_4/L\,sewage}$

To estimate the amount of methane gas dissolved in the effluent of the AFBR, it was assumed that the liquid was saturated with CH_4. The corresponding methane concentration was calculated using Henry's law.

Metcalf and Eddy (2003), proposes the following equation to estimate the Henry's constant.

$$Log_{10}(H) = \left(\frac{-A}{T} + B\right)$$

Where:

H= Henry's constant at temperature T, K.

A=empirical constant that takes into account the enthalpy change in water due to the dissolution of a component in water and the universal gas law constant. For methane A=675.74.

T=temperature, K=°C+273.15

B=empirical constant. For methane B=6.88

Rearranging the previous equation we obtain:

$$H = 10^{\left(\frac{-A}{T}+B\right)} \quad \Rightarrow \quad H = 10^{\left(\frac{-675.74}{T}+6.88\right)}$$

Therefore the Henry's constant for methane at 28°C is

$$H = 10^{\left(\frac{-675.74}{(28+273.15)}+6.88\right)} = 43264.8 atm$$

The relationship between the mole fraction of the methane in the gas above the liquid and the mole fraction of the methane in the liquid is given by the following form of Henry's Law

$$Y_{CH_4} = H \times X_{CH_4} = \frac{V_{CH_4}}{V_T}$$

Where:

Y_{CH_4} =mole fraction of methane in the gas phase

V_{CH_4} =volume of methane in the gas mixture

V_{T_4} =total volume of the gas mixture

X_{CH_4} = mole fraction of methane in the liquid phase

H = Henry's constant in atm

According to the previous equation, the molar fraction of methane in the liquid phase is:

$$X_{CH_4} = \frac{\frac{V_{CH_4}}{V_T}}{H} = \frac{0.57}{43{,}264.8} = 13 \times 10^{-6}$$

The molar fraction of methane in the in liquid phase is defined as:

$$X_{CH_4} = \frac{n_{CH_4}}{n_{CH_4} + n_{H_2O}}$$

One liter of water contains $1000\text{g} = 55.6 \text{ mole}$ 1000g=55.6 mole. Therefore, the number of moles of dissolved gas in a liter of water is much less than the number of moles of water.

$$n_{CH_4} = X_{CH_4} \times n_{H_2O} = 13 \times 10^{-6} \times 55.6 = 7.2 \times 10^{-4}\ mole CH_4/L$$

The volume that would be occupied by the moles of methane dissolved in the liquid phase can be calculated by the ideal gas law.

$$V = \frac{n \times R \times T}{P} = \frac{7.23 \times 10^{-4} \times 0.082\ atm.L/mole.K \times (25 + 273.15)K}{1 atm} \times 1000\ ml/L$$

Therefore, the volume (25 °C, 1 atm) of methane dissolved is:

$$\boxed{V = 18 ml\ CH_4/L}$$

Finally, the total volume of methane produced per liter of sewage is

$$V_{Total} = V_{observed} + V_{dissolved} = 4.17 ml CH_4/L + 18 ml\ CH_4/L = 22 ml\ CH_4/L$$

$\boxed{V_{Total} = CH_4/L\ sewage}$ at 25 °C and 1 atm

9. APPENDIX C

Table C-1 AFBR Influent and Effluent TCOD, mg/L

	TCOD (mg/L)		
Date	**Influent**	**Mixed Effluent**	**Settled Effluent**
1/14/2003	336	249	N/A
1/15/2003	346	278	N/A
1/16/2003	304	240	N/A
1/31/2003	358	291	N/A
2/7/2003	355	285	N/A
2/8/2003	323	186	N/A
2/10/2003	349	291	N/A
2/12/2003	340	285	N/A
2/14/2003	406	288	N/A
2/15/2003	403	313	N/A
3/14/2003	275	224	N/A
3/16/2003	272	182	N/A
5/17/2003	330	272	N/A
5/20/2003	378	362	N/A
5/21/2003	318	317	N/A
5/23/2003	375	279	N/A
5/24/2003	380	303	N/A
5/26/2003	394	294	N/A
5/29/2203	335	284	N/A
5/31/2003	352	294	N/A
6/5/2203	269	201	N/A
6/6/2003	256	205	N/A
6/9/2003	390	250	N/A
6/10/2003	416	256	N/A
6/11/2003	213	245	N/A
6/13/2003	210	144	N/A
6/14/2003	237	181	N/A
6/17/2003	238	195	N/A
6/18/2003	234	227	N/A
6/20/2003	230	173	N/A
6/21/2003	211	173	N/A
6/22/2003	250	235	N/A
6/25/2003	221	184	122
6/26/2003	269	160	109
6/27/2003	147	90	90
6/28/2003	199	98	60
6/29/2003	155	98	53
7/3/2003	171	145	104

Table C-1 (Continued)

7/5/2003	319	259	179
7/6/2003	313	195	136
7/7/2003	235	280	36.7
7/8/2003	174	122	87
7/12/2003	90	57	41
7/16/2003	216	129	38
7/19/2003	93	45	10
7/24/2003	244	141	114
7/25/2003	237	213	131
7/27/2003	217	154	107
7/28/2003	236	195	141
7/29/2003	227	242	116
7/30/2003	242	161	128
8/2/2003	265	191	157
8/13/2003	284	217	153
8/15/2003	389	262	170
8/18/2003	378	271	192
8/21/2003	307	215	141
8/23/2003	265	217	176
9/5/2003	296	207	135
9/10/2003	479	358	167
9/13/2003	471	308	166
9/28/2003	293	210	183
10/1/2003	317	256	204
10/3/2003	348	269	217
10/4/2003	325	277	228
10/6/2003	383	303	267
10/10/2003	1073	286	212
10/12/2003	189	194	158
10/14/2003	286	221	166
10/15/2003	280	228	182
10/23/2003	324	295	208
10/24/2003	372	307	219
10/25/2003	379	311	240
10/27/2003	338	291	244
10/28/2003	368	289	235
10/30/2003	319	286	230
11/6/2003	348	315	232
11/7/2003	327	291	232
11/14/2003	374	303	251
11/18/2003	406	337	274
11/21/2003	374	310	238
11/22/2003	361	317	238
11/26/2003	413	324	275
12/5/2003	334	274	218

Table C-2 AFBR Influent and Effluent TSS, mg/L

TSS (mg/L)			
Date	**Influent**	**Mixed Effluent**	**Settled Effluent**
1/14/2003	160	82	N/A
1/15/2003	175	88	N/A
1/16/2003	121	75	N/A
1/31/2003	145	125	N/A
2/7/2003	163	107	N/A
2/8/2003	138	94	N/A
2/10/2003	138	50	N/A
2/12/2003	156	114	N/A
2/14/2003	201	101	N/A
2/15/2003	190	92	N/A
5/17/2003	124	112	N/A
5/20/2003	150	134	N/A
5/21/2003	176	127	N/A
5/23/2003	157	96	N/A
5/24/2003	170	90	N/A
5/26/2003	193	112	N/A
5/29/2203	197	114	N/A
5/31/2003	180	117	N/A
6/5/2203	159	101	N/A
6/6/2003	136	93	N/A
6/9/2003	181	89	N/A
6/10/2003	179	80	N/A
6/11/2003	145	112	N/A
6/13/2003	91	46	N/A
6/14/2003	132	82	N/A
6/17/2003	136	94	N/A
6/18/2003	120	100	N/A
6/20/2003	128	78	N/A
6/21/2003	116	57	N/A
6/22/2003	144	115	N/A
6/25/2003	127	87	43
6/26/2003	167	82	27
6/27/2003	73	41	30
6/28/2003	89	61	27
6/29/2003	85	63	29
7/3/2003	93	68	32
7/5/2003	85	63	29
7/6/2003	167	93	51

Table C-2 (Continued)

7/7/2003	122	162	71
7/8/2003	98	77	36
7/12/2003	97	61	50
7/16/2003	129	106	40
7/19/2003	87	56	27
7/24/2003	127	64	40
7/25/2003	126	114	45
7/27/2003	104	75	40
7/28/2003	107	102	46
7/29/2003	95	68	52
7/30/2003	105	60	35
8/2/2003	120	129	40
8/4/2003	103	53	32
8/6/2003	88	44	27
8/8/2003	92	62	43
8/13/2003	389	262	170
8/18/2003	88	44	27
8/23/2003	137	86	46
9/5/2003	132	97	48
9/10/2003	247	197	58
9/13/2003	239	172	61
9/28/2003	139	78	57
10/1/2003	141	97	61
10/3/2003	152	98	62
10/4/2003	145	109	68
10/6/2003	165	99	76
10/10/2003	1340	122	63
10/12/2003	97	72	42
10/14/2003	181	99	51
10/15/2003	152	103	62
10/23/2003	136	96	47
10/24/2003	167	112	54
10/25/2003	180	118	162
10/27/2003	144	101	70
10/28/2003	148	92	58
10/30/2003	132	105	64
11/6/2003	151	119	64
11/7/2003	126	103	64
11/14/2003	165	110	73
11/18/2003	195	139	80
11/20/2003	165	137	72
11/21/2003	153	103	94
11/22/2003	158	111	52
11/26/2003	156	117	70
12/5/2003	142	196	68

Table C-3 AFBR Influent and Effluent VSS, mg/L

VSS (mg/L)			
Date	**Influent**	**Mixed Effluent**	**Settled Effluent**
5/17/2003	105	104	N/A
5/20/2003	138.5	123.5	N/A
5/21/2003	152.5	125	N/A
5/23/2003	140	92.5	N/A
5/24/2003	157	84.5	N/A
5/26/2003	183.5	107	N/A
5/29/2203	179	102	N/A
5/31/2003	167	103	N/A
6/5/2203	146	90	N/A
6/6/2003	122	80	N/A
6/9/2003	167	85	N/A
6/10/2003	167	72	N/A
6/11/2003	129	89	N/A
6/14/2003	123	68	N/A
6/17/2003	126	84	N/A
6/18/2003	114	93	N/A
6/20/2003	117	72	N/A
6/21/2003	104	52	N/A
6/22/2003	126	95	N/A
6/25/2003	109	73	39
6/26/2003	144	69	27
6/27/2003	64	35	25
6/28/2003	71	54	21
6/29/2003	69	52	24
7/3/2003	76	59	30
7/5/2003	69	52	24
7/6/2003	146	81	42
7/7/2003	106	133	69
7/8/2003	87	57	35
7/12/2003	87	52	43
7/16/2003	124	92	38
7/19/2003	72	44	22
7/24/2003	109	55	37
7/25/2003	109	97	37
7/27/2003	93	62	33
7/28/2003	92	84	40
7/29/2003	90	62	51
7/30/2003	98	64	35
8/2/2003	117	105	37
8/4/2003	90	48	28
8/6/2003	77	38	25

Table C-3 (Continued)

8/8/2003	85	56	36
8/18/2003	77	38	25
8/23/2003	126	79	46
9/5/2003	119	87	47
9/10/2003	223	166	48
9/13/2003	209	130	53
9/28/2003	130	86	57
10/1/2003	136	93	59
10/3/2003	142	99	58
10/4/2003	135	95	65
10/6/2003	147	93	71
10/10/2003	1218	111	60
10/12/2003	87	67	42
10/14/2003	165	81	49
10/15/2003	124	91	55
10/23/2003	127	87	38
10/24/2003	149	100	47
10/25/2003	166	110	60
10/27/2003	136	89	68
10/28/2003	136	83	57
10/30/2003	125	96	67
11/6/2003	139	114	62
11/7/2003	116	94	61
11/14/2003	147	100	66
11/18/2003	180	126	73
11/20/2003	150	124	65
11/21/2003	139	95	88
11/22/2003	144	107	51
11/26/2003	149	117	69
12/5/2003	134	181	63

Table C-4 Methane concentration in the biogas produced in the AFBR

Date	% CH_4
5/17/2003	27%
5/22/2003	19%
5/23/2003	17%
5/24/2003	23%
5/25/2003	43%
5/26/2003	43%
5/27/2003	50%
5/28/2003	52%
5/29/2003	57%
6/5/2003	50%
6/11/2003	75%
6/12/2003	60%
6/13/2003	78%
6/14/2003	79%
6/16/2003	83%
6/18/2003	86%
6/19/2003	79%
6/20/2003	73%
6/22/2003	84%
6/24/2003	81%
6/26/2003	74%
6/27/2003	74%
6/28/2003	74%
6/29/2003	71%
9/28/2003	57%
10/1/2003	60%
10/3/2003	32%
10/4/2003	29%
10/6/2003	34%
10/10/2003	34%
10/12/2003	29%
10/14/2003	54%
10/15/2003	58%
10/23/2003	45%
10/24/2003	59%
10/25/2003	60%
10/27/2003	43%
10/28/2003	30%
10/30/2003	35%
11/6/2003	56%
11/7/2003	52%

Table C-5 Volume of biogas produced and its corresponding temperature

Date	Biogas Volume , L	temperature, °C
9/28/2003	22	28
10/1/2003	20	24
10/3/2003	50	22
10/4/2003	38	23
10/6/2003	44	25
10/10/2003	38	23
10/12/2003	40	24
10/14/2003	20	25
10/15/2003	24	25
10/23/2003	37.5	26
10/24/2003	26	25
10/25/2003	22	26
10/27/2003	40	17
10/28/2003	47	19
10/30/2003	40	24
11/6/2003	42	25
11/7/2003	46	22

Table C-6 AFBR Influent and Effluent NO_3^- concentration, mg/L

Date	Influent NO_3^- Concentration, mg/L	Efluent NO_3^- Concentration, mg/L
11/14/2003	4.45	1.35
11/18/2003	4.2	0
11/21/2003	0.15	0
11/26/2003	0.35	0

Table C-7 Solids profile test 09/13/2003

Port	Dilution factor	Sample volume (ml)	Weight 1 (g)	Weight 2 (g)	Weight 3 (g)	TSS (mg/L)	VSS (mg/L)	Average	Average
1	1	81	1.112	1.1526	1.118	501	427	**TSS (mg/L)**	**VSS (mg/L)**
	1	81	1.1122	1.1524	1.117	496	437	509	427
	1	81	1.109	1.152	1.1125	531	488		
2	0.045	10	1.1135	1.1302	1.12	37111	22667	**TSS (mg/L)**	**VSS (mg/L)**
	0.045	10	1.118	1.1376	1.1254	43556	27111		
	0.045	10	1.114	1.132	1.1208	40000	24889	40222	24889
3	0.051	10	1.1247	1.1384	1.1294	26863	17647	**TSS (mg/L)**	**VSS (mg/L)**
	0.051	10	1.1198	1.1335	1.1247	26863	17255	26863	17386
	0.051	10	1.1134	1.1272	1.1184	27059	17255		
4	0.052	10	1.1028	1.1132	1.1066	20000	12692	**TSS (mg/L)**	**VSS (mg/L)**
	0.052	10	1.1233	1.134	1.1269	20577	13654	20256	13141
	0.052	10	1.102	1.1125	1.1057	20192	13077		
5	1	10	1.1149	1.1193	1.1164	440	290	**TSS (mg/L)**	**VSS (mg/L)**
	1	10	1.1171	1.1216	1.1186	450	300	447	293
	1	10	1.1235	1.128	1.1251	450	290		
Effluent	1	50	1.5105	1.5188	1.5135	166	106	**TSS (mg/L)**	**VSS (mg/L)**
	1	50	1.1102	1.1191	1.1114	178	154	172	130
	1	50	1.1149	1.1235	1.117	172	130		

Table C-8 Solids profile test 10/24/2003

Port	Dilution factor	Sample volume (ml)	Weight 1 (g)	Weight 2 (g)	Weight 3 (g)	TSS (mg/L)	VSS (mg/L)	Average	Average
1	1	30	1.1164	1.1201	1.1189	123	40	**TSS (mg/L)**	**VSS (mg/L)**
	1	30	1.1119	1.1159	1.1145	133	47	128	41
	1	30	1.0969	1.1007	1.0996	127	37		
2	0.041	10	1.0923	1.1097	1.0995	42439	24878	**TSS (mg/L)**	**VSS (mg/L)**
	0.041	10	1.0913	1.1101	1.0991	45854	26829		
	0.041	10	1.1018	1.1186	1.1086	40976	24390	43089	25366
3	0.039	10	1.0997	1.1075	1.1024	20000	13077	**TSS (mg/L)**	**VSS (mg/L)**
	0.039	10	1.0984	1.103	1.0979	11795	13077	20128	13162
	0.039	10	1.0977	1.1056	1.1004	20256	13333		
4	0.029	10	1.1007	1.1078	1.1029	24483	16897	**TSS (mg/L)**	**VSS (mg/L)**
	0.029	10	1.0998	1.1066	1.1021	23448	15517	24483	17011
	0.029	10	1.102	1.1094	1.104	25517	18621		
5	0.074	10	1.096	1.1051	1.098	12297	9595	**TSS (mg/L)**	**VSS (mg/L)**
	0.074	10	1.0969	1.1051	1.0981	11081	9459	11847	9730
	0.074	10	1.1024	1.1114	1.1039	12162	10135		
Effluent	1	50	1.0948	1.1031	1.0958	166	146	**TSS (mg/L)**	**VSS (mg/L)**
	1	50	1.0934	1.1018	1.0942	168	152	167	149
	1	50	1.102	1.1103	1.1029	166	148		

Table C-9 Solids profile test 11/18/2004

Port	Dilution factor	Sample volume (ml)	Weight 1 (g)	Weight 2 (g)	Weight 3 (g)	TSS (mg/L)	VSS (mg/L)	Average	Average
1	0.427	40	1.1162	1.1171	1.1166	53	29	**TSS (mg/L)**	**VSS (mg/L)**
	0.427	40	1.1127	1.1141	1.1134	82	41	82	41
	0.427	40	1.104	1.1054	1.1047	82	41		
2	0.0417	10	1.1218	1.1376	1.1265	37920	26640	**TSS (mg/L)**	**VSS (mg/L)**
	0.0417	10	1.125	1.141	1.1305	38400	25200		
	0.0417	10	1.1181	1.1336	1.1232	37200	24960	37840	25600
3	0.0517	10	1.1255	1.1408	1.1302	29613	20516	**TSS (mg/L)**	**VSS (mg/L)**
	0.0517	10	1.1158	1.1313	1.1205	30000	20903	29806	20710
	0.0517	10	1.5132	1.5327	1.5184	37742	27677		
4	0.05	10	1.1133	1.1365	1.1188	46400	35400	**TSS (mg/L)**	**VSS (mg/L)**
	0.05	10	1.1308	1.1529	1.1366	44200	32600	49867	34000
	0.05	10	1.1235	1.153	1.1293	59000	47400		
5	0.057	10	1.1135	1.1205	1.1149	12353	9882	**TSS (mg/L)**	**VSS (mg/L)**
	0.057	10	1.1077	1.1165	1.1092	15529	12882	15971	13059
	0.057	10	1.1251	1.1344	1.1269	16412	13235		
Effluent	1	50	1.1029	1.1095	1.1035	132	120	**TSS (mg/L)**	**VSS (mg/L)**
	1	50	1.111	1.1183	1.1117	146	132	139	126
	1	50	1.108	1.115	1.1087	140	126		

10. APPENDIX D

Table D-1 Statistical Analysis

AFBR Influent TCOD vs. Influent TSS

Regression Statistics	
Multiple R	0.850451927
R Square	0.723268479
Adjusted R Square	0.708562597
Standard Error	43.75615469
Observations	69

Analysis of Variance						
	df	*SS*	*MS*	*F*	*Significance F*	
Regression	1	340273.4934	340273.5	177.7255	1.61122E-20	
Residual	68	130192.873	1914.601			
Total	69	470466.3664				
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*
Intercept	0	#N/A	#N/A	#N/A	#N/A	#N/A
X1	2.084775541	0.035171051	59.27533	2.85E-60	2.014592844	2.154958239

Table D-2 Statistical Analysis

AFBR Influent TCOD vs. Influent VSS

Regression Statistics	
Multiple R	0.863121346
R Square	0.744978458
Adjusted R Square	0.727121315
Standard Error	43.01951109
Observations	57

Analysis of Variance						
	df	*SS*	*MS*	*F*	*Significance F*	
Regression	1	302751.1599	302751.1599	163.5892928	4.07732E-18	
Residual	56	103637.9867	1850.678334			
Total	57	406389.1467				
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*
Intercept	0	#N/A	#N/A	#N/A	#N/A	#N/A
X1	2.273782569	0.041949063	54.20341677	4.40592E-50	2.189748553	2.357816585

Table D-3 Statistical Analysis

AFBR Influent TCOD vs. Effluent TCOD

Regression Statistics	
Multiple R	0.915293098
R Square	0.837761455
Adjusted R Square	0.823872566
Standard Error	29.32246895
Observations	73

Analysis of Variance						
	df	*SS*	*MS*	*F*	*Significance F*	
Regression	1	319668.5402	319668.5	371.7909616	6.18327E-30	
Residual	72	61906.11734	859.8072			
Total	73	381574.6575				
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*
Intercept	0	#N/A	#N/A	#N/A	#N/A	#N/A
X1	0.780863603	0.010887994	71.71785	1.03792E-68	0.759158797	0.80256841

Table D-4 Statistical Analysis

AFBR Influent TSS vs. Effluent TSS

Regression Statistics	
Multiple R	0.848634816
R Square	0.720181051
Adjusted R Square	0.706096544
Standard Error	17.54379056
Observations	72

Analysis of Variance						
	df	*SS*	*MS*	*F*	*Significance F*	
Regression	1	56243.16931	56243.16931	182.7354963	3.39698E-21	
Residual	71	21852.70569	307.7845872			
Total	72	78095.875				
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*
Intercept	0	#N/A	#N/A	#N/A	#N/A	#N/A
X1	0.656065693	0.013400157	48.95955284	1.87123E-56	0.629346526	0.68278486

Table D-5 Statistical Analysis

AFBR Influent VSS vs. Effluent VSS

Regression Statistics	
Multiple R	0.786796479
R Square	0.619048699
Adjusted R Square	0.602382032
Standard Error	15.10412567
Observations	61

Analysis of Variance						
	df	*SS*	*MS*	*F*	*Significance F*	
Regression	1	22243.22654	22243.22654	97.50044638	4.12233E-14	
Residual	60	13688.07674	228.1346123			
Total	61	35931.30328				
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*
Intercept	0	#N/A	#N/A	#N/A	#N/A	#N/A
X1	0.663557087	0.014528392	45.67312673	2.58492E-48	0.634495986	0.692618189

VITA

Harold Jose Padron Bozo was born in Cabimas, Venezuela, on August 18, 1975. In 1992, He graduated from ISB High School in Ciudad Ojeda. Later on, he graduated from Universidad del Zulia in March 2000, obtaining a degree of Bachelor of Sciences in Chemical Engineering. In fall 2002, he started at the University of New Orleans, pursuing a Master's of Science in Environmental Engineering.

During graduate school, he was a Graduate Assistant for two years at the Urban Waste Management and Research Center in the Department of Civil and Environmental Engineering. His academic emphasis is focused in the areas of water and wastewater treatment processes in the environmental field.