EVALUATING BIOLOGICAL TREATMENT SYSTEMS

I. MOVING BED BIOFILM REACTOR VERSUS BIOLOGICAL AERATED FILTRATION

AND

II. SULFIDE-INDUCED CORROSION IN ANAEROBIC DIGESTER GAS PIPING.

KOFI ASIEDU

A thesis submitted to the facualty of Virginia Polytechnic Institute and State University in partial fulfillment of the requirements for the degree of

Master of Science
in
Environmental Engineering

Department of Civil and Environmental Engineering

Blacksburg, Virginia
September, 2001

TABLE OF CONTENT – SECTION I

TABLE OF FIGURES – SECTION I

LIST OF FIGURES – SECTION II

LIST OF TABLES – SECTION I

LIST OF TABLES – SECTION II

ABSTRACT

The research presented in this report is in two sections. Section I involved the performance of a moving bed biofilm reactor (MBBR) versus a biological aerated filtration (BAF) and Section II involved study on causes of deposition in anaerobic digester gas piping.

The first section evaluated and compared the performance of a laboratory-scale MBBR and BAF for organic carbon and suspended solids removal. A kinetic study was also performed on the MBBR to evaluate the system performance. The purpose was to recommend one of the systems for the Force Provider project, which provides a containerized "city" for the U.S. Army. The effluent criteria against which the systems were evaluated were total 5-day biochemical oxygen demand ($TBOD_5$) and total suspended solids (TSS) of 30 mg/L each. The report is based on a 5-month laboratory -scale study of the two reactors.

The MBBR performance depended on the percent of media provided in the reactor and the organic loading. At a media volume, which displaced the reactor volume by 40 % (heretofore called 40 % media volume), and surface area loading rate (SALR) of 20 g BOD_5/m^2-d, the system performance deteriorated with time. At 40 % media volume and SALR below 15 g BOD_5/m^2-d, the system performance improved but still did not meet effluent criteria or average. $TBOD_5$ reduction was generally poor (approximately 50 %). Soluble BOD_5 ($SBOD_5$) concentrations were frequently below 30 mg/L and TSS concentrations were often higher than influent TSS. Overall, TSS wastage from the system (both effluent TSS and intentional wastage) averaged 0.032 kg/d.

BAF system performance was excellent for $TBOD_5$, $CBOD_5$, $SBOD_5$ and TSS removal, and were consistently less that 30 mg/L. Overall TSS wastage from the BAF (both via effluent and backwash) average 0.027 kg/d and was 16 % less than for the MBBR. Based on demonstrated performance, the BAF was the only viable reactor for the project.

Section II of the report focused on possible causes of deposition in an anaerobic digester gas piping at a local wastewater treatment facility (Peppers ferry regional wastewater treatment facility).

Industrial waste input to the treatment facility has increased lately and accounts for 40 % of the plant's wastewater inflow. An industry in Pulaski, VA, Magnox Inc. generates and disposes highly concentrated sodium sulfate, (70,000 mg/L) which is a by-product of its activities, to PFRWTF wastewater influent stream. As a result of Magnox industrial waste input, a pilot study was carried out to determine the effect of its waste on the activated sludge treatment units. Results indicated that Magnox industrial waste input would not have adverse effect on the aeration basins. However production of H_2S, which can have effect on the anaerobic digester was reported (Olver Inc., 1995). Field analysis of data reported by Olver Inc. (2000) showed that H_2S concentration in PFRWTF anaerobic digester gas was rising. X-ray photoelectron spectroscopy analysis of deposits found in the digester pipe together with results obtained from the laboratory-scale study revealed that iron and sulfur played a role in the deposition in the digester gas pipe. The laboratory scale study revealed that ferrous ion in the digester feed possibly precipitated over 90 % of the hydrogen sulfide gas produced in the digester, thus protecting the digester from adverse effects caused by H_2S.

SECTION I MOVING BED BIOFILM REACTOR VERSUS BIOLOGICAL AERATED FILTRATION

CHAPTER 1.0 INTRODUCTION AND BACKGROUND

The need to provide an efficient wastewater collection and treatment facility within the constraints of Force Provider, a containerized highly deployable "city" for the U.S. Army, formed the basis of this research. This thesis reports on research conducted to determine the suitability of two wastewater treatment systems to be used as a subsystem in a proposed Force Provider Wastewater Treatment System (FPWTS). The FPWTS consists of a collection system (equalization tank), a waste filtration unit (WFU) using a geotextile material to treat raw and possibly secondary wastes, a biosystem to treat the filtrate, and an incinerator to manage disposal of the filtered solids and geotextile filter. The FPWTS is intended to produce a high quality liquid effluent and efficient solid waste management to meet standards both at home in the United States and other countries.

One of the requirements of Force Provider is to provide facilities capable of being transported from one point to another in 8 x 8 x 20 ft ISO containers during deployment of the army in operations both at home and outside. Therefore, size and weight were among those considerations limiting the selection of facilities to be provided. The maximum weight capable of being lifted is 10,000 lb. The elevation dimension of any facility cannot be bigger than 8 x 8 ft. For this reason, it was necessary to use high performance treatment systems, which have small footprints, and are capable of being moved to new locations with rapid process startup periods. (UTD Inc., 2000)

Four " off-the-shelf " technologies were recommended for consideration during the experimental phase. These technologies and their potential vendors were:

- Moving Bed Biofilm Reactor (MBBR), Waterlink, Fall River, MA
- Biological Aerated Filter (BAF), Infilco Degremont, Inc, Richmond, VA or Kruger Inc., Cary, North Carolina.
- Continuously Fed Sequencing Batch Reactor (CFSBR), Eco process, Inc., Terrebonne, Quebec, Canada.
- Membrane Bioreactor (MBR), Zenon Environmental, Inc., Oakville, Ontario Canada.

Each of the above systems was evaluated based on each system's ability to treat the highly concentrated wastewater, efficiency of the solid/liquid separation step, and operational and maintenance simplicity (Novak et al., 2000). Each of the systems above had noted advantages and disadvantages relative to the goals of the project. During a meeting that occurred between Virginia Tech and UTD, Inc, it was decided that the MBBR and BAF were the biological systems most feasible for testing during the experimental study. The advantages and disadvantages of each of these systems will be discussed later in the thesis.

Laboratory-scale MBBR and BAF systems were designed and constructed based on anticipated filtered wastewater characteristics in Force Provider (UTD Inc., 2000). The systems were operated and monitored over 5 months for organic matter and suspended solids removal. Other wastewater characteristics monitored to evaluate performance of the reactors were ammonia-N concentration, temperature, pH and dissolved oxygen. Wastewater from a sewer on the Virginia Tech campus was used as influent for the study. The wastewater was characterized and modified to meet influent requirements expected from the FPWFU.

Composite samples of influent wastewater and effluent from the reactors were analyzed for total, carbonaceous and soluble 5-day biochemical oxygen demand ($TBOD_5$, $CBOD_5$ and $SBOD_5$, respectively), total suspended solids (TSS) and ammonia-N. Composite samples were collected at 3 hr intervals over 24 hrs and kept in a constant 4 degree Celsius refrigerator during the collection period.

CHAPTER 2.0 OBJECTIVES

The objectives of the research were to:

- Monitor performance of the two biosystems and evaluate results.
- Select one system for use in the FPWTS's biological treatment subsystem;
- Use the results obtained from the laboratory study in the design and construction of a pilot scale plant of the selected system.

Secondary objectives included estimating kinetic parameters associated with biodegradation in the MBBR to explain performance of the system. The pseudo analytical model for analyzing attached growth systems was used with both measured and typical kinetic and stoichiometric parameters to determine substrate flux into the biofilm, the media surface area and the thickness of biofilm. Results from this theoretical analysis were compared with actual reactor performance.

CHAPTER 3.0 LITERATURE REVIEW

The use of MBBR and BAF has been reported in both pilot plant studies and full-scale plants (Gilmore et al., 1999; Rusten et al., 1997; Sunner et al.,1999). They are usually employed in carbonaceous BOD removal and nitrification, but nutrient removal is possible with different system configurations and conditions. The review of these systems will concentrate on organic matter and suspended solids removal. Although nitrification was not an objective of this research, both systems nitrified to some extent. Therefore, nitrification would be discussed in the Results and Discussion section of this report.

The two systems use aerobic attached biofilm growth processes for wastewater treatment. The MBBR uses media suspended in liquid and kept in constant circulation by the aeration system of the reactor. If anoxic conditions are desired, mixing is provided by mechanical means. The upflow BAF media operates in a packed (without movement) mode and is submerged in water. Due to the different system configurations and operations, the two systems will be presented separately.

3.1 The Moving Bed Biofilm Reactor

The Norwegian University of Science and Technology (NTNU) and Kaldnes Miljoteknologi A/S (KMT) of Norway developed the MBBR system employed for this study. The system is sometimes referred to as KMT reactor or Kaldnes MBBR. The system is currently used in 16 different countries all over the world and over 60 plants are either in operation or under construction (www.kmt.no).

The system consists of a reactor vessel containing mixed liquor suspended solids with specially designed carrier media suspended and kept in constant circulation. A screen is provided at the outfall end of the reactor to keep media from clogging the effluent spout or passing out of the reactor. Experiments are currently underway to increase the carrier media size. This will enable screens with larger openings to be used and thus prevent the use of primary settlers in most MBBR system configurations (Odegaard et al., 1994).

The carrier medium is made of polyethylene and has a specific gravity of 0.95. The medium is shaped in a form of a wheel (see Figure 1.1) and has a width of 7 mm and diameter of 10 mm. It is reinforced in the inside with a cross, which provides harborage for microorganisms. The effective specific area of the medium is 500 m^2/m^3. The media volume usually does not exceed 70% of the reactor volume.

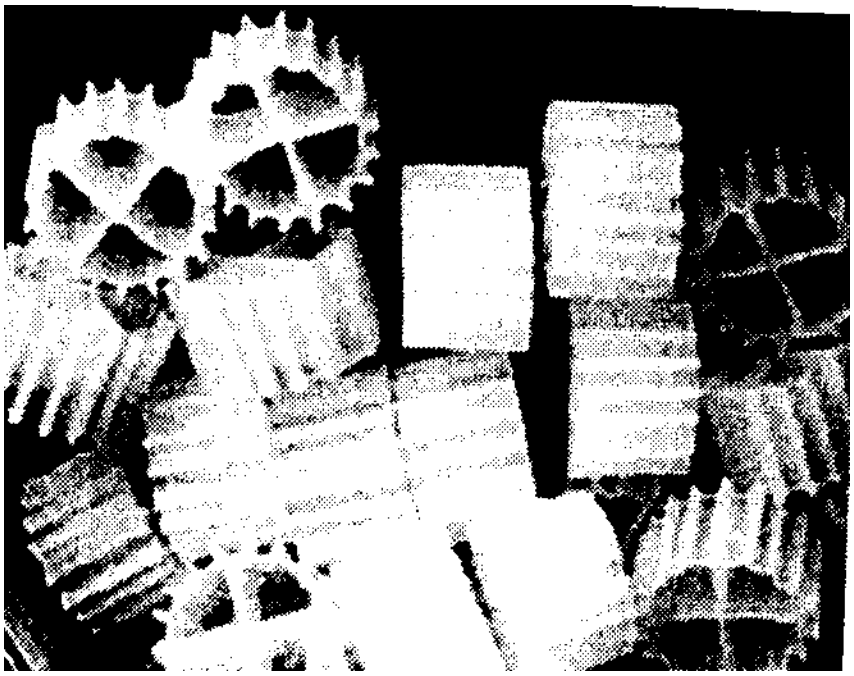


Figure 1.1 MBBR carrier elements by Kaldnes

The system finds several uses in both industrial and municipal wastewater treatment. Several configurations are possible to meet different treatment objectives. The different flow diagrams and treatment objectives are presented in Figures 1.2(a) and 1.2(b). In situations where phosphorus removal is a consideration, chemical coagulation is incorporated as a treatment step (see Figure 1.2b), either in pretreatment or in post treatment.

3.1.1 Factors Affecting Performance

The high specific area of the carrier media, which allows very high biofilm concentrations in a small reactor volume, controls the system performance. It was reported that typical biofilm concentrations range from 3000 to 4000 g TSS /m^3 (Odegaard et al., 1994), which is similar

to values obtained in activated sludge processes with high sludge ages. It was inferred that, since the volumetric removal rate in the MBBR is several times higher than that in the activated sludge process, the biomass in the former are much more viable (Odegaard et al., 1994). Very high performance efficiencies have been reported for different processes, as summarized Table 1.1 (Odegaard et al., 1994).

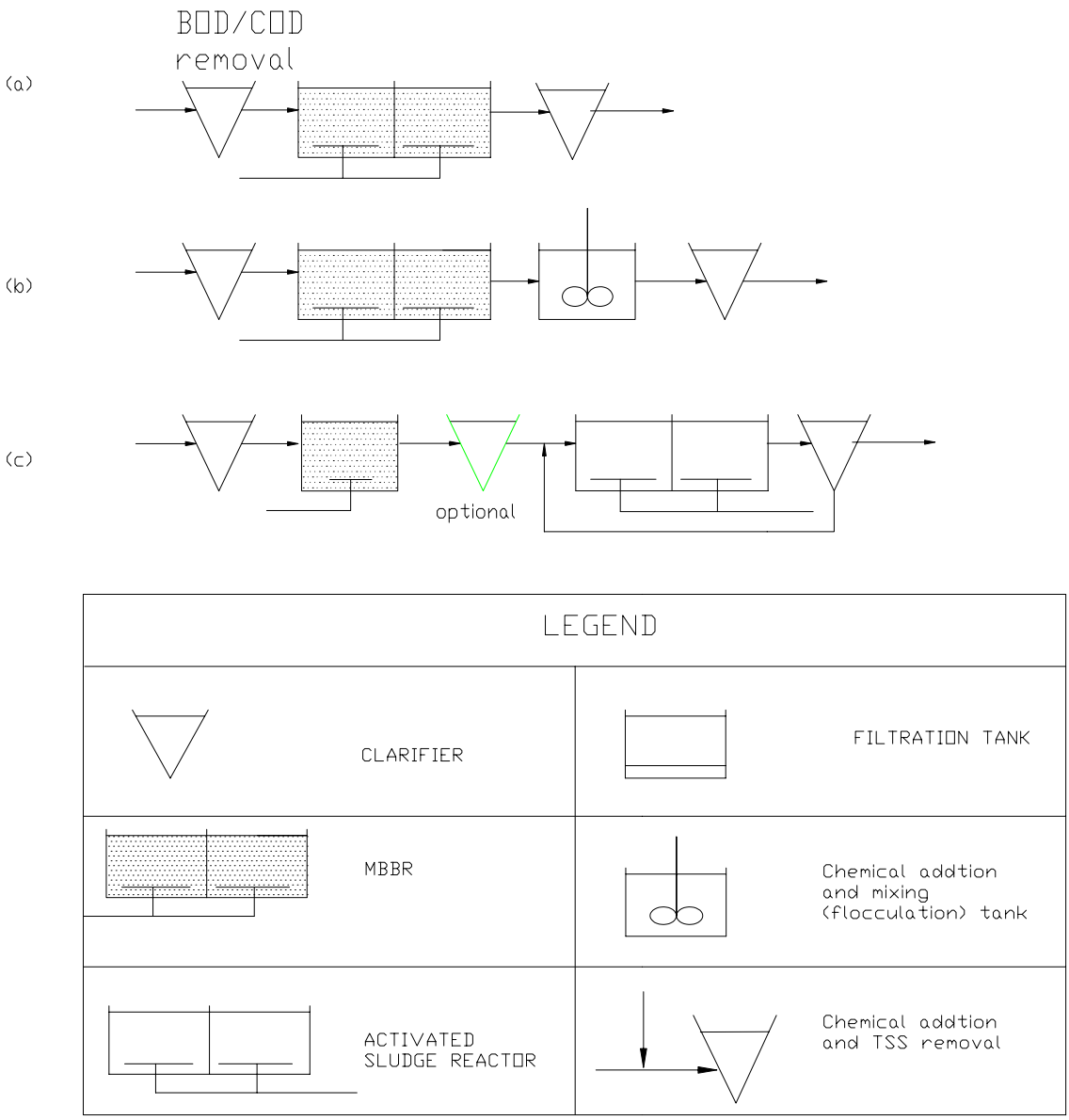


Figure 1.2(a) Different MBBR configurations for BOD5 removal.

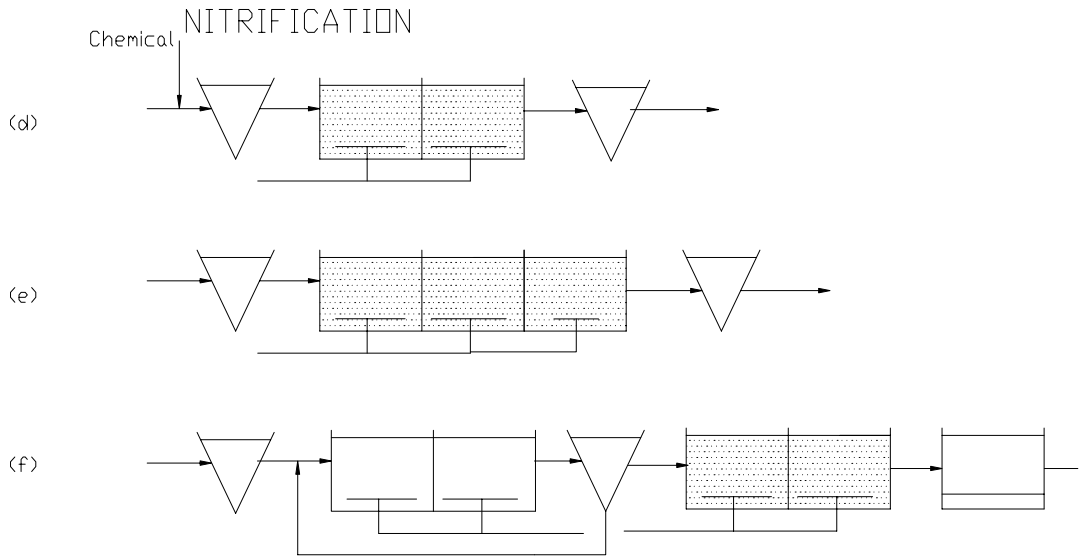


Figure 1.2(b) Different MBBR configurations for nitrification.

Table 1.1 Typical design values for KMT reactors at 15° C[a]

Purpose	Treatment achieved % removal	Design loading rate g/m^2d [b]	Design loading rate kg/m^3d[c] at 67 % fill
<u>BOD-removal</u>			
High-rate	75-80 BOD_7	25 BOD_7	8 BOD_7
Normal rate	85-90 BOD_7	15 BOD_7	5 BOD_7
Low rate	90-95 BOD_7	7.5 BOD_7	2.5 BOD_7
<u>Nitrification (O_2>5mg/L)</u>			
BOD-removal stage (O_2>3mg/l)	90-95 BOD_7	6 BOD_7	2.0 BOD_7
NH_4-N >3mg/L	90 NH_4-N	1.00 NH_4-N	0.35 NH_4-N
NH_4-N <3mg/L	90 NH_4-N	0.45 NH_4-N	0.15 NH_4-N
<u>Denitrification</u>			
Pre-DN (C/N>4) (gBOD7/NO_3-N equiv)	70 NO_3-N	0.90 NO_3-N	0.3 NO_3-N
Post-DN (C/N>3) (gBOD7/NO_3-N equiv)	90 NO_3-N	2.0 NO_3-N	0.7 NO_3-N

a. organic loads based on BOD_7, which is unconventional. BOD_7 approx. equals 1.17 x BOD_5 (Rusten et al., 1994)

b. unit refers to surface area of biofilm surface

c. volume refers to total reactor volume

d. Odegaard et al., 1994

The percent of reactor volume comprised of media is limited to 70%, with 67% being typical (Odegaard et al., 1994). However, the percentage of media required is based on wastewater characteristics and specific treatment goals. Values lower than 67% are frequently used.

Other factors reported to affect performance are flow and mixing conditions in the reactor. Adequate turbulence is ideal for efficient system performance. The nature of the carrier media used requires development of a very thin, evenly distributed and smooth biofilm to enable transport of substrate and oxygen to the biofilm surface. In this regard, thick and fluffy biofilms are not desired for this system. Adequate turbulence sloughs off excess biomass and maintains adequate thickness of biofilm. Biofilm thickness less than 100 μm for full substrate penetration is usually preferred. Adequate turbulence also maintains flow velocities necessary for effective system performance (Odegaard et al., 1994). Extremely high turbulence detaches biomass from the carrier and therefore is not recommended. In addition, collision and attrition of media in the reactor causes biofilm detachment from the outer surface of the Kaldnes media (carrier media used in experiment). Because of this, the MBBR carrier media is provided with fins on the outside to protect biofilm loss and promote growth of biofilm. The surface area of the fins does not contribute to the specific area reported (Odegaard et al., 1994). The effective area of the MBBR carrier medium is reported to be 70% of the total surface area due to less attachment of biofilm on the outer perimeter of the media. This can be seen with the media containing attached biofilm shown later in Figure 1.8.

Due to the uncertainty in the actual area covered by the biofilm on the surface of the carrier, it is preferred to report system performance in terms of reactor volume instead of media surface area, which would have been most appropriate. The reactor volume assessment, however, enables the system to be compared with other systems, which use the entire reactor volume for treatment.

For secondary treatment only, a volumetric loading of 4 to 5 kg BOD_7/m^3-d equivalent to 12 to 15 g BOD_7/m^2-d at 67 % carrier filling (providing 335 m^2 of media surface area per m^3 of reactor volume. See appendix A1 for specific area determination) and 15^oC have been

reported. These reported values and units (BOD_7) are unconventional in the United States of America. However, they conform to Norwegian Standard Methods and their applications to reactor design in the USA must be done cautiously. Rusten et al., 1997 reported that 60 g BOD_5/d is approximately equivalent to 70 g BOD_7/d (i.e. BOD_7 equals 1.17 x BOD_5), although it was not specified whether the BOD values were total or soluble (filtered). Nevertheless, this conversion will be used to translate loading values to the BOD_5 basis.

3.1.2 Organic carbon removal

MBBRs have been applied for organic matter removal (Rusten et al, 1995). The process quickly degrades biodegradable, soluble organic matter. It is reported by Odegaard et al. (1994) that particulate organic matter is partly trapped in biofilm, hydrolyzed and utilized although the hydrolyzed fraction is not significant compared to the readily soluble biodegradable organic matter utilized in the process. Part of the particulate organic matter passed through the reactor unchanged. Figure 1.2(a) for BOD_5 removal shows provision of a clarifier downstream of the MBBR reactor to separate suspended solids in the effluent stream for each of the flow configurations.

.

MBBRs have been used to upgrade existing activated sludge (AS) systems. In one example, the MBBR was proposed to replace a pretreatment system composed of biological filters and humus tank (Sunner et al., 1999). The effluent from the MBBR was fed directly into the AS reactor without an intermediate settling tank. It was anticipated that the sludge age of the AS could be influenced by the solids from the MBBR effluent. Additionally, it was reported that the direct feed improved the settling properties of solids in the secondary clarifiers.

3.1.3 Nitrification.

MBBRs have been used for nitrification (Hem et al., 1994; Odegaard et al., 1994; Rusten et al., 1995a). Although ammonia-N oxidation was not a goal of this research, levels of ammonia nitrogen expected in FPWTS wastewater and found in the wastewater used for the laboratory-scale MBBR study reported in this thesis are similar to that found in municipal wastewater (25 mg NH_3-N/L). Like other aerobic reactors, nitrification is likely to occur if favorable conditions are created (such as low organic matter and dissolved oxygen

concentrations greater than 2 mg O_2/L). At high organic loadings (5 g BOD_7/m^2-d), ammonia removal rates were low (0.2 g NH_3-N/m^2-d) even at high dissolved oxygen concentrations of 7 mg O_2/L (Odegaard et al., 1994). This has cost implications if nitrification is not a treatment objective. Odegaard et al. (1994) and Pastorelli et al. (1997) reported that oxygen levels above 2 and 3 mg O_2/L were needed in order for nitrification to take place.

MBBR systems have been very useful in upgrading schemes. The small footprint of the reactor saves the cost of acquiring land at a high premium for conventional reactors. It has also been used in treating industrial wastewaters from food processing and paper and pulp industries (Broch-Due et al., 1994; Odegaard et al., 1994; Rusten et al., 1992.). Where phosphorus removal is desired, chemical coagulation is incorporated in a pretreatment or post treatment step. If coagulation is used in the pretreatment stage, it was found that suspended solids were removed, leaving low molecular weight soluble organic matter in the influent wastewater stream. The low influent suspended solid concentration increases the overall system efficiency (Odegaard et al., 1994.).

3.2 **Biological Aerated Filter (BAF)**

BAF systems consist of a reactor container, media for supporting biofilm growth, an influent distribution system, an aeration system and an effluent collection system. The use of various media types has been reported (Kent et al., 1996). Most of these are dense (sinking, specific gravity > 1.0) natural materials such as sand, shale expanded clay, etc. However, low density (floating, specific gravity < 1.0) synthetic materials such as polystyrene beads and pure polypropylene have been used (Anderson et al., 1995; Mann et al., 1999, Vedry et al., 1994). The media sizes range from 2 to 6 mm and provide very high specific areas for biofilm attachment. Specific areas of 1000 to 1500 m^2/m^3_{media} have been reported for granular media (Mendoza-Espinosa and Stephenson, 1999). Recommended media depths are in the range of 2 to 4 m (Grady et al., 1999).

BAF systems use submerged attached growth processes for treatment of organic and inorganic pollutants in wastewater, typically under aerobic conditions. They are capable of

high removal efficiencies and come in small footprints. The BAF technology has been used widely in Europe since the early 70s (Mendoza-Espinosa and Stephenson, 1999). In the United States of America, a 2-stage BAF (for carbonaceous BOD removal followed by ammonia oxidation) was constructed at treatment plant at Roanoke, VA to upgrade the existing facility and was reported to be the first of its kind at full scale in the USA (Love et al., 1999). It is, therefore, a fairly new waste treatment process in the USA in recent years.

The nature of the media used in the BAF defines its operational mode. Sinking media BAFs enable influent to be fed in an upflow or downflow mode. However, a floating media is operated exclusively in an upflow mode. In the downflow mode, wastewater flows countercurrent to air. This is found to have a beneficial effect by providing sufficient oxygen in the lower regions of media bed, thus aiding nitrification if it is desired in the treatment process (Grady et al., 1999). In the upflow mode, wastewater flows cocurrent with air. Depending on the position of the aeration system in the reactor, the system is capable of producing different treatment effects. For example, if the aeration system is provided at the bottom, then carbonaceous and ammonia oxidation is predominant and suspended solids are filtered as well (Grady et al., 1999). However, if the aeration system is provided mid way through the media bed, then the bottom half of the reactor filters while the upper section performs carbonaceous removal and ammonia oxidation, depending on the influent C/N ratio.

BAF media eventually clogs due to biomass growth and filtration of suspended solids, creating resistance to flow of liquid and air. Regular backwashing is therefore required to remove excess biomass and particles, allowing treatment to be resumed directly after washing (Smith & Hardy, 1992) and to ensure acceptable performance. It is reported that ineffective backwashing results in the formation of mud balls, poor effluent quality during initial stages of filtration and increased head loss build up during subsequent filter runs (Amirtharajah, 1993; Robinson et al., 1994). Backwashing for floating media is carried out by reversing the flow of water through the system. Backwashing for sinking media could either be cocurrent if operated in upflow mode or countercurrent if operation is downflow. Solids dislodged from the backwash process require collection and disposal.

3.2.1 Performance

The performance of BAFs is defined relative to the specific pollutants, which must be removed. BAF can combine ammonia oxidation, carbonaceous oxidation and solids removal in a single-unit (Grady et al, 1999). In a 2-staged system it is possible to achieve carbonaceous BOD removal in the first stage and ammonia oxidation in the second stage (Gilmore et al., 1999). Reported removal rates for carbonaceous BOD_5, ammonia and nitrates are 4.1 kg BOD_5/m^3-d, 1.27 kg NH_3-N/m^3-d and 5 kg NO_3-N/m^3-d, respectively, where the volume reflects reactor empty bed volume (Dillon et al., 1990; Mendoza-Espinosa and Stephenson, 1999). Very high pollutant reductions (about 90%) in pollutants have been reported for both upflow and downflow reactor configurations (see Table 2).

Media depth is among those factors influencing performance and is recommended to be between 2 and 4 m for full-scale BAF plants (Stensel et al., 1988). The granular nature of media provides a very large surface area of attachment for biofilm growth, resulting in high removal efficiencies and small footprints. The media should be chemically inert and resistant to attrition (Valentis and Lesavre, 1989; Kent et al., 1996). The size of the granular media used affects the efficiency of treatment in term of physical removal of suspended solids and providing adequate surface area for biomass attachment (Smith and Marsh, 1995). It also affects the frequency of backwashing. The use of large media (greater than 6 mm) causes a reduction in nutrient and suspended solids removal through high void spaces and reduced surface area for biofilm growth (Stensel et al., 1988). However, it reduces the backwashing frequency and, hence, the overall maintenance and operational cost of the system. Very small sized media may result in short filter run time and frequent backwashing. In this regard, optimizations of the system in terms of media type and backwashing rates are needed for efficient running of BAF systems (Mendoza-Espinosa and Stephenson, 1999). Air and water flow rates used in backwashing are normally between 0.43 and 0.52 m^3_{air}/ m^3_{media}-min and 0.33 to 0.35 m^3_{water}/m^3_{media}-min, respectively (Condren, 1990). The amount of liquid required for backwashing full scale BAF plants is 12 to 35 % of total daily effluent flow for secondary treatment and 2% if the reactor is for tertiary treatment (Canler and Perret, 1994; Dillon and Thomas, 1990; Whealer and Cooper-Smith, 1995).

Reported flow velocities for efficient running and performance of BAFs are between 1 and10 m^3/m^2-h (Stensel et al., 1988; Paffoni et al., 1990; Vedry et al., 1994). Over 80 % ammonia removal was reported by Husovitz et al. (1999) at a hydraulic loading rate between 5.1 to 15.8 m^3/m^2-h, for an average ammonia loading rate of 1.35 kg NH_3-N/m^3-d and an organic loading rate of 0.65 kg $cBOD_5/m^2$-d. It was possible to have over 90% NH_3-N reduction of diluted wastewater and storm water at a flow rate of 30 m^3/m^2-h in upflow biofilters (Peladan et al., 1997).

3.2.2 BOD_5 removal

BAFs have been used extensively for carbonaceous BOD removal combined with suspended solids filtration (Mendoza-Espinosa and Stephenson, 1999). Removal of carbonaceous pollutants in BAFs is carried out by suspended solids filtration and hydrolysis, absorption of soluble and colloidal organic matter, and oxidation (Phipps, 2001; Stensel et al., 1988). BAFs are capable of producing high volumetric organic removal rates in secondary processes and high quality effluents. Organic removal rates of 4.1 kg BOD/m^3-d have been reported in partially nitrifying BAFs (Mendoza-Espinosa and Stephenson, 1999). The removal rate is very high compared to that typically found in trickling filters, oxidation ditches and activated sludge plants of 0.06 kg BOD/m^3-d, 0.35 kg BOD/m^3-d and 0.42 kg BOD/m^3-d respectively (Smith and Hardy, 1992; Pujol et al., 1994). See Table 1.2 for carbonaceous matter removal for various reactor configurations and organic loading rates. The system performance is affected by high peak loading, resulting in organic breakthrough due to limited treatment efficiency (Ruffer and Rosenwinkel, 1984).

3.2.3 Nitrification

BAFs can be used for nitrification in systems that combine suspended solids, chemical oxygen demand (COD) and ammonia removal (combined secondary and tertiary treatment) or in systems that remove suspended solids and ammonia (tertiary treatment). Two to three percent of influent ammonia is used to synthesize cells while the rest of ammonia is oxidized. (Akunna et al., 1994; Cecen and Gonenc, 1995). Nitrification rates of 1.27 kg NH_3-N/m^3-d are obtainable and depend on influent ammonia concentrations and hydraulic residence time.

Ammonia removal using BAFs for wastewater treatment are reported in Table 1.3 (Gilmore et al., 1999; Mendoza-Espinosa and Stephenson, 1999).

Table 1.2 Carbonaceous matter removal using biological aerated filters for wastewater treatment[a]

Reactor configuration	Reactor volume (m3)	Influent conc. (mg/L)	Organic loading rate (kg/m3-day)	Removal (%)	HRT (h)	Reference
Upflow	0.0085	3000–3500	3.3-15.4	33-82	4.5-23	Costa Reis and Sant'Anna, 1985
Downflow	0.3	424 (avg)	<9.2 COD	>90	0.4-0.76	Dillon and Thomas, 1990
Downflow	0.14	< 200	10.5 COD	Approx. 55	0.5	Rogalla et al., 1990
Downflow	0.2 – 0.3	324 (avg.)	< 15	86	0.4-0.6	Rogalla and Bourbigot, 1990
Downflow	0.2	350	8-10	90	-	Bacquet et al., 1991
Downflow	22 0.1 0.7	13.6 COD_{Mn} 9 $sCOD_{Cr}$	- 2.3COD_{Cr}	12 20 30	- - -	Sakuma et al., 1993
Upflow and downflow	Full scale plant 31.5– 90.3	35-607	0.5-6.3	55-85	-	Pujol et al., 1992
Downflow	4 cells, 143 m3 each	131	1.5 COD	93	1.3	Wheale and Cooper-Smith, 1995
Upflow	Full scale, two units 151.2 m3 each	25-43	Approx. 2.4 (per unit)	48-70	21	Peng et al., 1995
Upflow	Full scale, 8 cells 219 each	109-250 (BOD)	4	>93	-	Brewer, 1996
Upflow	Lab-scale, 2 stage, 0.0093 each	257 (BOD_5)	2.4	>90	3.88	Asiedu, 2001

a. culled from (Mendoza-Espinosa and Stephenson, 1999).

Table 1.3 Ammonia removal using biological aerated filters for wastewater treatment [a].

Reactor configuration	Reactor volume (m3)	Influent conc. (mg/L as N)	Organic loading rate (kg/m^3-day)	Removal (%)	HRT (h)	Reference
Upflow	0.55	22.7-37.5 NH_3		62-84	6-12.9	Faup et al., 1982.
Upflow (Intermittent aeration)	4.18	17.8 NH_3	0.03-0.05	88.4	7-8	Iida and Teranishi, 1984.
Downflow	1	13-20.9 NH_3	0.39-0.84 (NH4-N)	90-99	0.5-0.9	Rogalla and Payraudeau, 1988.
Upflow	Full scale		<0.46	-		Carrand et al., 1990
Downflow	0.3	40.7 NH_3	<0.58	65-100	0.32-0.83	Dillion and Thomas, 1990
Downflow	0.14	< 20 NH_3	<0.6	-	1	Rogalla et al., 1990.
Downflow	0.2-0.3	40 (TKN)	1 (NH_4-N)	> 95	1	Rogalla and Bourbigot.,1990
Downflow	22 0.1	11 NH_3 23.4 NH_3	0.9 (NH_4-N; 1.2 removal rates)	57 78	- -	Sakuma et al., 1993
Upflow	Full scale, 333	Approx. 22 NH_3	Approx. 0.9 (NH4-N)	89	-	Vedry et al., 1994
Upflow	0.81-1.16	-	3 (NH4-N)	80	-	Peladan et al., 1996
Upflow	Full scale, 18 cell, 292 m3 each	13-28 NH_3	0.15	92-96	-	Brewer, 1996
Upflow	Pilot plant 2-stage. Reactor volume:1st ($1.1\ m^3$); 2nd ($1.04\ m^3$)	10.3 ± 0.6 NH_3-N	<0.6	> 90 (winter period)	-	Gilmore et al., 1999
Upflow	Pilot plant 2-stage. Reactor volume:1st ($1.1\ m^3$); 2nd ($1.04\ m^3$)	15-40 NH_3-N	1.0	80-100 (summer period)	-	Husovitz et al., Unpublished paper

a. culled from (Mendoza-Espinosa and Stephenson, 1999).

Performance of the system is also influenced by temperature variations. Love et al. (1999) reported high (between 80 to 100 %) ammonia removal efficiencies for their pilot study for summer conditions with average temperatures of (20.1 ± 3.3^{o}C). During winter conditions (12.4 ± 0.1 ^{o}C) nitrification rate was greater than 90 % in a 2nd stage upflow BAF column when ammonia loading to the column was either equal to or less than 0.6 kg NH_3-N /m^3-d (Gilmore et al., 1999).

3.3 **Modeling Biofilm Systems**

Substrate transport to cells in biofilms is essential to maintain a viable biofilm for wastewater treatment. Aggregation of cells creates significant gradients in substrate concentrations. Since mass transport of bulk substrate from outside the biofilm to inside is driven by concentration differences, bacteria on the inside of biofilms are often exposed to substrate concentrations substantially lower than that measured in the bulk liquid. Therefore, the rates of substrate utilization and cell growth are not uniform throughout the depth of a biofilm, but depend on the cell location within the film (Rittmann and McCarty , 1992). Three concentration profiles are possible in a biofilm and are noted in Figure 1.3. They are:

- Deep biofilm, in which the substrate concentration approaches zero at some point in the biofilm,
- Shallow biofilm where S_f remains above zero at all points in the film, and
- Fully penetrated biofilm, which occurs when the substrate concentration has negligible gradient.

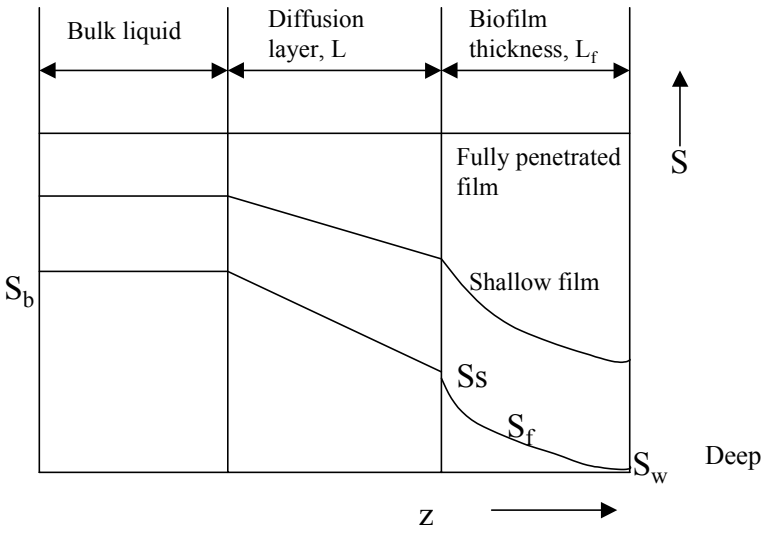


Figure 1.3 Ideal biofilm structure as it relates to substrate penetration profile.

Table 1.4 Kinetic parameters used in pseudo analytical model

Symbols	Parameters	Units	Value	Reference
L	Diffusion layer	cm	0.01	Rittmann et al., 2001
L_f	Thickness of biofilm	cm	390 μm	Asiedu, 2001
Y	True yield coefficient	mg/mg	0.43	Asiedu, 2001
K_s	Half-saturation constant	mg/cm^3	-	
X_f	Biomass density	mg/cm^3	319	Asiedu, 2001
q	Substrate utilization rate	d^{-1}	varies	Asiedu, 2001
$\hat{q}$	maximum substrate utilization rate	d^-	6.62	Asiedu, 2001
D	Molecular diffusion coefficient of substrate in water	cm^2/d	0.8	Rittmann et al., 2001
D_f	Effective molecular diffusion of substrate in biofilm	cm^2/d	0.16	Rittmann et al., 2001
J_s	Flux into biofilm	mg/cm^2-d	-	
A_s	Media specific surface area	m^2/m^3	500	www.kmt.no
V	Reactor volume	m^3	-	
S_{so}	Influent substrate concentration	mg COD/cm^3	0.513	UTD Inc., 2000
S_b	Effluent or bulk substrate concentration	mg COD/cm^3	0.05	UTD Inc., 2000
S_f	substrate concentration in biofilm			
b'	Overall biofilm loss coefficient	d^{-1}	0.02	Rittmann et al., 2001

The pseudo-analytical model was developed originally by Rittmann and McCarty (1980a) and was improved by Saez and Rittmann (1991). The model invokes several assumptions. These include:

1. The biofilm has a uniform biomass density,
2. It has a locally uniform thickness,
3. The depth of the effective diffusion layer reflects the degree of external-mass transport resistance. Internal mass transport resistance is represented by molecular diffusion.
4. Substrate concentrations in the biofilm (S_f) are usually lower than S_b in the bulk liquid.

For steady state conditions inside the biofilm layer, diffusion (Fick's second law) and the rate of substrate utilization occur simultaneously, as described in equation 1.1.

$$0 = D_f \frac{d^2 S_f}{dz_f^2} - \frac{\hat{q} X_f S_f}{K_s + S_f} \qquad \text{(Equation 1.1)}$$

where the parameters and typical values for domestic wastewater treatment (where known) are defined in Table 1.4.

Two boundary conditions are required to solve equation 1.1. The first assumes that there is no substrate flux into the attachment surface (that is the attachment surface is inert), and is expressed mathematically as:

$$\frac{dS_f}{dz} = 0 \text{ at } z = L_f \qquad \text{(Equation 1.2)}$$

The second boundary condition occurs at the biofilm/water surface, where substrate must be transported from the bulk liquid to the outer biofilm surface. Fick's first law applies.

$$J_s = \frac{D}{L}(S_b - S_s) \qquad \text{(Equation 1.3)}$$

At steady state, the flux of biomass growth is assumed to equal biomass loss across the biofilm. This is reflected in equation 1.4.

$$0 = YJ_s - b' X_f L_f \qquad \text{(Equation 1.4)}$$

The solution to differential equation (1.1), was provided by Rittmann and Saez (1987). Their approach was an improved version of the pseudoanalytical approach by Rittmann and McCarty 1980a.

The solution to the differential equation is presented in dimensionless form as follows:

$$J_s^* = \xi . J_{s,deep}^* \qquad \text{(Equation 1.5)}$$

where

$$\xi = \tanh\left[\alpha\left(\frac{S_s^*}{S_{\min}^*} - 1\right)^{\beta}\right] \qquad \text{(Equation 1.6)}$$

and

$$J_{s,deep}^* = \left\{2\left[S_S^* - In\left(1 + S_s^*\right)\right]\right\}^{0.5} \qquad \text{(Equation 1.7)}$$

where α and β are functions of S^*_{min} defined by Saez and Rittmann (1987). Equation (1.6) was assumed to be one since the biofilm thickness determined during the study suggested that the biofilm in the MBBR media was deep. S_s^* is a dimensionless substrate concentration at the liquid/biofilm interface.

$$S_b^* = S_s^* + \frac{\tanh\left[\alpha\left(\frac{S_s^*}{S_{\min}^*} - 1\right)^{\beta}\right]\left\{2\left[S_s^* - \ln(1 + S_s^*)\right]\right\}^{0.5}}{K_L^*} \qquad \text{(Equation 1.8)}$$

$$J_s^* = K_L^*(S_b^* - S_s^*) \qquad \text{(Equation 1.9)}$$

$$J_s = J_s^*(K\hat{q}X_f D_f)^{0.5} \qquad \text{(Equation 1.10)}$$

where parameters are defined in Table 1.5.

Table 1.5 Relationship between dimensionless variables and dimensioned variables.

Dimensionless variable	Dimensioned variables
S_b^*	$\frac{S_b}{K_s}$
K_L^*	$\frac{D}{L}\sqrt{\frac{K}{\hat{q}X_f D_f}}$
$S_{\min}^*$	$\frac{`S_{\min}}{Ks} = \frac{1}{K_s}\left(\frac{b'}{Y.\hat{q} - b'}\right)$

CHAPTER 4.0 MATERIALS AND METHODS

4.1 Biotreatment System Overview

Two laboratory-scale biological systems, a moving bed biofilm reactor (MBBR) and biological aerated filter (BAF) were designed, constructed and operated for 5 months. Wastewater from one of Virginia Tech (VT) sewer lines was used for this study. The wastewater was pumped daily with a submersible pump (0.5 HP, Teel Commercial and Industrial Pump) from a manhole into two 55-gallon wastewater storage tanks (WTSs), which were hydraulically connected. The wastewater was kept completely mixed with four small submersible pumps (1/10 HP, Little Giant Co., USA). The pumps were cleaned daily of debris to ensure effective mixing. The number of pumps used in each WST varied depending on the concentration of total suspended solids (TSS) estimated to be in the wastewater.

Influent and effluent samples taken daily during the reactor study were either grab (first 24 days) or composite (after day 24) for the MBBR. For the 2-stage BAF system, some grab samples were collected from the first column; however the overall system performance was determined from the 2^{nd} column effluent composite samples taken from day 52 to day 150. To obtain composite samples, peristaltic pumps were programmed to pump liquid six times a day at 3-hour intervals. The samples were collected in 1-liter containers, which were kept in a $4^{o}C$ constant temperature refrigerator. The samples were transported through 1/4-inch PVC tubes. The tubes were flushed daily prior to the start of the sampling process. Finally, reactors were allowed to develop biomass naturally so exogenous seed was not used.

The temperature in the shed was controlled to maintain a reactor temperature of 20 to $25^{o}C$. To achieve this, the shed was insulated with 1/2-inch thick polystyrene board. During winter conditions, a ceramic heater furnished with a thermostat device was used to maintain the desired temperature. During spring, heating was not necessary most of the time. An air conditioner was installed during the course of the study to control excessive warm-weather temperatures in the shed. A thermometer was used to monitor the reactor temperature.

4.2 Wastewater Characterization and modification

Target influent wastewater characteristics anticipated at the FPWTS were given as 300 mg/L BOD_5 and 133 mg/L TSS (UTD Inc, 2000)

It was necessary to characterize this wastewater to determine if its composition needed to be altered to meet the influent targets. The VT wastewater was characterized for periods when school was in session and recess. This was necessary because the period of the study spanned over school periods and breaks. Samples for the characterization study were collected for one month at about 2:00 p.m. (anticipated wastewater peak loading period) daily and analyzed for unfiltered BOD_5 ($TBOD_5$) carbonaceous BOD_5 ($CBOD_5$), volatile suspended solids (VSS), total suspended solids (TSS) and ammonia-N. Standard laboratory procedures were used (APHA, 1998) and are described in more detail below.

Based on the results from the characterization study, the wastewater was modified during school sessions and recesses to meet the BOD_5 and TSS targets. During periods (both school and recess periods) when TSS concentration in the influent stream was found to be approximately equal to the target value, the two WSTs were kept completely and continuously mixed with the 'Giant' submersible pumps. In order to meet the BOD_5 target, soluble BOD supplement was prepared daily and added to the VT wastewater. The BOD supplement consisted of a stock solution of 4.7 g COD/L sugar (dextrose, galactose and fructose), 126 g COD/L organic acid (acetic acid in the form of potassium acetate and glycerol) and 10 g COD/L protein (yeast extract). An established relationship between BOD_5 and biodegradable chemical oxygen demand (COD_{BO}) of 1: 1.71, reported by Grady et al. (1999), was used to determine the BOD_5 amount of each constituent of the supplement.

Due to daily variations in the BOD concentration, probably due to infiltration and rainfall, it was necessary to have a simple means of determining the amount of supplement to be added daily (see typical sample calculation in Appendix A1). A relationship was established between VT sewage BOD_5 and its dissolved organic carbon (DOC). DOC values were relatively simple to obtain and thus served as a quick means of estimating the amount of

supplement needed to meet the BOD_5 target. Additionally the DOC to BOD_5 correlation was useful in determining the proper dilutions to use when preparing effluent BOD_5 assays.

A settling experiment was conducted to develop a protocol for reducing the TSS concentration in the wastewater during incidents of high TSS. Six graduated cylinders were filled with VT wastewater and allowed to settle. The supernatant in each cylinder was analyzed for TSS concentration at 1-hour intervals. A wide bore pipette was used to carefully draw samples directly from the cylinder to avoid re-suspension of settled solids. A plot of TSS concentration with time was created and used to determine the time to settle the designated amount of suspended solids from the VT sewage. In the WST, a number of 'Giant' submersible pumps were used at a time to achieve the desired influent TSS. Although this procedure was quite successful, it was also possible to obtain the desired TSS in the influent wastewater by putting the WST in series. With this arrangement, the second tank served as a wastewater intake tank which was completely and continuously mixed while the first tank was not mixed and allowed to settle suspended solids. Influent to both the BAF and MBBR was pumped from the completely mixed (second) tank.

4.3 **Reactor Design and Operation.**

4.3.1 MBBR

The laboratory scale MBBR was an 11-liter rectangular Eckenfelder basin with a 2 mm perforated baffle, constructed from a Plexiglas. The baffle was located at the downstream end to keep media out of the reactor outlet and to maintain a less turbulent region for settling residual suspended solids. This region was about one tenth of the reactor volume and served as the intake point during composite sampling of the effluent. A schematic of the system is given in Figure 1.4.

The aeration system consisted of a manifold of 1/4 inch id cPVC pipe, a vertical pipe connector to the manifold and rising above the liquid level in the reactor to prevent flow of liquid to the compressor when the compressor was turned off. A temperature resistant

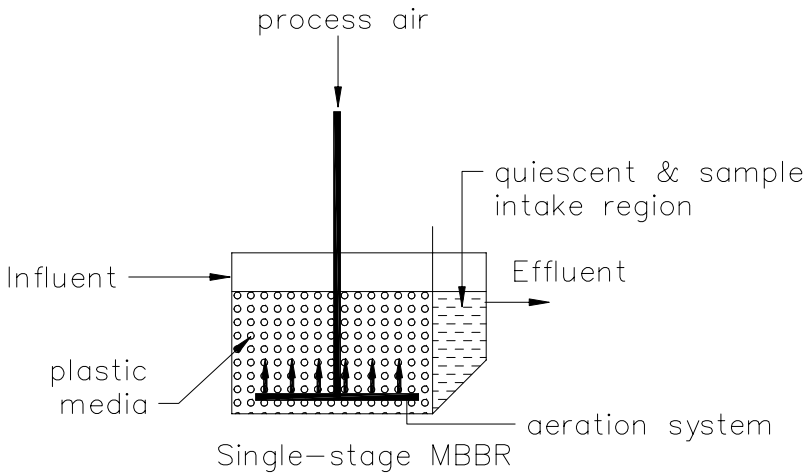


Figure 1.4 Schematic of MBBR system with 40 % media (same configuration for 67 % media volume).

flexible tube, connected the vertical pipe and the compressor. An airflow meter was installed to measure air supply to the reactor. Air supply was maintained between 0.8 and 1.0 cfm. Valves mounted on the compressor were used to regulate the airflow. The aeration system provided coarse bubbles and kept the media in circulation.

A two head peristaltic pump with a variable flow rate of 4.8 to 480 ml/min and an rpm of 6 to 600 was used to pump influent waste to the reactor. Each pump was provided with a controller, which enabled flow rates to be adjusted. The influent was pumped from the 55-gallon intake WST through 1/4-inch id PVC tube, which was connected to the pump head tubes by fittings. The tube was occasionally flushed to remove built up solids by operating the peristaltic pump in reverse mode. It was possible to maintain a continuous influent flow of wastewater supply except during periodic maintenance of the pumps.

The MBBR systems were operated under a range of conditions. Based on previous studies reported in Waterlink technical papers on the system, a media volume representing 40 % of the reactor volume was used (www.kmt.no). An initial theoretical empty bed HRT of 90 minutes was selected, which required a flow of 122 ml/min. Given a target BOD_5 influent

concentration of 300 mg/L, a BOD surface area loading rate (SALR) of 24 g BOD_5/m^2_{media}-day was initially applied. The system was operated under these loading conditions for 80 days, after which the flow rate was reduced to 80 ml/min, resulting in a SALR of 15 g BOD_5/m^2_{media}-day and an HRT of 2.4 hours. The lower SALR was recommended by Waterlink Inc (vendor for the MBBR technology). After day 120, a new single stage reactor with a media of 67 % was set up to ascertain the effect of increased media volume on effluent quality. Flow rates, aeration and reactor temperature were 80 ml/min, 0.8 to 1.0 cfm and 20 to 25 ^oC, respectively, similar to the 40 % media volume single stage reactor.

A two-stage MBBR was also studied in an attempt to improve process performance. A 2nd reactor was introduced downstream of the single stage 40 % media volume reactor to achieve this (see Figure 1.5). The 2nd stage reactor was similar to the 1st stage in terms of reactor and media volume and aeration system, except that the baffle in the first stage was removed and introduced in the 2nd stage. Effluent from the 1st stage flowed through a 4-in. long 1/4 inch. id PVC tube to the 2nd stage. Biofilm was allowed to build naturally on the new media provided in the 2nd stage reactor without addition of exogenous seed (media with biofilm) from the first. The 2-stage system was operated for 33 days.

Due to the low solids retention capabilities of the system, return of solids back to the reactors was initiated on day 145 of the study to increase the SRT of the suspended solids and determine its effect on system performance. The system configuration for the return sludge is shown in Figure 1.6.

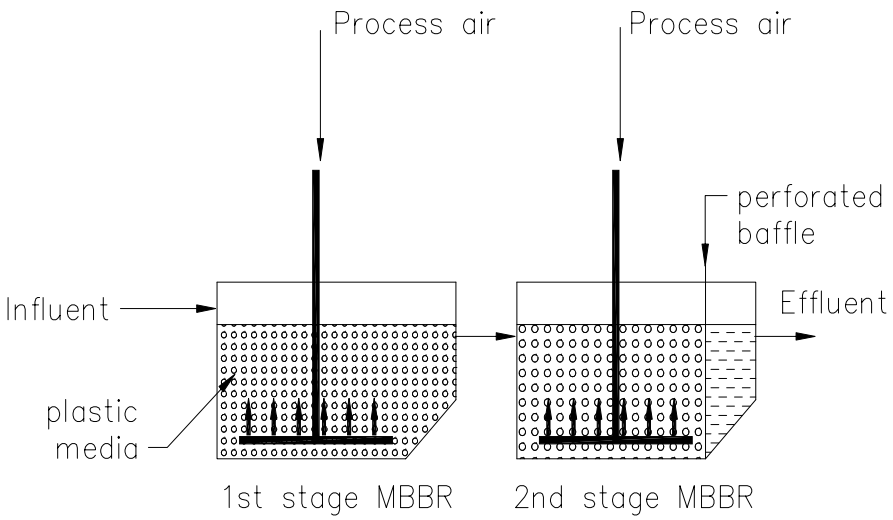


Figure 1.5 Schematic of 2-stage MBBR with 40 % media volume.

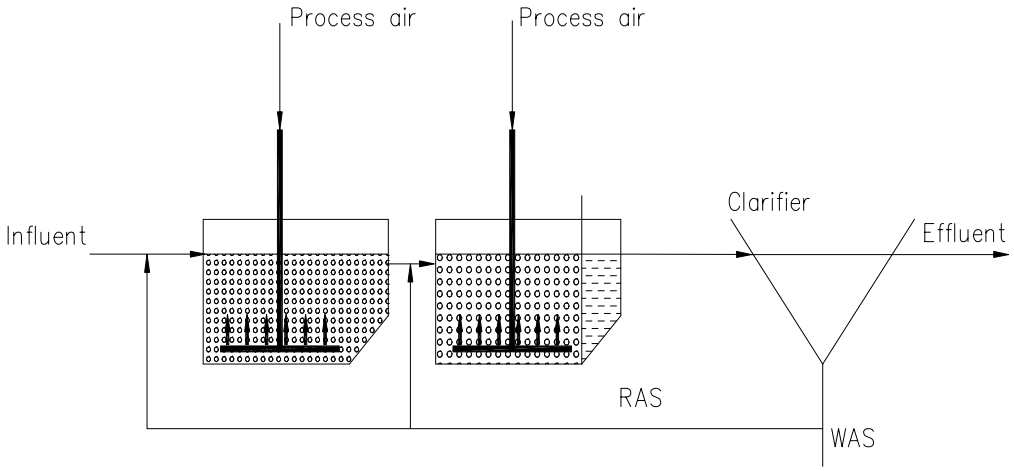


Figure 1.6 Schematic of MBBR with 40 % media operated with return sludge.

4.3.1.1 Stimulus – Response Study (Tracer Study).

Prior to the start of operation of the MBBR, a tracer study was performed to determine the actual HRT of the system. The reactor was filled with 40 % media volume (based on reactor volume) and water to the desired reactor volume. A known concentration of potassium chloride (0.1 M) solution was prepared. A conductance meter (YSI model 32, Yellow Spring Instrument Co. Inc.) was used to standardize the solution by relating different concentrations of the solution to conductance (APHA, 1999). The system was fed a step input of a known concentration of potassium chloride at the design flow rate of the reactor. The initial conductance was determined and samples were collected at 2 minutes interval until the conductance remained fairly constant. The procedure was repeated twice. Ideal and non-idealized flow models for analyzing flows through reactors were used to analyze the data and the actual HRT of the system determined. (Octave Levenspiel, 1972). The tracer test data were manipulated by normalizing the corrected conductance (after subtracting the background potassium chloride conductance) of the effluent samples at time *i (Ci*) with the conductance of the feed (Co). A plot of Ci/Co versus time yields an F-curve. The F-curve is related to the E-curve or the exit age distribution curve by equation (1.11).

$$E = \frac{dF}{dt},\ F = \int_0^t E dt \qquad \text{(Equation 1.11)}$$

Predictions can be made about flows through a vessel once the E-curve is derived. The mean residence time, $\bar{t}$, and the spread of the curve is given by the variance σ^2 , are used to characterize the E-curve.

For normalized distributions with discrete measurement, these parameters can be determined from equations (1.12) and (1.13).

$$\bar{t} = \sum t_i E_i \Delta t \qquad \text{(Equation 1.12)}$$

$$\sigma^2 = \sum t_i^2 E_i \Delta t - \bar{t}^2 \qquad \text{(Equation 1.13)}$$

Other models used to characterize flow behavior are the tanks-in-series model and the PFR model with dispersion.

This model may also be used to determine the flow configuration based on continuous flow stirred tank reactors (CSTRs)-in-series of equal total volume that emulate the flow conditions in the real reactor. As the number of tanks in the model, N, increases, the response approaches that of a PFR. The number of CSTRs in series that would most closely approximate the behavior of the reactor can be determined from the mean and variance of the E-curve by the following equation:

$$N = \frac{\bar{t}^2}{\sigma^2} \qquad \text{(Equation 1.14)}$$

in which N is the number of equally sized tanks (Teefy et al., 1990).

4.3.2 BAF

The upflow laboratory-scale BAF system consisted of 2 reactor columns in series. The column was constructed with 3" id, 1/4 inch. thick clear PVC pipes. Since the headroom in the shed (the housing unit for the reactors) was less than 9 feet (2.7 m) and the ISO containers to be used for the FPWTF were only 8 feet (2.4 m) tall, it was necessary to stage the reactor to achieve the overall 9 feet (2.7 m) media depth recommended for BAFs. One-foot depths were provided between the bottom of the media and the influent, and between the top of the media and the effluent. An additional 2-inch freeboard was maintained above the effluent outlet, giving an overall column depth of 80 inches for each reactor. The pipes were coupled at 16 in. centers using 3 in. diameter PVC couples with clean outs and a PVC end cap at the bottom. At each coupled point on the column, a port was provided that allowed air and water injection during backwashing operation of the BAF. 1.37 m (54 inches) depth of floating polystyrene beads with a density of 24 kg/m^3 (1.5 lb/ft^3) and specific area of 2500 m^2/m^3 (762 ft^2/ft^3)((Radva, Radford, VA) was used for media in each column. The total empty bed volume per column was 6.25 mm^3 (0.22 ft^3). About 4 inch depth of 1/2 inch. size crushed aggregates were provided at the bottom of each column to disperse the influent wastewater and air. A line aeration system of perforated 1/4 inch PVC pipe provided coarse

bubbles. Air supply was through a 3/8-inch high temperature resistant tube (to resist the heating effect of compressed air) and was supplied by a 1.0 HP, 4.7 cfm, 20-psi piston air compressor (GAST brand). The tubes were furnished with check valves (one for each column) to prevent back flow of liquid from the column to the compressor during backwash operations and periods when air supply was halted. Peristaltic pumps were used to feed wastewater at a flow rate of 80 ml/min into the reactor. This corresponded with a target mass-loading rate (MLR) of 2000 g BOD_5/m^3_{media} vol.-day. This loading was selected based on experience obtained with BAF studies in Radford, VA. (Love et al., 1999).

A 16" x 16" (16 openings in 1 inch wire cloth) size screen with 1.52 mm size openings supported on a rigid wire screen with 5 mm x 5 mm openings was mounted approximately 450 mm (18 in.) below the top of the reactor and just below the topmost port to retain the floating media. Figure 1.7 shows a schematic diagram of the reactor configuration. A system of PVC tubes was used to transport influent and effluent wastewater to and from the reactor. Since the system was a 2-stage BAF, the effluent from the first stage was collected in a 20-liter bucket and pumped as influent for the 2^{nd} stage reactor. Retention time in the bucket was about 30 min. Coarse bubble aeration was provided at the bottom of each column. Dissolved oxygen was measured using the DO probe (YSI Model 58, Yellow Spring Instrument Co. Inc., USA) in the liquid above the screen.

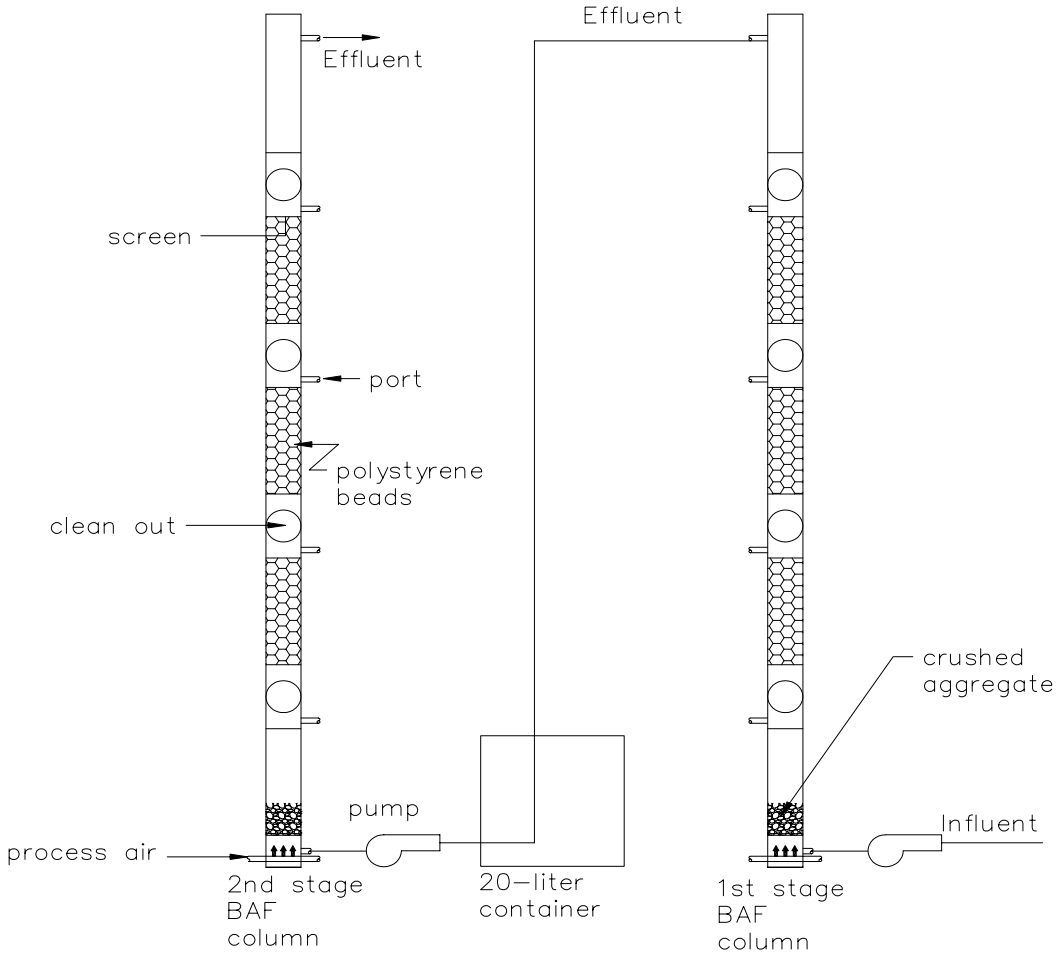


Figure 1.7 A schematic diagram of 1st and 2nd stage BAF reactor.

Backwashing was performed daily for the 1st stage column and every other day for the 2nd column. Thirty to 45 minutes were required to backwash both the 1st and 2nd stage BAF columns. Between 10 to 20 % of the total daily flow of effluent was required to backwash the 1st column and about 2 % to backwash the 2nd column. The backwashing operation was carried out by draining, agitating and flushing the column to remove particulates from the system. Columns were drained through the influent line and the wall ports. During the draining cycle, air supply was turned off. Filling involved closing the influent line and filling the reactor from the top. This was followed by agitation (injection of air through the ports) to disperse the media and dislodge particulates. Once the media was adequately dispersed, the liquid was drained. At this point the outlet valve on the influent line was opened and treated effluent (stored in a 30 gal holding tank) was fed continuously in a downflow manner from the top of the column until most solids were removed. Samples of the backwash liquid were collected and analyzed for TSS concentration and backwash volumes were recorded. Operation of the system began after backwashing by turning airflow and influent pumps on and ensuring that valves to all other ports were closed.

4.4 Kinetic And Stoichiometric Parameter Determination

An attempt was made to determine the maximum specific heterotrophic growth rate $(\hat{\mu}_H)$, the half saturation constant (K_s) and true growth yield (Y). A substrate utilization experiment was performed to determine $\hat{u}_H$ and Ks. Using a 3.5-liter reactor and a food to microorganism ratio of 0.18 (COD basis), which was the same as that maintained in the lab-scale MBBR. The kinetic experimental reactor was inoculated with 350-ml of media from the laboratory scale reactor and comprised 10 % of the kinetic experimental reactor volume.

A synthetic substrate, which consisted of 174.3 mg COD/L acetic acid and 176.3 mg COD/L dextrose, was used. 21.4 mg/L of ammonia chloride and 4.3 mg/L of potassium biphosphate were added as nutrients. The experiment was performed in triplicate. Each reactor was put on a magnetic stir plate and completely mixed with magnetic stirrers. Aeration stones supplied humidified air in each reactor and the DOC of samples was measured at 10 minute intervals. To determine the K_s, it was necessary to monitor the DOC concentration until a stable concentration was reached. However, the experiment was monitored for 3 hours without

achieving a stable concentration. The experiment was repeated with 20 % media by volume of reactor to enable a stable concentration to be achieved in a reasonable period of time. No distinct stable concentration was achieved with this amount of media after 3 hours. For this reason a typical K_s for biofilm in completely mixed systems similar to the MBBR would be used in computations.

An attempt was made to determine true growth yield using fresh support media with no biofilm. A food to microorganism ratio of 120 on a COD basis was used. The food consisted of 66.3 mg/L acetic acid, 67.1 mg/L dextrose and 71.7 mg/L yeast extract for a final initial concentration of 205 mg/L COD. About 50 ml of MLSS from the suspended culture phase of the laboratory-scale MBBR [100 mg/L TSS or approximately (2.0 mg/L COD)] was added to the reactor. Ammonia chloride (51.0 mg/L) and KH_2PO_4 (1.7 mg/L) were added as nutrients. It was very difficult to obtain sufficient biofilm attachment on the media for the duration of the experiment. For this reason, the experiment was repeated without media to obtain a true growth yield for suspended biomass. Samples were collected at 10-minute intervals and analyzed for both total and filtered COD using method 5220 B (APHA, 1998). Yield was determined from the slope of a linear regression line through a plot of biomass COD versus substrate COD concentration.

Both the biofim density and biofilm thickness on MBBR media was determined. Samples of media with biofilm were taken from each of the 3 laboratory-scale reactors (1^{st} stage 40 % media volume, 2^{nd} stage 40 % media volume and 67 % media volume reactors). Biomass density was determined from mass of biomass per media and volume of biomass per media, which were experimentally determined. Biofilm thickness was determined by slicing biofilm-containing media in a manner to enable good magnification and clear pictures of the biofilms to be taken. The sliced media pieces were viewed with an Olympus BH-2 phase contrast microscope supplemented with an Olympus PM-6 35-mm camera. Pictures of 3 different sliced media from each of the reactors were taken to enable an average biofilm thickness to be determined. Pictures were taken with an objective lens providing 4x magnification and with an ocular providing 15x magnification, giving a magnification of 60x. Using the same magnification as the biofilm, a picture was taken of a slide with a metric scale, which enabled

measurements to 20 µm. This enabled direct measurement of the biofilm thickness. Figure 1.8 shows a picture of a typical biofilm attached to a piece of sliced media and the metric scale.

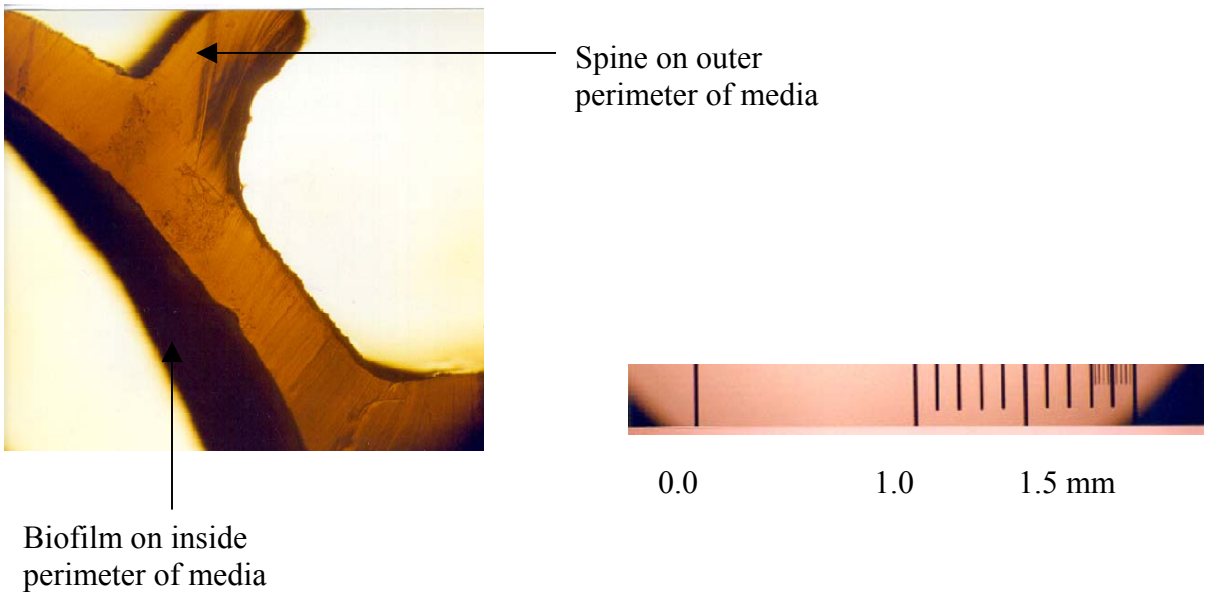


Figure 1.8 (A) Sliced MBBR media with biofilm.(B) Metric scale used to determine biofilm thickness.

(Note that the figures were expanded to varying degrees to facilitate viewing; therefore, the scale cannot be used to estimate the biofilm thickness here).

4.5 **Analytical methods**

Composite samples were obtained daily from the laboratory-scale treatment systems for analysis. The frequency at which the various analytical procedures were performed was as follows: $TBOD_5$, (unfiltered) 4x/week; $CBOD_5$ (unfiltered, nitrification inhibitor, Hach's formula 2533, added), 2x/week; $SBOD_5$ (filtered through a 1.5 µm filter), 2x/week; TSS (defined as that retained by a 1.5 µm), 4x/week; ammonia-N, 2x/week. Other analyses included DOC, nitrate-N of 1st and 2nd stage BAF effluent and 2-stage MBBR effluent, pH, dissolved oxygen and COD during the kinetic experiment.

Influent wastewater and effluent from the MBBR and BAF were analyzed for total carbonaceous and soluble BOD_5. Samples collected for BOD_5 were analyzed within 24 hours of collection and kept in a 4^oC refrigerator prior to analysis. Procedures recommended in Standard Methods (APHA, 1998) were used. Nitrification inhibitor (Hach's formula 2533) was added to the $CBOD_5$ samples (0.16 g per 300 ml of sample). Samples were filtered using a 55 mm diameter, 1.5µm-glass microfibre filter (Whatman Inc., USA) for soluble BOD determination. Estimation of sample dilutions for BOD analysis was accomplished by using a relationship established between DOC and $TBOD_5$ and recommendations by Metcalf and Eddy, Inc. (1991) Each BODs ($TBOD_5$, $CBOD_5$, $SBOD_5$) was performed in triplicate and the average is presented. Some results from the BOD test when final DO readings were below 1.0 mg O_2/L were discarded as recommended in Standard Methods (APHA,1998). BOD test results with a DO change less than 2 mg O_2/L but greater than 1.0 mg O_2/L as a result of over dilution were used in data analysis. An oxygen probe (YSI Model 58, Yellow Spring Instrument Co.) was used to determine dissolved oxygen concentrations.Both influent wastewater and effluent samples contained sufficient bioparticles so that seeding was not necessary. However, BAF effluent samples had very low TSS; therefore, they were seeded with VT sewage (influent sample) to ensure that adequate microorganisms were present to degrade BOD in the sample. TSS and VSS, MLSS and MLVSS were determined using method 2540 D in APHA, 1999. A 1.5µm filter paper was used to filter the samples. Samples were oven dried at 105^oC. VSS determination involved burning the weighed TSS sample in a muffle furnace at 500^oC. Ammonia-N was determined by the titrimetric method (4500-NH_3 C) as specified in Standard Methods (APHA, 1999). Nitrate-N concentration was determined by the ion chromatography method (4500 NO_3-N C).

4.6 **Development of the MBBR model**

Results of the tracer study performed with the single stage 40 % media volume MBBR showed that at least 3 CSTRs in series were required to best fit the data collected during the study. For this reason the single stage 40 % media volume reactor was modeled as 3 CSTRs in series and analyzed with the pseudoanalytical model. The flow diagram of the 3 CSTRs in series is shown in Figure 1.9

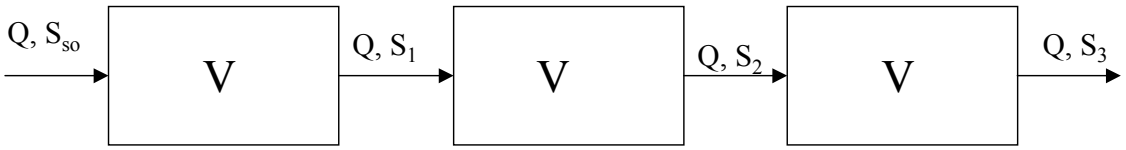


Figure 1.9 Single stage, 40 % media volume MBBR modeled as 3 CSTRs in series.

For a single stage, 40 % media volume with reactor volume V_t, each of the tanks in series represents one third of the reactor volume. Steady state mass balance implies that:

$$S_i = S_{so} - \frac{J_s a V}{Q} - \frac{q_H X_{B,Hb} V}{Q}, \text{ where } a = \frac{A_s}{V} \qquad \text{(Equation 1.15)}$$

where S_i represents S_1,S_2,S_3 and S_b. Equation (1.15) would be used to determine the reactor volume required to achieve the desired effluent quality. The pseudoanalytical model predicts substrate flux into biofilms in each of the tanks based on biofilm kinetics, stoichiometry and diffusion. The model is presented in chapter 3.3. Experimentally determined kinetic and stoichiometric parameters together with typical diffusion and kinetic parameters presented in Table 1.4 were used as inputs to the model and allowed substrate flux and influent substrate into each of the tanks to be determined. To do this, S_b (S_3) in the last tank was held constant at 30 mg $TBOD_5$/L (approximately equal to 50 mg COD/L) and the influent concentration to the tank calculated. The process was repeated moving up the tanks in series using the calculated influent concentrations as effluent (S_2) in the second tank and the influent concentration to the tank calculated. The iteration process continued, until the predicted value for S_{so} to the first tank equaled 257 mg $TBOD_5$/L (approximately equal to 440 mg COD/L),

the actual feed concentration. During this process q was calculated and used in equation 1.15 for each tank using the Monod expression (Grady et al., 1999).

The model enabled the MBBR system to be validated by comparing predicted and measured concentrations. The validated model was used to determine the volume required for the MBBR tank.

CHAPTER 5.0 RESULTS AND DISCUSSION

5.1 Virginia Tech Sewage Characterization

5.1.1 Initial Wastewater Quality Screening

The results of the wastewater characterization study for both the recess and school periods are summarized in Tables 1.6 and 1.7, respectively.

Table 1.6 Wastewater characteristics during recess periods.

Periods	VT sewage			
	$TBOD_5$ (mg/L)	$CBOD_5$ (mg/L)	TSS (mg/L)	VSS (mg/L)
20-Nov. 2000	209	155	337	301
21-Nov. 2000	203	201	220	201
13-Dec.2000	164	141	-	-
19-Dec.2000	168	146	-	-
Average ($\overline{X}$)	**186**	**161**	**279**	**251**
Std. Dev. (SD)	**23**	**28**	**83**	**71**
Count (N)	**4**	**4**	**2**	**2**

Table 1.7 Wastewater characteristics during school periods.

Periods School sessions	VT sewage			
	$TBOD_5$ (mg/L)	$CBOD_5$ (mg/L)	TSS (mg/L)	VSS (mg/L)
27-Nov. 2000	-	-	335	298
28-Nov. 2000	-	-	383	317
30-Nov. 2000	215	181	277	257
1-Dec. 2000	151	141	242	222
9-Dec. 2000	307	263	200	199
10-Dec. 2000	273	249	223	211
11-Dec. 2000	277	258	220	206
12-Dec. 2000	237	214	-	-
Average ($\overline{X}$)	**243**	**218**	**269**	**244**
Std. Dev. (SD)	**55**	**49**	**68**	**47**
Count (N)	**6**	**6**	**7**	**7**

The combined average BOD_5 in VT sewage during school recess periods and school sessions was 215 mg BOD_5/L (SD = 53, N = 10). This was less than the target design $TBOD_5$ of 300 mg/L required for FPWTS. Tables 1.6 and 1.7 show TSS trends in the VT sewage. TSS concentrations for periods when school was in session and on break was 265 mg/L (SD=64, N=10) and exceeded target design concentration of 133 mg/L.

The results from the wastewater quality screening indicate that soluble BOD_5 had to be added to the sewage and suspended solids had to be removed. Dissolvable readily biodegradable $SBOD_5$ supplement was added to the WTSs daily, as outlined in Materials and Methods. Additionally, partial settling of suspended solids had to be achieved to meet the target influent TSS goal. Figure 1.10 shows the results of a suspended solids settling experiment performed using VT sewage. Approximately 70% of the suspended solids settled within 4 hours of the experiment. These results were used to modify the TSS concentration in an effort to achieve the target influent TSS concentration, as outlined in Materials and Methods.

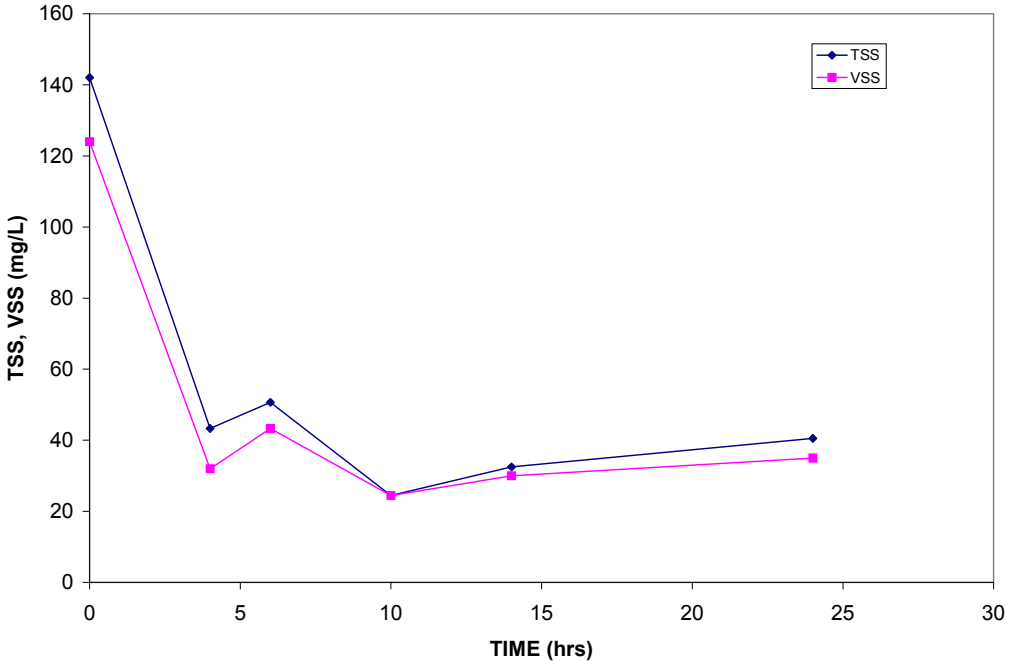


Figure 1.10 Plot of suspended solids versus time in settling experiment on VT sewage.

The DOC of the VT sewage was also measured during the characterization study. A plot of $TBOD_5$, $CBOD_5$ and DOC trends in VT sewage is presented in Figure 1.11, and shows that DOC tracked BOD_5 trends well. A good correlation (r^2=0.96) was obtained between the sewage $TBOD_5$ and DOC, as shown in Figure 1.12. Since the DOC analysis was easy and rapid, DOC values of the feed and effluent streams were monitored daily to estimate influent and effluent $TBOD_5$. This information was used to determine how much supplement $SBOD_5$ to add, and to determine the dilution factors required for BOD_5 assays. Figure 1.12 also shows that approximately 11 mg/L of the DOC was not readily biodegradable.

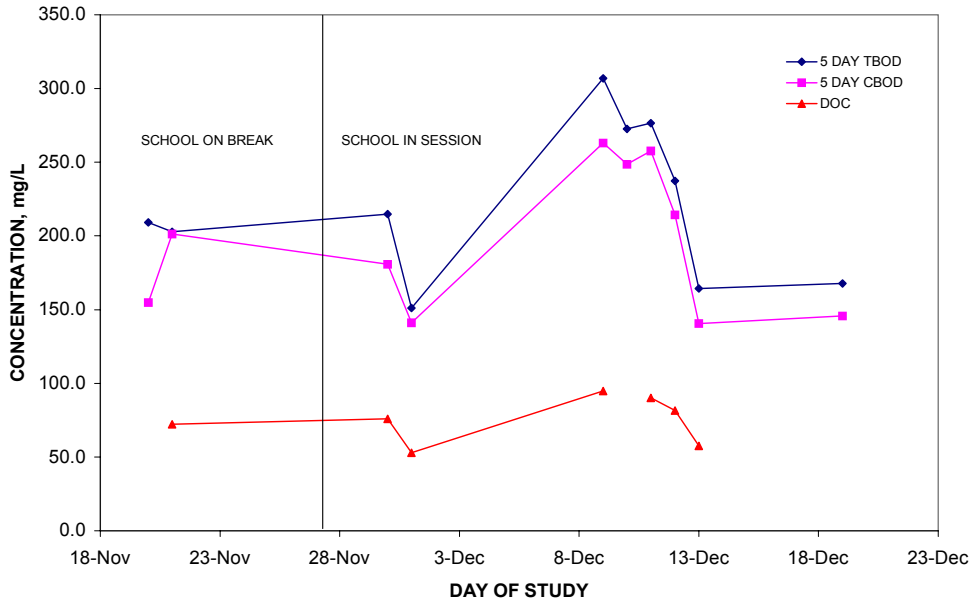


Figure 1.11 $TBOD_5$, $CBOD_5$ and DOC trends in VT sewage.

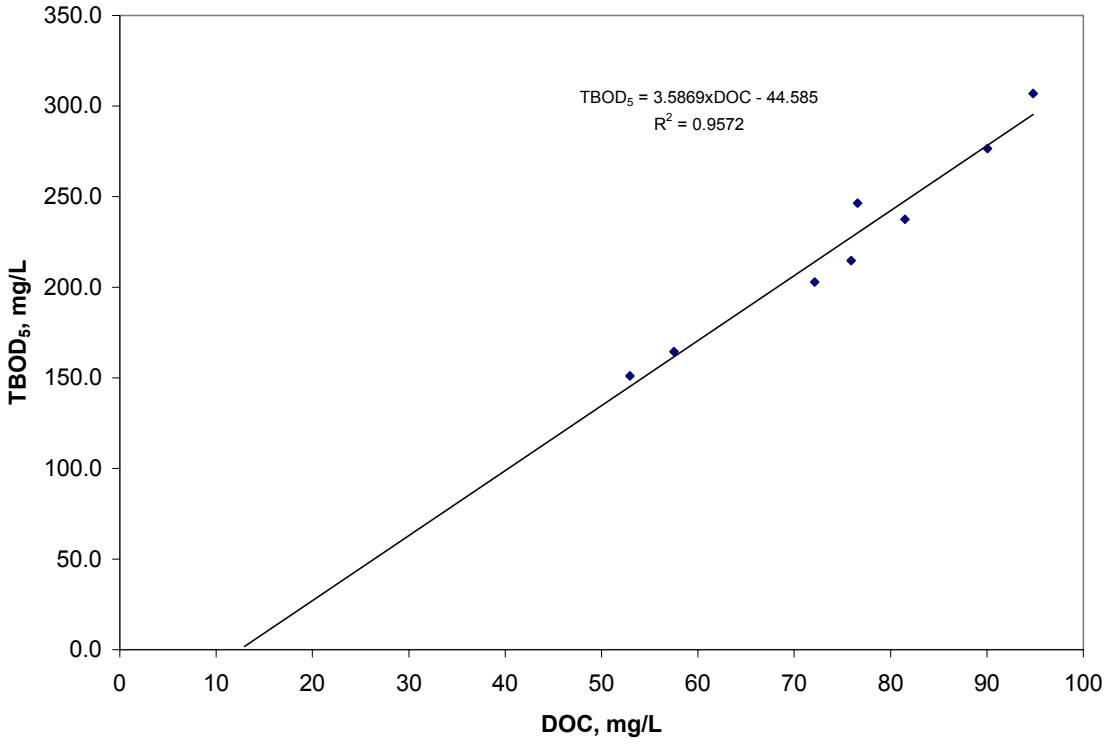


Figure 1.12 Influent $TBOD_5$ and DOC correlation for VT sewage.

5.1.2 Influent Wastewater Characteristics

The influent wastewater characteristics for the duration of the reactor study are summarized in Table 1.8. The results reflect the impact of supplementing $SBOD_5$ with a synthetic additive and implementing partial TSS settling, as described previously. The average $TBOD_5$ obtained was 257 mg/L (300 mg/L targeted) and average TSS was 129 mg/L (133 mg/L targeted). The influent wastewater parameters compared very well with the targeted values defined in the Phase I FPWTS UTD report. (UTD, Inc.,2000).

Table 1.8 Influent wastewater characteristics obtained during the reactor study.

Parameter		Influent
$TBOD_5$, mg/L	Average	257
	Std. Dev. (SD)	63
	Count (N)	44
$CBOD_5$, mg/L	Average	225
	Std. Dev. (SD)	61
	Count (N)	27
$SBOD_5$, mg/L	Average	189
	Std. Dev. (SD)	41
	Count (N)	26
TSS, mg/L	Average	129
	Std. Dev. (SD)	76
	Count (N)	67
VSS, mg/L	Average	122
	Std. Dev. (SD)	67
	Count (N)	61
NH_3-N, mg/L	Average	24.3
	Std. Dev. (SD)	5.92
	Count (N)	33

5.2 **Stimulus - Response (Tracer) Study**

Figure 1.13 shows an F-curve of conductance versus time during the tracer study and the corresponding E-curves, which were derived from the F-curves. Values calculated for the hydraulic residence time and the required number of tanks (reactors) in series are given in Table 1.9 for the three experiments.

Table 1.9 Summary of results from tracer study on a single MBBR with 40 % media volume

Experiment	Empty bed HRT (min.)	Theoretical HRT based on 40 % media volume (min)	Mean residence time, $\bar{t}$, with 40 % media volume (min.)	Number of tanks-in-series (N)
1	90	56	53	2
2	90	56	49	3
3	90	56	52	3

The average measured HRT (based on 40 % media volume) was 51 min and approximately 3 tanks in series of equal volumes were required to best fit the data collected during the study. The results also indicate that the volume of the reactor had to be increased or the liquid flow rate reduced to achieve an effective treatment HRT of 90 min.

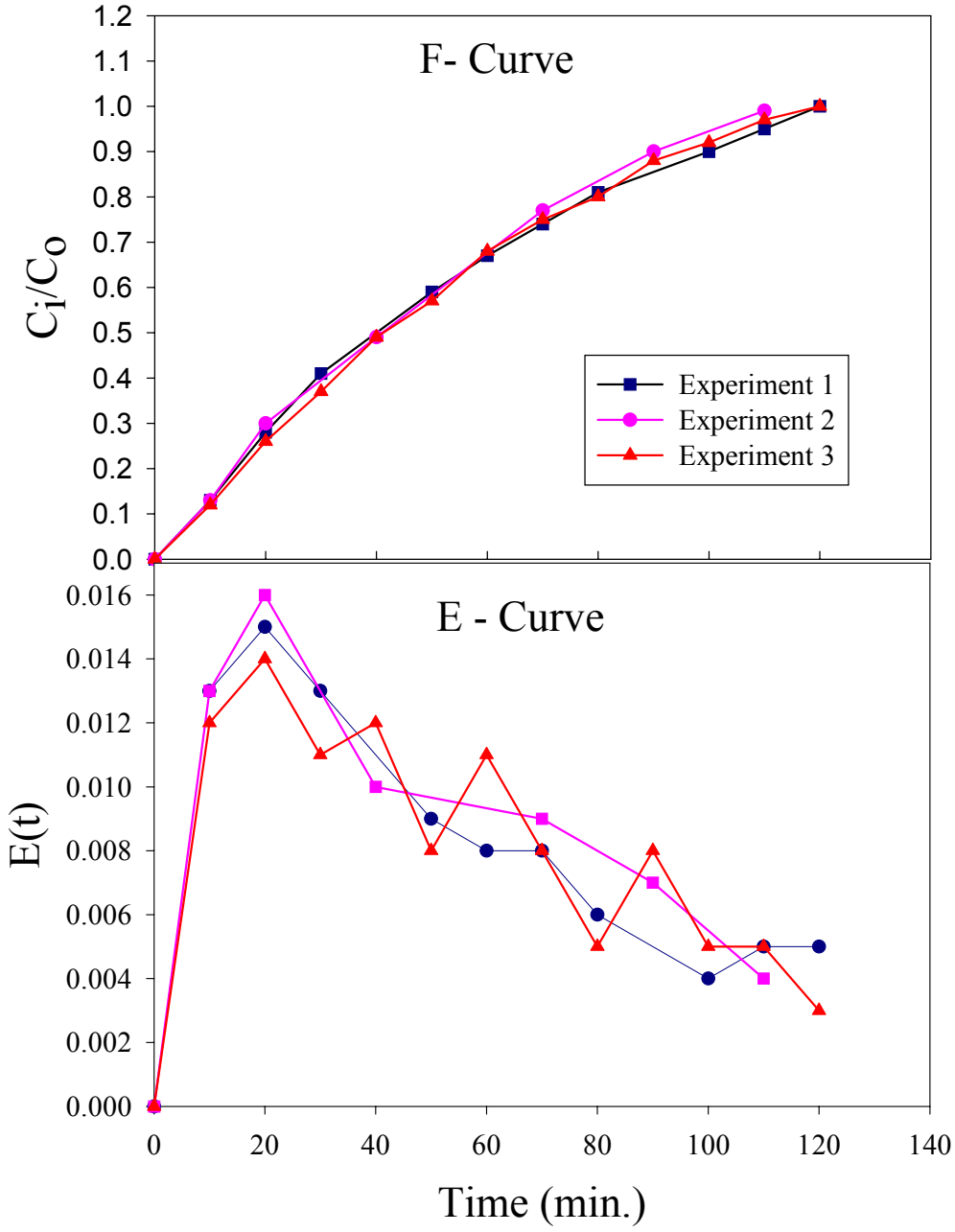


Figure 1.13 F-Curve measured during replicate tracer experiments for a single MBBR with 40 % media volume, E-Curves derived from F- Curves.

5.3 MBBR performance

5.3.1 BOD_5 Removal

A plot of total and soluble BOD_5 versus time is presented in Figure 1.14 and shows that the system did not meet the effluent $TBOD_5$ target during most of the study.

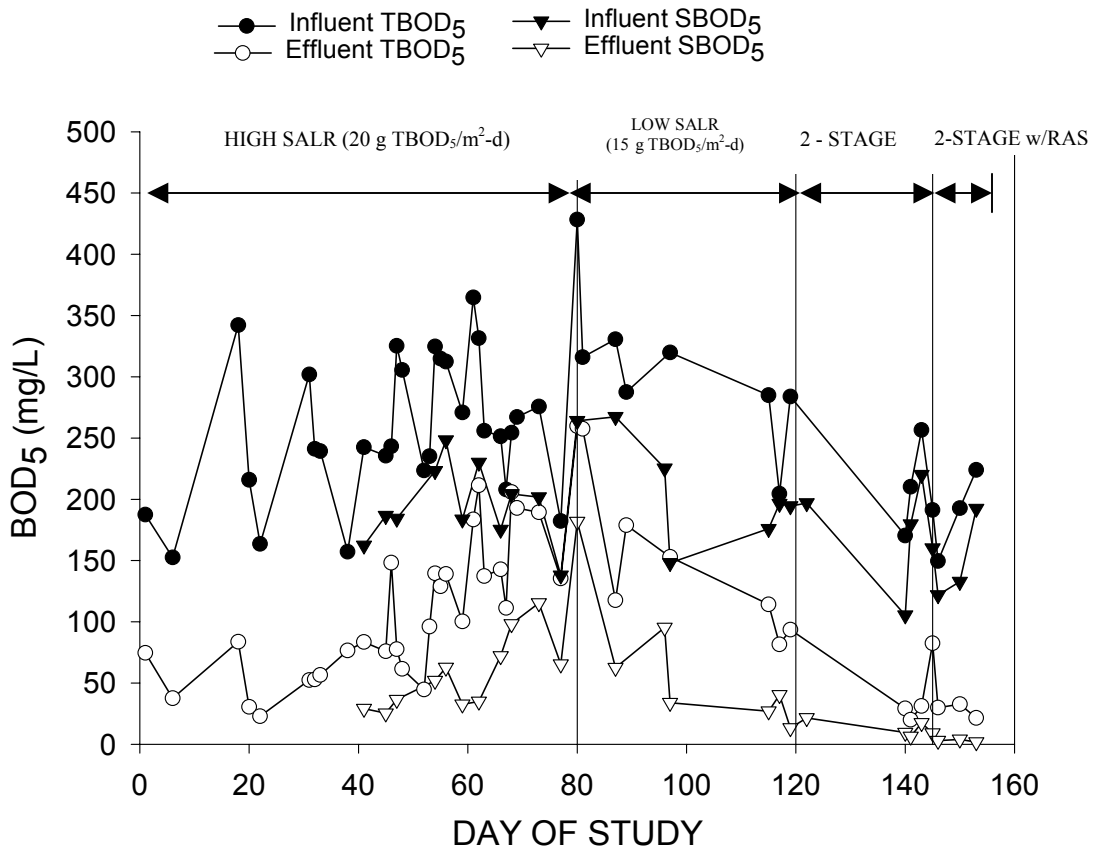


Figure 1.14 Influent and effluent BOD_5 trends in MBBR with 40 % media volume.

Effluent $TBOD_5$ deteriorated from the beginning of the experiment to day 80 when surface area loading rate (SALR) was high (averaging 20 g $TBOD_5/m^2$-day, SD = 5.8, N = 30). The deteriorating trend in the effluent quality reversed after day 80 when the SALR was reduced to an average of 15 g $TBOD_5/m^2$-day (SD = 2.2, N = 7). The average effluent $TBOD_5$ concentrations for the single stage MBBR with 40 % media volume were 112 mg/L (SD = 64, N = 30) before day 80 and 142 mg/L (SD = 61, N = 7) after day 80 but before 2-stage operation of the MBBR began. Reducing the SALR did not improve the average $TBOD_5$

removal efficiency since steady state was not achieved before the reactor configuration was changed to a two-staged system on day 120. As a result, the effluent $TBOD_5$ was well above the effluent target and it appeared that the single-stage system was unlikely to meet the design criteria.

$TBOD_5$ reductions continued to improve during the period when the system was operated with return of settled sludge to the 1st and 2nd stage reactors. Average $TBOD_5$ and $SBOD_5$ concentrations during the period were 42 mg/L (SD = 28, N = 4) and 5 mg/L (SD = 3, N = 4) respectively. Further studies would be required to study the effect of sludge return on the performance of the system since the system was operated in this mode for only 9 days and steady state conditions may not have been achieved.

Although the MBBR performed poorly with respect to $TBOD_5$ removal, effluent $SBOD_5$ concentrations were relatively low with average concentrations of 67 mg/L (SD = 45, N = 12) before day 80, 45 mg/L (SD = 29, N = 6) after day 80 and before day 120, and 13 mg/L (SD= 6, N= 5) after day 120 when the system configuration was changed to 2-stage. The trend in the $SBOD_5$ was similar to that of the $TBOD_5$, deteriorating during the first 80 days until the loading rate was reduced. The shift from a single reactor to a two-stage system had a small improvement on $SBOD_5$. The poor performance of the system before day 80 was possibly due to the high SALR or insufficient media provided (40 % reactor volume) (Johnson, 2001.). The improved performance of the system after day 80 when the SALR was reduced confirmed the suspected link between the systems low efficiency and the higher SALR.

The MBBR showed improved $TBOD_5$ removal efficiency when it was operated as a single reactor with 67 % media volume and no sludge recycle (Figure 1.15), relative to 40 % media volume (Figure 1.14), the TBOD improvement averaged 60 percent and yielded an average effluent $TBOD_5$ concentration of 57 mg/L (SD = 27, N = 6). Therefore, operating with a higher media volume may have allowed the MBBR system to meet effluent criteria for the FPWTS.A summary plot of $TBOD_5$ mass removal rate versus loading rate is given in Figure 1.16, and shows that performance improved at lower SALRs and higher media volumes.

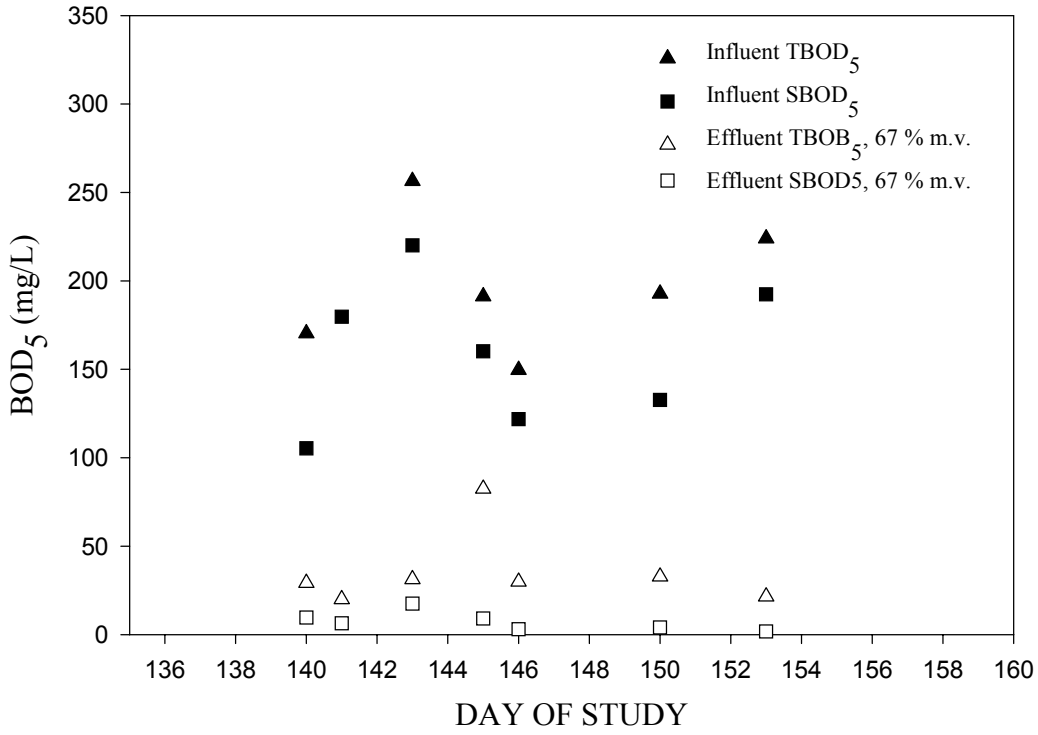


Figure 1.15 Plot of BOD_5 trends obtained with 67 % media volume reactors.

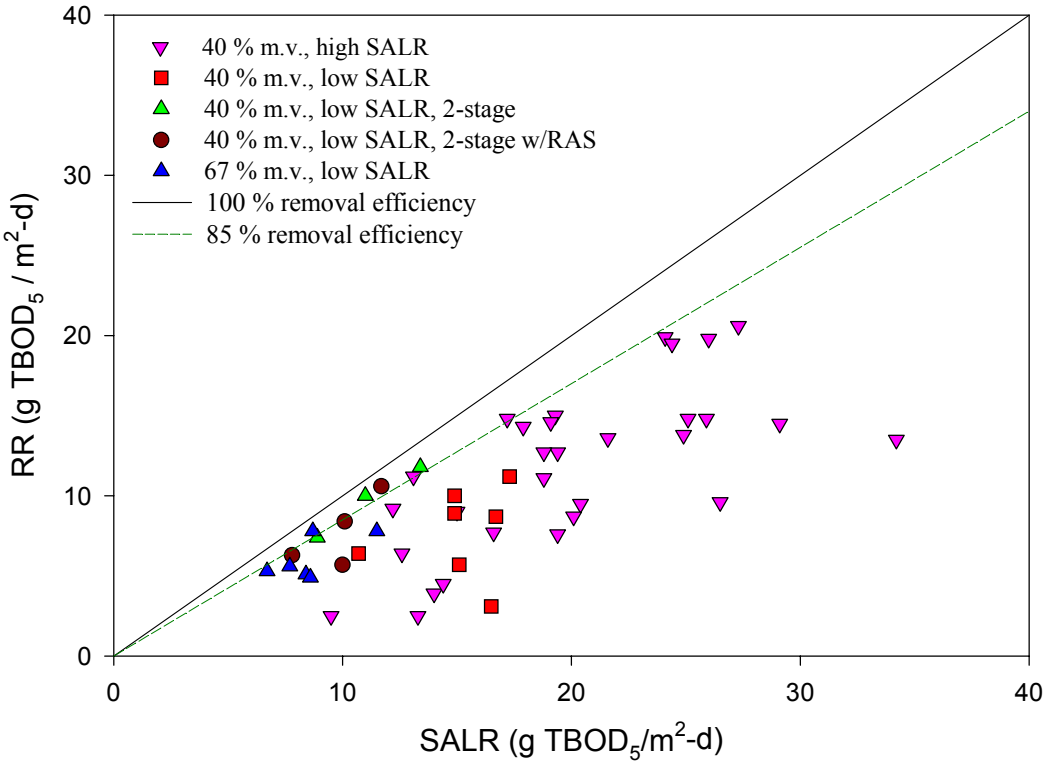


Figure 1.16 Plot of BOD_5 loading rate versus removal rate for 40 % media volume.

5.3.2 TSS Removal

Net TSS removal in the MBBR system was poor for all 5 of the system configurations monitored. The trends in the MBBR influent and effluent TSS concentrations are shown in Figure 1.17. The average effluent TSS concentration for the entire period of study was 145 mg/L (SD = 109, N = 67).

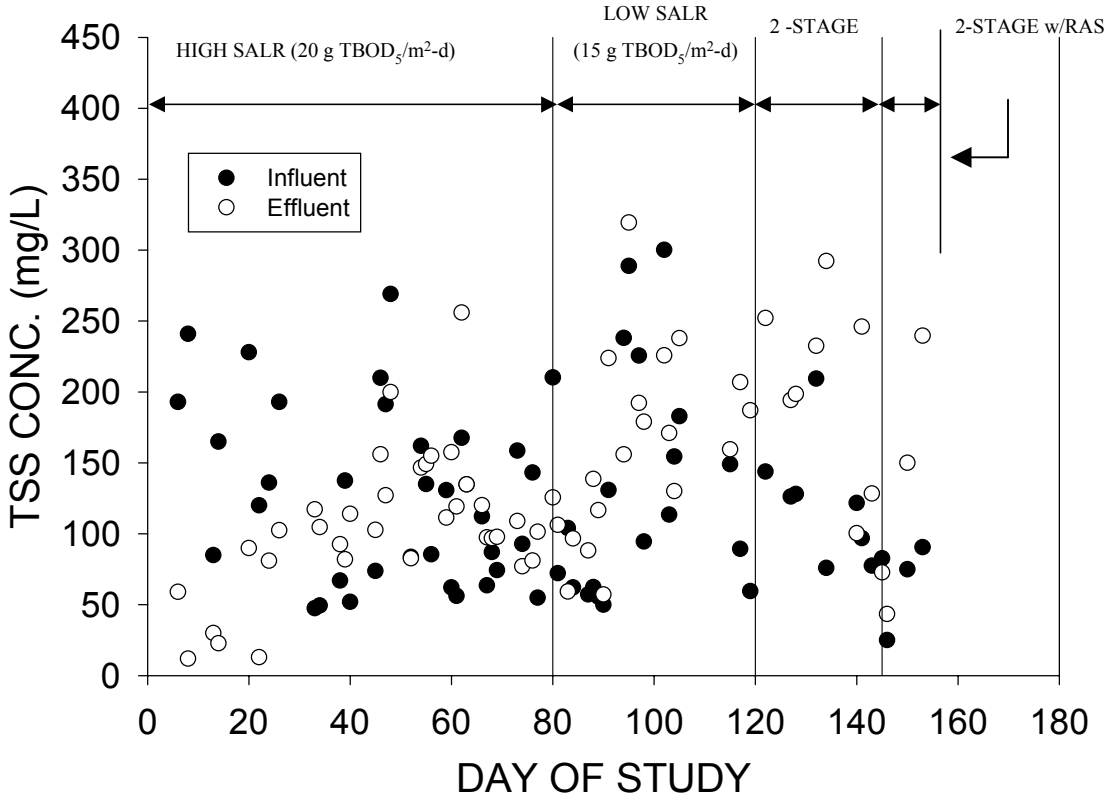


Figure 1.17 Influent and effluent TSS concentration trends with 40 % media volume.

During the first 24 days of the study, over 50 % average TSS removal was observed. There was no defined removal trend in the TSS after this period, and effluent TSS concentrations were generally higher than the influent concentrations. This change in TSS removal efficiency over time is believed to be due to two factors. First, a change in sampling protocol from a grab-sampling mode to a composite sampling mode may have influenced the effluent TSS. A second factor believed to contribute to the high effluent TSS is the low residence

time of the suspended solids in the MBBR. Sludge age (solids retention time, SRT) has significant effect on settling properties of suspended solids in a bioreactors. High SRT results in decreased sludge volume index (SVI) (good settling of solids) in activated sludge systems (Phillips, 1998). SRT calculated for the period of the study was in the range of 2 to 3 hours. In order to achieve acceptable settling quality, the SRT for suspended growth reactors should be at least 3 days (Grady et al., 1999). The SRT of the system increased to 1.5 days during the period when the system was operated with return sludge to the reactors. Due to time constraints, this configuration was operated for only 9 days, which was not long enough to achieve steady state. Despite implementing sludge return, high effluent TSS concentrations were recorded in the composite samples over the brief experimental period. However, a grab sample collected from the MBBR effluent holding tank (which functioned as a supplemental clarifier) was found to have an effluent TSS of 16 mg/L. This observation suggests that addition of a clarifier downstream of the MBBR with sufficient retention time and sludge thickening for recycle is likely to yield effluent quality with respect to TSS.

High effluent TSS concentrations were reported in a study of a high rate MBBR system used to pretreat chemical wastewater (Rusten et al. 1999). It is not uncommon to find a solid separation unit down stream of an MBBR plant, as shown in the flow schemes in the literature review. When phosphorus removal is a requirement in the wastewater treatment, chemical addition to the effluent reduces TSS levels to very low concentrations through coagulation (Odegaard et al., 1994).

5.3.3 Ammonia Oxidation

Although nitrification was not a goal of this study, ammonia concentrations in the influent and effluent waste streams were monitored. Results are presented in Figure 1.18.
An average of 30 % of influent ammonia concentration was removed by the MBBR during the entire period of the study. The average effluent ammonia-N concentration prior to day 120 was 18.75 mg/L (SD = 4.79, N = 27), which corresponded with an average removal of 24 %. Lack of nitrification is very likely due in part to high organic loading to ammonia-N loading in the influent sewage, which exceeded an influent BOD_5:$NH4^+$-N of 5.

Additionally, limiting dissolved oxygen concentrations in the reactor may have prevented nitrification. DO remained between 1.5 and 2 mg O_2/L in the MBBR throughout the operation of the single stage MBBR Odegaard et al., (1994) reported that DO levels above 2-3 mg O_2/L were required for nitrification to take place in the MBBR.

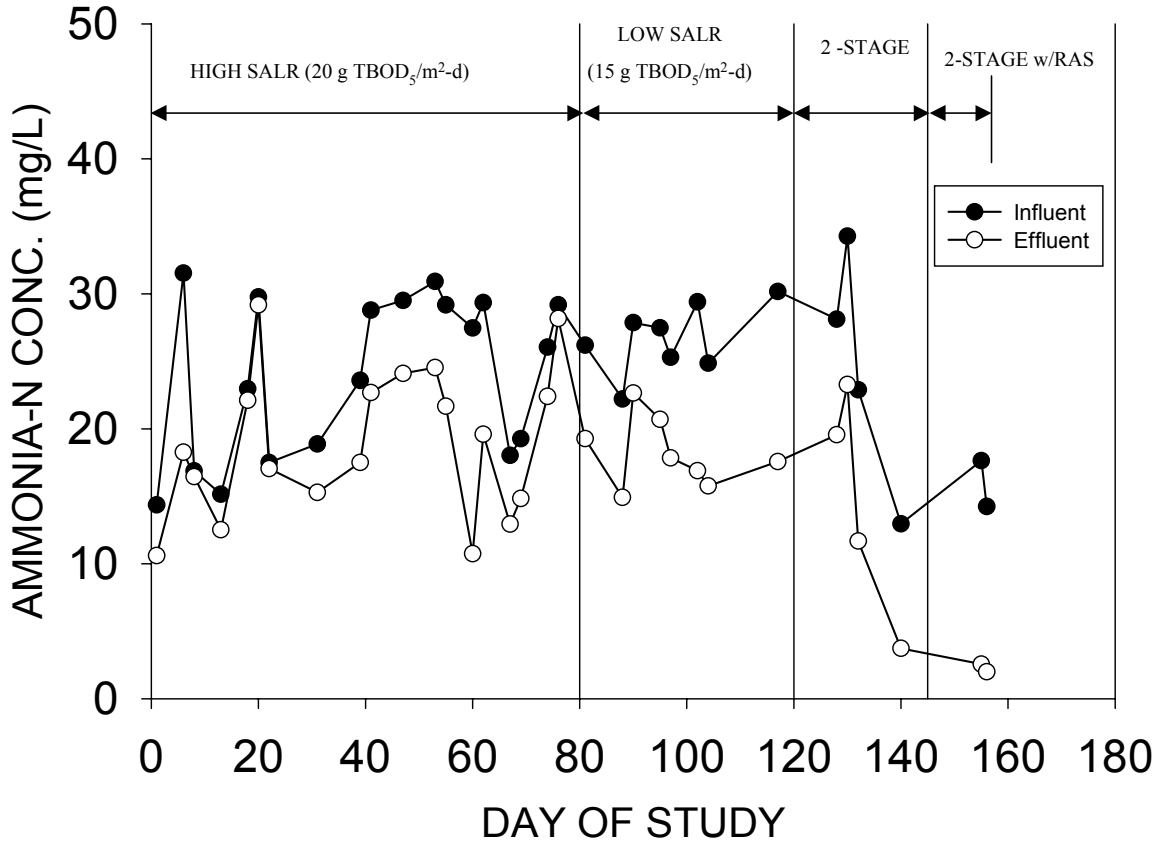


Figure 1.18 Ammonia trends in MBBR with 40 % media volume.

High ammonia removal levels were observed when the system was staged, after day 120 as shown in Figure 1.18. This could be due to a low influent BOD_5:NH_4^+-N to the second stage, which would support nitrification in the second stage. On one occasion when ammonia reduction was 71 %, the nitrate concentration measured in the effluent was 0.21 mg N/L,

confirming partial nitrification and possible denitrification in the system. The concentration of nitrate-N in a subsequent test was zero. Further studies may be required in the future to examine the potential for simultaneous nitrification and denitrification in staged MBBR systems.

5.3.4 MBBR Performance Summary

Despite its potential for improved performance, the MBBR did not demonstrate acceptable treatment capacity to meet the effluent guidelines for the FPWTS. Due to limited time, it was not possible to adequately evaluate the three reactor configuration modifications that were most likely to improve performance (increased media volume, sludge recycle to increase the SRT of the suspended biomass and increased clarifier surface area). The poor effluent quality was heavily linked to the high effluent TSS, which also increased the $TBOD_5$. The $SBOD_5$ data obtained during two-stage operation averaged below 13 mg/L, suggesting that treatment strategies that are effective at reducing effluent TSS would enable the MBBR system to meet the target effluent criteria.

Application of chemicals by coagulation and flocculation would be an option to remove suspended solids in the MBBR effluent stream if settling of biosolids in the systems was poor.

Odegaard et al., 1994, reported that a typical design-loading rate of 25 g BOD_7/m^2-day (equivalent to 21.4 g BOD_5/m^2-day, assuming 60 mg BOD_5 equals to 70 mg BOD_7 as reported by Rusten et al. (1997)) with 67 % media volume, achieved between 75 and 80 % reduction in BOD_7. For secondary treatment only, the process is normally designed for a volumetric loading of 12 to15 g BOD_7/ m^2-d (10 to 13 g BOD_5/m^2-d) at 67 % carrier filling and 15^oC. This implies that SALRs less than 10 g $TBOD_5/m^2$-day at 40 % media volume were required for efficient performance of the system.

5.4 Biological Aerated Filters

5.4.1 BOD_5 removal

The average BOD_5 mass loading rate to the BAF system was 3492 g/m^3-d (SD = 767, N = 39) and achieved efficient organic carbon removal (Figure 1.19). The BAF effluent $TBOD_5$ and $CBOD_5$ concentrations averaged 16 mg/L (SD = 8, N = 39) and 11 mg/L (SD = 8, N = 25), respectively, during the entire period of the study. These results show that the BAF performed well within the target effluent criteria given for organic matter (UTD Inc, 2000). Effluent $SBOD_5$ concentrations in Appendix A3 averaged 15 (SD = 8, N = 10). Effluent $TBOD_5$ and $SBOD_5$ values were similar; indicating the efficiency of the system in filtering suspended solids.

First stage effluent BOD_5 concentrations showed that over 80 % of the organic carbon removal occurred in the first BAF column. Thus, the second column served as a polishing unit for BOD_5 and led to excellent effluent BOD_5 quality. Without the second column, the average $TBOD_5$ concentration to be expected from the effluent of the single column was 36 mg/L (SD= 22, N=9). This average did not meet target effluent criteria and suggests that some additional media beyond that contained in the first stage would be needed.

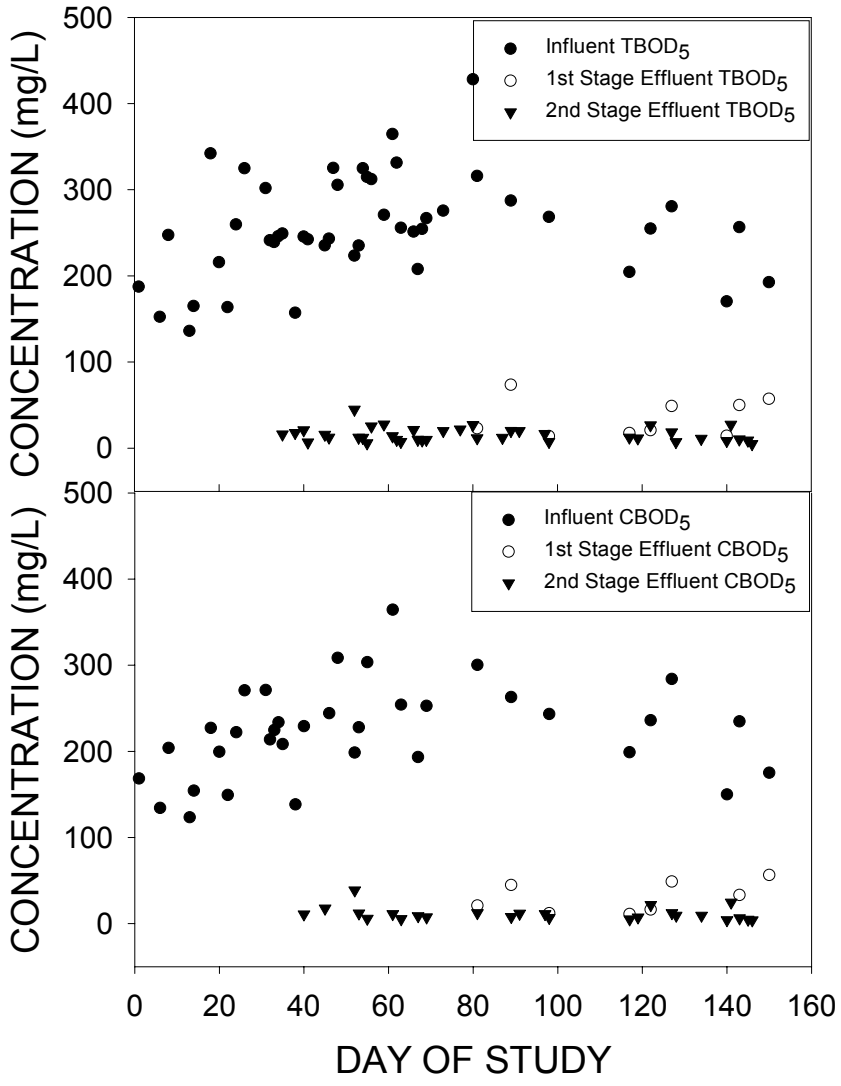


Figure 1.19 BOD_5 trends in 1st and 2nd stage BAF columns.

5.4.2 TSS Removal

High TSS removal efficiency was observed in the BAF with effluent concentrations less 30 mg/L on most occasions, as shown in Figure 1 20. The average BAF effluent TSS over the course of the study was 18 mg/L (SD = 22, N = 45). The columns required daily backwashing to remove solids (both filtered solids and biomass) entrapped in the BAF media and to reduce head losses as a result of clogging in the media. High backwash TSS concentrations with an average of 6559 mg/L (SD = 3606, N = 20) were measured between day 56 and day 81 of the study, and 1860 mg/L (SD = 1443, N = 7) between day 85 and day 109 of the study. Evaluation of the BAF system based on measured BOD_5 loading, an assumed yield typical of BAF systems achieving organic carbon removal and measured backwash volumes, the latter backwash concentration was more consistent with the calculated concentration. The high backwash TSS concentration before day 81 was probably due to high BOD_5 and TSS loadings to the system. These concentrations demonstrate the high efficiency with which BAFs remove suspended solids. The low TSS concentration in the effluent could thus be attributed primarily to solids filtration in the BAF.

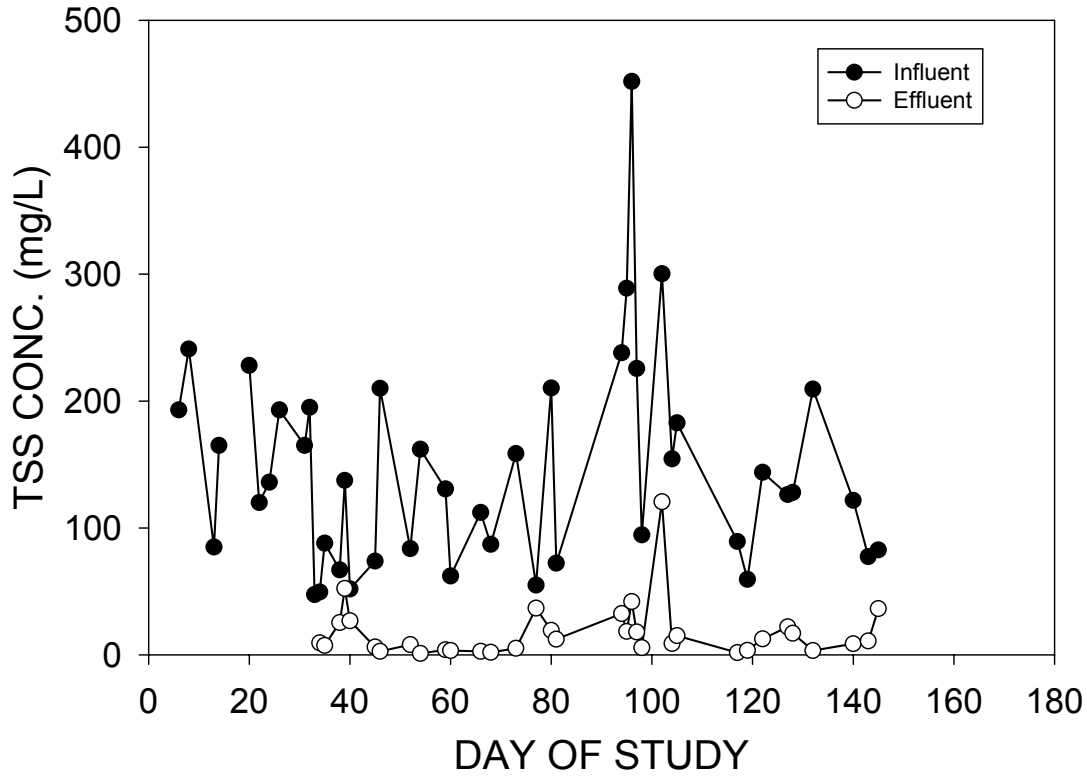


Figure 1.20 Influent and 2nd stage effluent TSS concentrations in BAF.

5.4.3 Ammonia Oxidation

Ammonia utilization in the BAF columns is illustrated in Figure 1 21 for influent, 1st and 2nd stage effluent concentrations. Ammonia removal was generally above 80 % with periodic deviations from the trend on a few occasions. Analysis of 1st stage effluent revealed that nitrification mainly occurred in the 2nd column. This occurred because much of the organic carbon in the waste stream was removed in the 1st column, reducing the $TBOD_5$: NH_4^+-N ratio to the second column to be within a range acceptable to nitrification. This trend is consistent with what has been observed previously for 2-stage BAF systems (Gilmore et al., 1999). DO concentrations measured in the effluent of the BAF were above 6.0 mg O_2/L, suggesting that favorable conditions existed for ammonia oxidation. Traces of nitrates (0.13 mg N/L and 0.3 mg N/L) were measured in the 1st stage effluent, indicating occurrence of partial nitrification in the first column.

The average nitrate-N concentration in the 2nd column was 4.21 mg NO_3-N/L (SD = 1.24, N = 4). The significant difference of nitrate-N concentration in the first and second column effluents indicates that nitrification mainly occurred in the latter column.

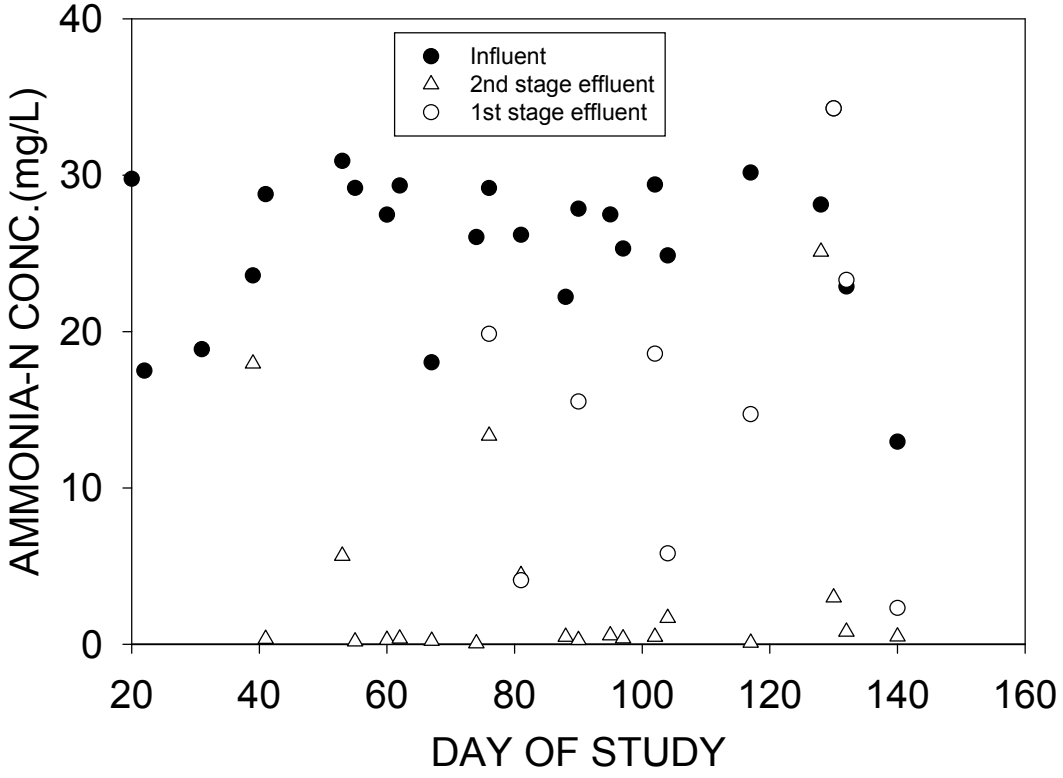


Figure 1.21 Influent, 1st and 2nd stage effluent ammonia concentration in BAF

5.4.4 BAF Backwashing

Backwash volumes for the 2nd stage BAF constituted between 10 and 20 % of daily total effluent flow for the first column and 2% for the second column. These figures compare very well with backwash volumes reported in full-scale BAF plants by others (Dillon and Thomas, 1990; Canler and Perret, 1994 ; Wheale and Cooper-Smith, 1995). Between 12 and 35% of total daily flow were reported for BAFs achieving secondary treatment of BOD_5 and 2 % in tertiary treatment for nitrification.

The backwash operation during the study could be said to be effective since the reactor efficiency was consistently high and no solids break through were observed in the effluent during the period of the study, except on day 102 when a high TSS concentration of 121 mg/L was observed. The influent TSS on that occasion was 298 mg/L, indicating a high

incidence of TSS concentration in the influent and was more than two times that desired in the influent stream. The TSS reduction on that day was 60 %.

In order to compare the suspended solids removal efficiencies of the two systems, a mass balance for TSS was determined for the BAF and MBBR. Suspended solids from the MBBR included those in the effluent and daily wastage stream of settled solids from the quiescent zone at the end of the reactor. The BAF suspended solids were that in the effluent and backwash streams. The average mass of solids generated daily by the MBBR and BAF were 32 g/d (SD = 26, N = 11) and 27 g/d (SD = 29, N = 6), respectively. Therefore, approximately 16 percent more solids came out of the MBBR than the BAF. This result indicates that neither system had a significant advantage over the other in terms of suspended solids removal.

5.4.5 BAF Performance Summary

The 2-stage BAF produced effluent quality, which met the desired effluent criteria of FPWTS. The average effluent $TBOD_5$ achieved in the first column was 36 mg/L, which was slightly less than the target effluent $TBOD_5$. This implies that a single stage column with media volume more than that provided in the first column but less than the total media volume provided for the two columns can achieve the targeted effluent $TBOD_5$. The specific area of the media used in the laboratory-scale BAF was estimated to be 2500 m^2/m^3 and was relatively high compared to typical values (between 1000 - 1500 m^2/m^3) normally reported for BAFs. The high specific area of the media contributed to the high performance of the BAF system.

Backwashing was a major draw back to the performance of the BAF. The backwash operation required between 10 to 20 % of the total daily effluent flow, air supply to disperse the media, a system to manage the high TSS in the backwash stream and between 30 to 45 minute duration for the operation. Effective backwashing is an essential requirement for the effective performance of the BAF and the cost of backwashing can be a substantial component of the operation and maintenance cost of the BAF.

The $TBOD_5$ removal efficiency of the 2-stage BAF as shown in Figure 1 22 was above 85 % and should be adopted for the FPWTS.

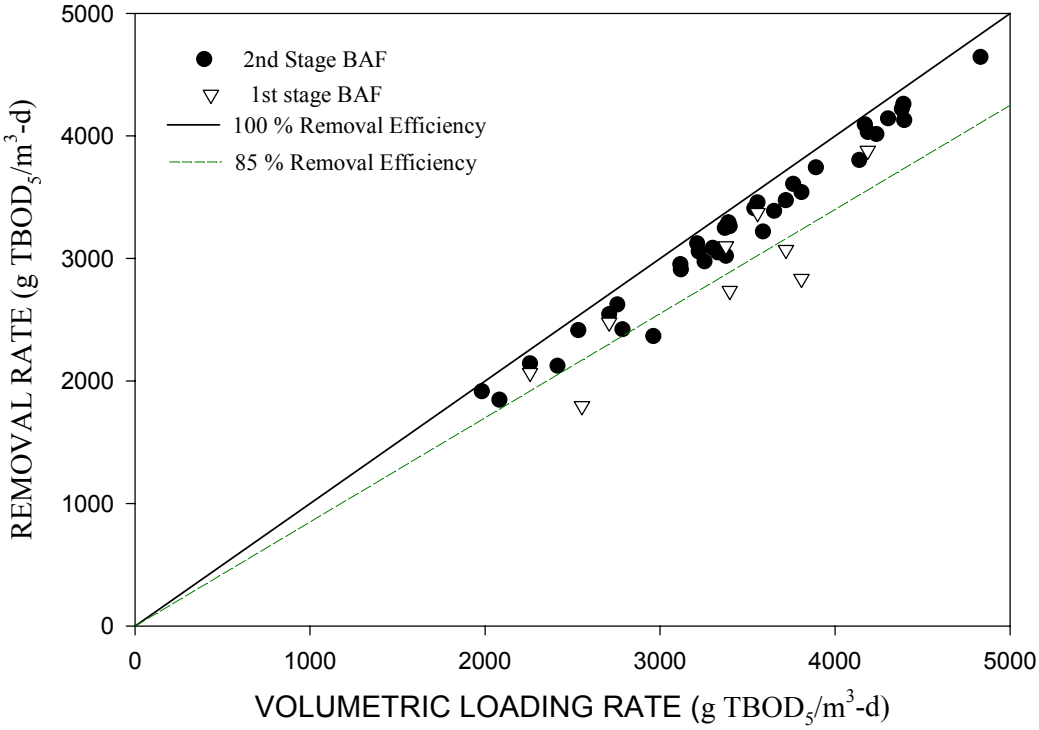


Figure 1.22 Plot of BOD_5 mass loading rate versus removal rate for 1st and 2nd column BAFs.

5.5 MBBR Kinetic and Yield Study and Model

The CSTRs in series were analyzed for reactor performance based on the experimentally determined microbial parameters, stoichiometric parameters and assumed kinetic and diffusion parameters typical of biofilms achieving carbon removal in a CSTR. The assumed parameters were D, D_f, L, K_s and b' as defined in Table 1.4.

The following parameters, which were determined during the kinetic study and the stoichiometric experiment, were also used to analyze the model.

5.5.1 Biofilm thickness.

The thickness of biofilm measured on media from each of the MBBR reactors is presented in Table 1.10. The average biofilm thicknesses for the 1st stage, 40 % media volume reactor, 2nd stage, 40 % media volume reactor and the single sage, 67 % media volume reactor were 310 µm, 290 µm and 250 µm respectively.

Table 1.10 Biofilm thickness for the MBBR reactors used during the study.

	Biofilm thickness (cm)		
	1st stage of 40 % media vol. Reactor	2nd stage of 40 % media vol. reactor	Single stage, 67 % media reactor
Average	0.031	0.029	0.025
Max	0.11	0.042	0.040
Min	0.02	0.015	0.008
Std. Dev. (SD)	0.24	0.13	0.12
Count (N)	19	9	9

5.5.2 Biofilm density

The biofilm density was determined to be 319 mg/cm^3. This was calculated from the following inputs determined from TSS analysis conducted on the biofilm media from the 40 % media volume, first stage MBBR.

Mass of Biofilm per media: = $0.0071 g / media$

Volume of Biofilm per media = $2.227x10^{-8} m^3_{biofilm} / media$

Specific Area of media = $500 m^2 / m^3$

Average biofilm thickness = 270 μm.

5.5.3 Substrate utilization rate

The substrate utilization rate was estimated to be 6.34 / day based on an assumed typical K_s of 0.01 mg/cm^3 and a substrate COD of 0.228 mg/cm^3 used in the substrate utilization experiment. The value however varied from reactor to reactor when applied to the tanks in series model and each value was estimated with the Monod equation. The maximum substrate utilization rate determined was 6.62 /day.

5.5.4 Yield

The true growth yield, Y, was found to be 0.43 mg COD of biomass per mg COD of substrate.

5.5.5 Pseudo – analytical model and results

The Pseudo-analytical approach was used to determine substrate flux into the biofilm and it was found to be 2.7 $mg / cm^3 - d$. This was achieved at an assumed K_s of 20 mg/L COD. At steady state conditions, the models predicted a biofilm thickness of 390μm. The predicted thickness compared very well with experimentally determined average thicknesses and within the range of values presented in Table 9 for media in each of the MBBR reactors. The results also suggested that the biofilm was deep.

Also, the model predicted a reactor volume of 10 liters as that required to achieve a BOD_5 reduction of 90 % based on a media volume of 40 % and a media specific area of 500 m^2/m^3. The results indicate that adequate reactor volume and media were provided to achieve the desired BOD_5 reduction during the study. This discrepancy with what was observed, experimentally may be due to one of the following reasons:

(a) diffusion of dissolved oxygen at high organic loads;

(b) readily biodegradable substrate used in the kinetic and stoichiometric experiment versus slowly biodegrable (particulate) substrate used in the actual reactor and

(c) the deviations of the assumed parametrs from actual.

The model thus over predicted the substrate flux by 10 % and as a result predicted a reactor volume, which was 10 % smaller than the actual reactor volume.

5.5.6 ISO Area Demand for MBBR versus BAF

Table 1.11 ISO Surface area demand for MBBR versus BAF.

Bioreactor	Area required for treatment (m^2)	Clarifier area/Effluent holding tank for backwash. (m^2)	Design criteria
MBBR	0.0055	0.005	* Substrate flux = 2.7 g/m^2-d, * Reactor depth = 1.8 m * Surface area loading to clarifier = 600 gal/ft^2-d (Metcalf and Eddy Inc., 1991) * Flow rate to clarifier = 0.115 m^3/d * Assumes adequate aeration provided to overcome limitation implicated by modeling exercise
BAF	0.0055	0.001	* Mass loading = 3492 g/m^3-d * Media depth = 1.8 m * Surface area loading to tank = 600 gal/ft^2-d (Metcalf and Eddy Inc., 1991) * Flow rate to holding tank = 0.023 m^3/d

Examination of Table 1.11 shows that equal footprints (areas) were required by the bioreactors to meet the effluent BOD_5 criteria for the FPWTS. However, the surface area of clarifier required to remove suspended solids in the MBBR effluent to achieve the desired effluent quality was five times that required to hold BAF effluent for the backwash operation. Due to size limitation of the ISO container for the FPWTS, the BAF bioreactor providing smaller footprint for its treatment units should be preferred to the MBBR.

CHAPTER 6.0 CONCLUSIONS AND RECOMMENDATIONS

6.1 MBBR

The performance of the MBBR was assessed for each of the following reactor configuration:

* single stage 40 % media volume with high SALR;
* single stage 40 % media volume with low SALR;
* single stage 67 % media volume with low SALR;
* 2-stage, 40 % media volume with low SALR and
* 2-stage 40 % media volume low SALR with return activated sludge.

The single stage 67 % media volume reactor achieved the best organic carbon removal among the single stage system configurations. This shows that providing adequate media volume and maintaining appropriate surface area loading rate can ensure effective performance of the MBBR. Conclusions for the 2-stage systems cannot be made since steady state conditions were not achieved. However, the following observations were made: that staging and return of settled effluent sludge back to the reactors achieved lower effluent $TBOD_5$ concentrations. Therefore, staging and staging with return sludge back to the reactors would be possible means to improve on the MBBR system.It is anticipated that return of sludge to the reactors will results in good settling of suspended solids in the effluent. Since $SBOD_5$ was consistently lower than $TBOD_5$ reduced suspended solids in the effluent would improve the effluent quality of the MBBR.

Nitrification and denitrification were also observed when the system was staged. Further investigation on simultaneous nitrification and denitrification would be required especially if nutrient removal would be a goal of the Force provider project.

Comparatively, the MBBR process was easy to operate and maintain and the overall cost of operation and maintenance may be lower than the BAF system, which was an alternative to the MBBR during the studies. The clogging of the BAF and the backwash operation made the BAF relatively more difficult to operate and maintain.

6.2 **BAF Reactor**

- The BAF system achieved excellent suspended solids removal and concentrations were less than the desired effluent concentration of 30 mg/L.
- Effluent BOD_5 concentrations were consistently less than 30 mg/L and it was possible to achieve the target BOD_5 effluent concentration in the first stage column only. Further more, staging of the system allowed organic carbon removal in the first stage column and nitrification and residual carbon removal in the 2^{nd} stage. It may seem reasonable to adjust the size of the system to achieve only carbon removal in a single column if the BAF is selected for the FPWTS. Since space and weight are limitations a single column capable of meeting the effluent goals would be preferred to the 2-stage system.
- Between 10 to 20 % of total daily flow was required for backwashing the system and backwash water TSS concentration was 1860 mg/L.

6.3 **Recommendations**

The BAF met FPWTS effluent guidelines and provided a smaller reactor footprint compared with the MBBR, thus satisfying the ISO container size limitation of FPWTS. Additionally, the BAF removed 16 % more suspended solids than the MBBR.

The BAF seemed to be a viable bioreactor for the project. However cost-benefit evaluation (not provided in this report) of the two systems may be used as the basis of selection for the better option since clogging of the media was a major drawback and made the BAF operation and maintenance difficult.

CHAPTER 7.0 REFERENCES

1. Akunna, J., Bizeau, C., Moletta R Bernet N and Heduit A. (1994). Combine organic and complete nitrogen removal using anaerobic and aerobic upflow filters. *Water science Technology*. **30**(12), 297-306

2. APHA-AWWA-WEF. 20th Ed. (1998). Standard methods for the examination of water and wastewater. APHA, 1015 15th St., DC 20005

3. Bacquet, G., Joret, J.C. Rogalla, F., and Bourbigot, M. M., (1991), Biofilm startup and control in aerated biofilter. *Environ Technol*. **12** 747-756

4. Bae, Wookeun and Rittmann, Bruce E., 1996. Responses of Intracellular Cofactors to Single and Dual Substrate Limitations. Biotechnology and Bioengineering , Vol. 49, pp. 690-699.

5. Bae, Wookeun and Rittmann, Bruce E., 1996. A Structured Model of Dual-Limitation Kinetics. Biotechnology and Bioengineering , Vol. 49, pp. 683-689.

6. Bisogni (Jr.), J. J. and Lawrence, Alonzo, WM., 1970. Relationships between Biological Solids, Retention Time and Settling Characteristics of Activated Sludge. Water Research vol. 5 pp. 753 – 763.

7. Brewer, P.(1996). Poole wastewater treatment BAF plant. *Proceedings of the 2nd BAF symposium*, Cranfield, UK.

8. Canler, J. P., and Perret, J. M. (1994). Biological aerated filters: Assessment of the process based on 12 sewage treatment plants. *Water Sci. Technol.* **29**(10-11), 13-22.

9. Costa Reis, L. G., and Sant'Anna, G.L.(1985). Aerobic treatment of concentrated wastewater in a submerged bed reactor. *Water Res.* **19**, 1341-1345.

10. Desbos, G., Rogalla, F., Sibony, J., and Bourbigot, M.M (1990). Biofiltration as a compact technique for small wastewater Plants. *Water Sci Technol.* **22**(3/4), 145-152.

11. Dillon, N.R., and Thomas V. K. (1990). A pilot-scale evaluation of the "Biocarbone process" for the treatment of settled sewage and for tertiary nitrification of secondary effluent. *Water Sci Technol*. **22**(1/2),305-316.

12. Faup, G. M., Leprince, A., Pannier, M. (1982). Biological nitrification in an upflow fixed bed reactor (UFBR). *Water Sci. Technol.* **14**, 795-810.

13. Gilmore, K. R., Husovitz, K. J., Holst, T. and Love, N. G., 1999. Influence of Organic and Ammonia Loading on Nitrifier Activity and Nitrification Performance For a

Two-Stage Biological Aerated Filter System. Water Sci. Tech. Vol. 39, No. 7, pp. 227-234.

14. Grady, C. P., Glen T. Daigger, H. C. Lim (1999), 2nd Ed. Biological Wastewater Treatment, Marcel Dekker Inc., New York, Basel.

15. Hem, L. J., Rusten, B., Odegaard, H., 1994. Nitrification in a Moving Bed Biofilm Reactor. Water Research vol. 28 (6) pp. 1425 – 1433

16. Husovitz, K.J., Gilmore, K. R., Delahaye, A., Love, N. G., Little, J.C., 1999. Influence of Upflow Liquid velocity on Nitrification in a Biological Aerated Filter. (Unpublished paper).

17. Johnson Chandler, 2001. www.kmt.no.

18. Kent, T. D., Fitzpatrick, C.S.B., and Williams, S.C. (1996). Testing of biological aerated filter (BAF) media. *Water Sci Technol*. **34**(3-4) 363-370

19. Love, N. G., Gilmore, K. G., Husovitz, K. J., Delahaye, A. P., Novak, J. T. and Little, J. C. (1999). Performance of a Biological Aerated Fiter System Treating Domestic Wastewater for BOD, Ammonia and TSS Removal: Pilot Plant Results. Proceedings of the Water Environmetal Federation 72nd Annual Conference and exposition, New Orleans, LA, October 9 – 13, 1999.

20. Mendoza-Espinosa, L.G., and Stephenson, T. (1998). A process model to evaluate the performance of a biological aerated filter. *Biotechnol. Tech.* **12**, 373-376

21. Mendoza-Espinosa, L., Stephenson, T., 1999. A Review of Biological Aerated Filters (BAFs) for Wastewater Treatment, Environmental Engineering Science v.16 (3) 201 – 216.

22. Metcalf and Eddy Inc., Wastewater Engineering, Treatment, Disposal and Reuse, 3rd Ed. McGraw Hill, New York, St. Louis San Francisco, Tokyo Toronto.

23. Odegaard, H., Rusten, B., Siljudalen, J., The Development of the Moving Bed Biofilm Process – From Idea to Commercial. www.waterlink .com

24. Novak J. T. and Love, N. G., Hughes M. (2000). Testing of a Package Wastewater Treatment System and Consultation Services for UTD, Inc.

25. Paffoni, C., Gousailles, M., Rogalla, F., and Gilles, P. (1990) Aerated biofilters for nitrification and effluent polishing. *Water Sci Technol.* **22**(7/8), 181-189

26. Pastorelli, G., Andreottola, G., Canziani, R., Darriulat, C., de Fraja Frangipane and Rozzi, A., 1997. Organic Carbon and Nitrogen Removal in Moving Bed Biofilm Reactors. Water Sci. Tech. Vol. 35, No. 6, pp. 91 – 99

27. Pastorelli, G., Andreottola, G., Canziani, R., de Fraja Frangipane, E., De Pascalis, F., Gurrieri, G. and Rozzi, A. 1997. Pilot-Plant Experiments With Moving Bed Biofilm Reactors. Water Sci. Tech. Vol. 36 No. 1. Pp 43-50.

28. Peladan, J.G., Lemmel, H., and Pujol, R.(1996). High nitrification rate with upflow biofiltration. *Water Sci. Technol.* **34**, 347-353

29. Peng, J., Stevens, D. K., and Yiang, X. (1995). A pioneer project of wastewater reuse in China. *Water Res*. **29**, 357-363.

30. Phipps Scott, (2001). Masters Thesis, Virginia Polytechnic Institute and State University.

31. Pujol, R., Canler, J.P., and Iwema, A. (1992). Biological aerated filters: an attractive and alternative biological process. *Water Sci. Technol*. **26**, 693-702.

32. Pujol, R., Hamon, M., Kandel, X., and Lemmel, H. (1994). Biofilters: Flexible, reliable biological reactors. *Water Sci. Technol.* **29**(10-11), 33-38

33. Rittmann, B.E. and P. L. McCarty, Environmental Biotechnology, principles and applications. McGraw Hill, New York.

34. Rittmann, B. and Saez, P. B., 1992. Communications to the Editor: Accurate Pseudoanalytical Solution foor Steady-State biofilms. Biotechnology and Bioengineering, Vol. 39, pp. 790 – 793

35. Rogalla, F., and Sibony, J. (1992). Biocarbone aerated filters-ten years after: Past, present and plenty potential. *Water Sci Technol.* **26**(9-11), 2043-2048.

36. Rogalla, F., Lamouche, A., Specht, W., and Kleiber, B., (1994). High rate aerated biofilters for plant upgrading. *Water Sci. Technol.* **29**(12), 207-216.

37. Ros, M., and Mejac, B. (1991). Treatment of wastewater in an upflow packed bed reactor. *Water Sci. Technol.* **24**(7), 81-88

38. Ruffer, H., and Rosenwinkel, K.H. (1984). The use of biofiltration for further wastewater treatment. *Water Sci. Technol.* **16**, 241-260

39. Rusten, B. (1984). Wastewater treatment with aerated submerged biological filters. J. Water Pollut Control Fed. 56, 424-431

40. Rusten Bjorn,Kolkinn Odd, Odegaard Hallvard, 1997. Moving Bed Biofilm Reactors and Chemical Precipitation for High Efficiency Treatment of Wastewater from Small Communities. Water Sci. Tech. Vol. 35, No. 6, pp. 71-79.

41. Rusten Bjorn, McCoy Mike, Proctor, Robert and Siljudalen, Jon G., 1998. The Innovative Moving Bed Biofilm Reactor/ Solids Contact Reaeration Process for

Secondary Treatment of Municipal Wastewater. Water Environmental Research, Vol. 70, No. 5 pp. 1083 – 1083.

42. Rusten, Bjorn, Hem, Lars J., Odegaard, Hallvard, 1995. Nitrication of Municipal Wastewater in Moving Bed Biofilm Reactors. Water Environmental Research, Vol. 67, No. 1, pp. 75 – 86.

43. Rusten, B., Johnson, C. H., Devall, S., Davoren, D., Cashion, B. S., 1999. Biological Pretreatment of a Chemical Plant Wastewater in High-Rate Moving Bed Biofilm Reactors. Water Sci. Tech. Vol. 39, No. 10-11, pp. 257-264.

44. Sunner, N., Evans, C., Siviter, C., Bower, T., 1999. The Two-Stage Moving Bed/Activated Sludge Process: An Effective Solution for High-Strength Wastes. Water and Environmental Management Journal, vol. 13 (5) 353 – 358.

45. Teefy, S. M. and Philip C. Singer, (1990) Performance and Analysis of Tracer tests and determine compliance of a disinfection scheme with the SWTR. Journal AWWA 88 – 98

46.

47. Valentis, G., and Lesavre, J. (1989). Wastewater treatment by attached growth microorganisms on geotextile support. *Water Sci. Technol.* **22**(1-2), 43-51.

48. Vedry, B., Paffoni, C., Gousailles, M., and Bernard, C. (1994). First months operation of two biofilter prototypes in waste water plants of Acheres. *Water Sci. Technol.* **29**(10-11), 39-46.

49. Whealer, G., and Cooper-Smith, G.D. (1995). Operational experience with biological aerated filters. *J. CIWEM* **9**, 109-118.

CHAPTER 8.0 APPENDICES

APPENDIX A1 - GENERAL

BOD Supplement ('Candy') Preparation

1. Constituents:
   Protein
   Sugar (Dextrose, Fructose and Galactose)
   Organic acid (Acetic acid and glycerol)

2. Concentrations (in stock solution)

Protein	(yeast)	= 10.0 g/L
Sugar,	Dextrose	= 15.6 g/L
	Fructose	= 15.6 g/L
	Galactose	= 15.6 g/L
	Total	= 46.8 g/L
Organic acid,	Acetic acid	= 98 ml/L x 1 g/L
		= 98 g/L
	Glycerol	= 14 ml/L x 2 g/L
		= 28 g/L
Total		= 126 g/L

   Assuming 1 g COD for each constituents,

Protein	=	10 g/L COD
Sugar	=	47 g/L COD
Organic acid	=	126 g/L COD.
Total COD	=	183 g/L COD

   Assume $BOD_5 : COD_{BO} = 1 : 1.71$

Protein	=	6 g/L BOD5
Sugar	=	28 g/L BOD5
Organic acid	=	76 g/L BOD5
Total BOD	=	110 g/L BOD5

3. VT sewage BOD supplement

Target BOD_5	=	300 mg/L
VT sewage BOD_5	=	260 mg/L (typical during school session)
BOD supplement	=	44 mg/L

   Total flow per day (BAF and MBBR) = 80 gpd (303 lpd)

BOD_5 supplement loading = 13.3 g/d

Volume of stock solution ('Candy') required $= \frac{13.3g}{d} x \frac{L_{stock}}{110g} = \frac{0.12L_{stock}}{d}$

4. Measurement of Constituents

   The following ratios were used to determine the amount of each constituents required daily.

   Protein: Sugar: Organic acid = 1: 5: 13
   Dextrose: Fructose: Galactose = 1: 1: 1
   Acetic acid: glycerol = 1: 3.5

Media surface area relative to reactor volume calculation.

Specific area of media = 500 m^2/m^3

Percentage of media provided (Equivalent to volume fraction of liquid displaced by media in reactor) = 67 %.

Specific surface area of media (with respect to reactor volume)

= 500 $m^2/m^3_{plastic\ volume}$ x (67 $m^3_{plastic\ volume}$/ 100 $m^3_{reactor\ volume}$)

Wastewater characteristics data

Waste Characteristics

Date	Day of study	BOD mg/L	c-BOD mg/L	AMMONIA mg[NH4-N]/L	TOC mg-C/L	TSS, mg/L	VSS, mg/L
15-Nov	1	246.4	-	26.12	76.57	217.0	204.0
20-Nov	6	209.1	154.8	16.69	38.54	337.0	301.0
21-Nov	7	202.8	201.4	19.49	72.13	220.0	201.0
27-Nov	13	-	-	27.08	70.70	335.0	298.0
28-Nov	14	-	-	26.74	76.54	383.0	317.0
30-Nov	16	214.8	180.8	26.96	75.91	277.0	257.0
1-Dec	17	151.2	141.0	25.45	52.94	242.0	222.0
9-Dec	25	306.9	263.2	19.46	94.79	200.0	199.0
10-Dec	26	272.7	248.5	21.17	69.16	223.0	211.0
11-Dec	27	276.5	257.6	24.08	90.07	220.0	206.0
12-Dec	28	237.4	214.4	23.80	81.48	-	-
13-Dec	29	164.4	140.5	21.90	57.52	-	-
19-Dec	35	167.9	145.8	-	-	-	-
Averages		222.7	194.8	23.2	71.4	265.4	241.6

APPENDIX A2 - MBBR DATA

Summary of results for influent and effluent wastewater characteristics in MBBR.

Parameter		Influent	Effluent		
			Overall	High SALR (before day 80)	Low SALR (After day 80)
$TBOD_5$, mg/L	Average	257	104	112	89
	SD	63	64	61	71
	Count (N)	44	44	30	14
$CBOD_5$, mg/L	Average	225	98	81	176
	Max	365	245	179	245
	Min	119	19	19	117
	SD	61	60	47	52
	Count (N)	27	22	18	4
$SBOD_5$, mg/L	Average	189	44	67	25
	SD	41	42	45	27
	Count (N)	26	26	12	14
TSS, mg/L	Average	129	146	106	189
	SD	76	110	49	139
	Count (N)	67	67	35	32
VSS, mg/L	Average	122	135	101	173
	SD	67	99	45	126
	Count (N)	61	63	33	30
NH_3-N, mg/L	Average	24.3	17.24	19	14.9
	SD	5.92	6.52	6	7.23
	Count (N)	33	33	19	14

MBBR BOD_5 data.

MBBR $TBOD_5$

Feed date	Nth day of study	Influent $TBOD_5$ mg/L	Effluent $TBOD_5$ mg/L	% Red in TBOD	$TBOD_5$ SALR (g/m^2-day)	$TBOD_5$ RR (g/m^2-day)	
29-Dec	1	187.5	74.7	**60.17**	15.0	9.0	
3-Jan	6	152.5	37.6	**75.35**	12.2	9.2	
15-Jan	18	342.1	83.7	**75.54**	27.3	20.6	
17-Jan	20	215.9	30.5	**85.89**	17.2	14.8	
19-Jan	22	163.6	22.9	**85.98**	13.1	11.2	
28-Jan	31	301.9	52.5	**82.61**	24.1	19.9	
29-Jan	32	241.2	53.0	**78.01**	19.3	15.0	
30-Jan	33	239.4	56.7	**76.32**	19.1	14.6	
4-Feb	38	157.2	76.6	**51.27**	12.6	6.4	
7-Feb	41	242.5	83.5	**65.58**	19.4	12.7	
11-Feb	45	235.5	76.0	**67.72**	18.8	12.7	
12-Feb	46	243.2	148.2	**39.07**	19.4	7.6	
13-Feb	47	325.3	77.7	**76.11**	26.0	19.8	
14-Feb	48	305.6	61.5	**79.87**	24.4	19.5	
18-Feb	52	223.6	44.7	**79.99**	17.9	14.3	
19-Feb	53	235.2	96.1	**59.15**	18.8	11.1	
20-Feb	54	324.8	139.6	**57.03**	25.9	14.8	
21-Feb	55	314.8	129.0	**59.04**	25.1	14.8	
22-Feb	56	312.4	139.0	**55.50**	24.9	13.8	
25-Feb	59	270.8	100.2	**63.00**	21.6	13.6	
27-Feb	61	364.7	183.539	**49.67**	29.1	14.5	
28-Feb	62	331.5	211.4	**36.22**	26.5	9.6	
1-Mar	63	255.9	137.396	**46.31**	20.4	9.5	
4-Mar	66	251.4	142.875	**43.17**	20.1	8.7	
5-Mar	67	208.0	111.3	**46.49**	16.6	7.7	
6-Mar	68	254.4	205.9	**19.07**	13.3	2.5	
7-Mar	69	267.1	192.9	**27.78**	14.0	3.9	
11-Mar	73	275.7	189.329	**31.32**	14.4	4.5	
15-Mar	77	182.2	135.343	**25.72**	9.5	2.5	
18-Mar	80	428.3	259.8	**39.33**	34.2	13.5	20 g/m2-d
19-Mar	81	316.0	257.60	**18.48**	16.5	3.1	15g/m2-d
25-Mar	87	330.7	117.60	**64.44**	17.3	11.2	
27-Mar	89	287.5	178.90	**37.77**	15.1	5.7	
4-Apr	97	319.8	152.90	**52.18**	16.7	8.7	
22-Apr	115	284.98	114.20	**59.93**	14.9	8.9	
24-Apr	117	204.53	81.60	**60.10**	10.7	6.4	
26-Apr	119	283.88	93.60	**67.03**	14.9	10.0	2-STAGE
17-May	140	170.38	29.27	**82.82**	8.9	7.4	
18-May	141	210.26	20.00	**90.49**	11.0	10.0	
20-May	143	256.50	31.34	**87.78**	13.4	11.8	SLUDGE RETURN
22-May	145	191.25	82.50	**56.86**	10.0	5.7	
23-May	146	149.63	29.92	**80.00**	7.8	6.3	
27-May	150	192.75	32.79	**82.99**	10.1	8.4	
30-May	153	224.03	21.59	**90.36**	11.7	10.6	

MBBR $CBOD_5$ data

MBBR $CBOD_5$

Feed date	Nth day of study	Influent CBOD mg/L	Effluent CBOD mg/L	**% Red. in CBOD**	CBOD SALR (g/m2-day)	CBOD RR (g/m2-day)	
29-Dec	1	168.5	68.1	**59.58**	13.5	8.0	
3-Jan	6	134.3	33.0	**75.42**	10.7	8.1	
15-Jan	18	227.1	78.0	**65.65**	18.1	11.9	
17-Jan	20	199.5	27.2	**86.38**	15.9	13.8	
19-Jan	22	149.4	18.9	**87.32**	11.9	10.4	
28-Jan	31	271.2	55.5	**79.53**	21.7	17.2	
29-Jan	32	213.8	48.3	**77.40**	17.1	13.2	
30-Jan	33	224.8	48.3	**78.50**	17.9	14.1	
4-Feb	38	*138.4*	68.9	**50.21**	11.0	5.5	
12-Feb	46	244.2	133.0	**45.55**	19.5	8.9	
14-Feb	48	308.6	51.8	**83.23**	24.6	20.5	
18-Feb	52	198.5	38.7	**80.49**	15.9	12.8	
19-Feb	53	227.9	95.9	**57.94**	18.2	10.5	
21-Feb	55	303.45	112.4	**62.98**	24.2	15.3	
27-Feb	61	364.4	159.3	**56.29**	29.1	16.4	
1-Mar	63	254.2	130.6	**48.62**	20.3	9.9	
5-Mar	67	193.4	103.1	**46.70**	15.4	7.2	
7-Mar	69	252.7	178.5	**29.37**	20.2	5.9	
18-Mar	80						20g/m2-day
19-Mar	81	300.3	244.8	**18.49**	15.7	2.9	15g/m2-day
27-Mar	89	262.9	168.9	**35.75**	13.8	4.9	
29-Mar	91	284.1	172.8	**39.17**	14.9	5.8	
26-Apr	119	257.06					2-STAGE
29-Apr	122	236.06	117.1	**50.38**	12.4	6.2	
20-May	143	234.86					SLUDGE RETURN
22-May	145	143.40					
23-May	146	119.33					
27-May	150	175.10					

67 % media volume MBBR, BOD_5 and TSS data

MBBR, 67% media BOD5, TSS DATA

		Influent				MBBR Effluent			
Feed date	Nth day of study	TBOD mg/L	SBOD mg/L	TSS, mg/L	VSS, mg/L	TBOD mg/L	SBOD mg/L	TSS, mg/L	VSS, mg/L
29-Dec	1	187.5		-	-	74.7		-	-
17-May	140	170.38	105.36	122	114	46.85		119.00	-
20-May	143	256.50	220.07	78	76	82.50	9.17	188.34	184.29
22-May	145	191.25	160.17	83	-	59	1.50	118.43	109.72
23-May	146	149.63	121.85	25.00	20.33	31.74	3.74	91.77	84.74
27-May	150	192.75	132.73	75.00	64.50	20.04	*8.05*	201.00	174.00
30-May	153	224.03	192.43	90.50	-	-	23.73	356.61	346.42

MBBR $SBOD_5$ data

MBBR $SBOD_5$

Feed date	Nth day of study	Influent SBOD mg/L	Effluent SBOD mg/L	Influent $SBOD_5$ mg/L	Effluent $SBOD_5$ mg/L	**% Red. in SBOD**	$SBOD_5$ SALR (g/m^2 day)	$SBOD_5$ RR (g/m^2 day)
29-Dec	1							
3-Jan	6							
15-Jan	18							
17-Jan	20							
19-Jan	22							
21-Jan	24							
23-Jan	26							
28-Jan	31							
29-Jan	32							
30-Jan	33							
31-Jan	34							
1-Feb	35							
4-Feb	38							
6-Feb	40							
7-Feb	41	162.35	29.18	162.4	29.2	**82.03**	13.0	10.6
11-Feb	45	186.595	25.41	186.6	25.4	**86.38**	14.9	12.9
13-Feb	47	184.262	36.5	184.3	36.5	**80.18**	14.7	11.8
20-Feb	54	223.214	52.0	223.2	52.0	**76.69**	17.8	13.7
22-Feb	56	248.286	62.7	248.3	62.7	**74.75**	19.8	14.8
25-Feb	59	183.625	32.7	183.6	32.7	**82.21**	14.7	12.1
28-Feb	62	229.982	34.9371	230.0	34.9	**84.81**	18.4	15.6
4-Mar	66	175.268	72.125	175.3	72.1	**58.85**	14.0	8.2
6-Mar	68	204.339	98.275	204.3	98.3	**51.91**	16.3	8.5
11-Mar	73	201.804	115.313	201.8	115.3	**42.86**	16.1	6.9
15-Mar	77	137.869	65.3813	137.9	65.4	**52.58**	11.0	5.8
18-Mar	80	264.018	181.8	264.0	181.8	**31.14**	21.1	6.6
25-Mar	87	267.36	62.60	267.4	62.6	**76.59**	14.0	10.7
3-Apr	96	225.34	95.34	225.3	95.3	**57.69**	11.8	6.8
4-Apr	97	148.08	33.99	148.1	34.0	**77.05**	7.8	6.0
22-Apr	115	175.95	27.00	176.0	27.0	**84.65**	9.2	7.8
24-Apr	117	196.50	40.35	196.5	40.4	**79.47**	10.3	8.2
26-Apr	119	194.40	*13.29*	194.4	13.3	**93.16**	10.2	9.5
29-Apr	122	196.93	*21.54*	196.9	*21.5*	***89.06***	10.3	9.2
17-May	140	105.36	*9.65*	105.4	*9.6*	***90.84***	5.5	5.0
18-May	141	179.79	*6.44*	179.8	*6.4*	***96.42***	9.4	9.1
20-May	143	220.07	17.55	220.1	*17.6*	***92.03***	11.5	10.6
22-May	145	160.17	9.17	160.2	*9.2*	***94.27***	8.4	7.9
23-May	146	121.85	3.06	121.9	*3.1*	***97.49***	6.4	6.2
27-May	150	132.73	*4.08*	132.7	*4.1*	***96.93***	6.9	6.7
30-May	153	192.43	1.82	192.4	*1.8*	***99.05***	10.1	10.0

MBBR TSS Data

Feed date	Nth day of study	Influent TSS, mg/L	Effluent TSS, mg/L	**% Red. In TSS**	Influent VSS, mg/L	Effluent VSS, mg/L	**% Red. in VSS**	Solids in effluent g/day	Wasted sludge g/day	**Total Solids g/day**
29-Dec	1	-	-	**-**	-	-	**-**	-		
3-Jan	6	193.0	59.0	**69.4**	145.0	56.0	**61.4**	10.62		
5-Jan	8	241.0	12.0	**95.0**	196.0	11.0	**94.4**	2.16		
10-Jan	13	85.0	30.0	**64.7**				5.40		
11-Jan	14	165.0	22.8	**86.2**	159.0	21.8	**86.3**	4.10		
17-Jan	20	228.0	90.0	**60.5**	196.0	82.0	**58.2**	16.20		
19-Jan	22	120.0	13.0	**89.2**	109.0	13.0	**88.1**	2.34		
21-Jan	24	136.0	81.0	**40.4**	129.0	80.0	**38.0**	14.58		
23-Jan	26	193.0	102.5	**46.9**	168.0	88.8	**47.2**	18.45		
30-Jan	33	47.5	117.3	**INC**	47.5	115.3	**INC**	21.12		
31-Jan	34	49.5	104.7	**INC**	43.5	91.3	**INC**	18.84		
4-Feb	38	67.0	92.7	**INC**	53.0	77.3	**INC**	16.68		
5-Feb	39	137.5	82.0	**40.4**	118.5	80.0	**32.5**	14.76		
6-Feb	40	52.0	114.0	**INC**	48.8	108.0	**INC**	20.52		
11-Feb	45	73.8	102.7	**INC**	65.3	88.7	**INC**	18.48		
12-Feb	46	210.0	156.0	**25.7**	200.0	154.0	**23.0**	28.08		
13-Feb	47	191.4	127.2	**33.5**	169.6	119.0	**29.8**	22.90		
14-Feb	48	269.1	199.9	**25.7**	239.1	179.8	**24.8**	35.98	14.1	**50.1**
18-Feb	52	83.7	82.7	**1.2**	75.8	67.3	**11.2**	14.88		
20-Feb	54	162.0	146.7	**9.4**	149.8	143.7	**4.1**	26.40		
21-Feb	55	135.0	149.2	**INC**	127.6	141.7	**INC**	26.85		
22-Feb	56	85.5	155.0	**INC**	81.0	148.2	**INC**	27.89		
25-Feb	59	130.7	111.4	**14.8**	118.9	98.9	**16.8**	20.06		
26-Feb	60	62.2	157.5	**INC**	59.5	144.2	**INC**	28.34		
27-Feb	61	56.2	119.3	**INC**	53.0	103.4	**INC**	21.47	9.4	**30.9**
28-Feb	62	167.6	256.0	**INC**	153.4	229.3	**INC**	46.08		
1-Mar	63	134.9	134.9	**INC**	127.0	130.2	**INC**	24.29	8.2	**-**
4-Mar	66	112.3	120.2	**INC**	103.7	104.7	**INC**	21.63		
5-Mar	67	63.7	97.5	**INC**	58.2	85.4	**INC**	17.55		
6-Mar	68	87.1	97.0	**INC**	82.5	87.0	**INC**	17.46		
7-Mar	69	74.4	97.8	**INC**	72.1	89.8	**INC**	17.61		
11-Mar	73	158.6	109.0	**31.2**	144.1	92.3	**36.0**	19.63		
12-Mar	74	92.9	77.1	**17.0**	92.5	-		13.87	0.4	**14.3**
14-Mar	76	143.2	81.1	**43.4**	130.5	76.6	**41.3**	14.59		
15-Mar	77	55.0	101.4	**INC**	55.3	95.2	**INC**	18.26		
18-Mar	80	210.2	125.6	**40.2**	184.3	116.2	**36.9**	22.62	0.7	**23.3**
19-Mar	81	72.24	106.27	**INC**	66.27	93.05	**INC**	12.24		
21-Mar	83	104.05	59.29	**43.0**	83.62	54.73	**34.6**	6.83	7.66	**14.49**
22-Mar	84	62.17	96.67	**INC**	60.83	87.33	**INC**	11.14	6.68	**17.81**
25-Mar	87	57.24	88.31	**INC**	54.14	79.61	**INC**	10.17		
26-Mar	88	62.35	138.74	**INC**	61.00	132.00	**INC**	15.98		
27-Mar	89	55.70	116.65	**INC**	52.53	112.99	**INC**	13.44		
28-Mar	90	49.97	57.24	**INC**	48.68	54.14	**INC**	6.59		

29-Mar	91	130.86	223.91	**INC**	122.64	210.87	**INC**	25.79		
1-Apr	94	238.03	155.95	**34.5**	187.69	141.43	**24.6**	17.97	6.55	**24.51**
2-Apr	95	288.87	319.42	**INC**	249.16	277.11	**INC**	36.80		
3-Apr	96	451.77	844.79	**INC**	407.92	744.79	**INC**	97.32	7.00	**104.31**
4-Apr	97	225.66	192.30	**14.8**	201.53	170.38	**15.5**	22.15	6.07	**28.22**
5-Apr	98	94.50	179.00	**INC**	90.50	169.00	**INC**	20.62	6.08	**26.70**
9-Apr	102	300.23	225.84	**24.8**	252.91	195.45	**22.7**	26.02		
10-Apr	103	113.50	171.05	**INC**	101.50	151.95	**INC**	19.71		
11-Apr	104	154.50	130.10	**15.8**	153.00	125.44	**18.0**	14.99	2.00	**16.99**
12-Apr	105	182.83	237.99	**INC**	163.02	209.86	**INC**	27.42		
22-Apr	115	149.00	159.57	**INC**	104.00	115.96	**INC**	18.38		
24-Apr	117	89.33	206.89	**INC**	88.00	196.90	**INC**	23.83		
26-Apr	119	59.50	187.03	**INC**	**-**	173.61	**0.0**	21.55		
29-Apr	122	143.86	252.15	**INC**	135.96	233.39	**INC**	29.05		
4-May	127	126.24	194.29	**INC**	111.29	173.01	**INC**	22.38		
5-May	128	128.00	198.49	**INC**	122.37	193.52	**INC**	22.87		
9-May	132	209.43	232.46	**INC**	198.00	217.67	**INC**	26.78		
11-May	134	76.00	292.29	**INC**	**-**	268.06	**0.0**	33.67		
17-May	140	121.72	100.46	**17.5**	114.22		**-**	11.57		
18-May	141	97.00	246.00	**INC**	87.50	234.00	**INC**	28.34		
20-May	143	77.50	128.44	**INC**	76.00	120.00	**INC**	14.80		
22-May	145	82.67	72.80	**11.9**		70.60	**-**	8.39		
23-May	146	25.00	43.50	**INC**		38.50	**-**	5.01		
27-May	150	75.00	150	**INC**	64.50	131.62	**INC**	17.24		
30-May	153	90.50	239.62	**INC**			**-**	27.60		
	INC	No reduction in TSS								

MBBR Ammonia Data

Feed date	Nth day of study	Influent NH_3-N mg/L	Effluent NH_3-N mg/L	**% Red. in NH_3-N**	
29-Dec	1	14.36	10.61	**26.12**	
3-Jan	6	31.53	18.26	**42.10**	
5-Jan	8	16.88	16.46	**2.49**	
10-Jan	13	15.15	12.52	**17.38**	
15-Jan	18	22.96	22.12	**3.66**	
17-Jan	20	29.76	29.18	**1.98**	
19-Jan	22	17.50	17.05	**2.56**	
28-Jan	31	18.87	15.29	**18.99**	Composite sampling on the Feb. 1
5-Feb	39	23.58	17.50	**25.77**	
7-Feb	41	28.78	22.68	**21.21**	
13-Feb	47	29.51	24.1	**18.41**	
19-Feb	53	30.91	24.53	**20.65**	
21-Feb	55	29.18	21.67	**25.72**	
26-Feb	60	27.47	10.75	**60.86**	
28-Feb	62	29.34	19.60	**33.21**	
5-Mar	67	18.03	12.94	**28.26**	
7-Mar	69	19.26	14.84	**22.97**	
12-Mar	74	26.04	22.40	**13.98**	
14-Mar	76	29.18	28.17	**3.45**	Reduced flow from 125ml/min to 80ml/min on Mar 18
19-Mar	81	26.18	19.26	**26.42**	
26-Mar	88	22.21	14.92	**32.83**	
28-Mar	90	27.85	22.65	**18.70**	
2-Apr	95	27.48	20.70	**24.69**	
4-Apr	97	25.30	17.84	**29.48**	
9-Apr	102	29.40	16.90	**42.51**	
11-Apr	104	24.86	15.76	**36.62**	
24-Apr	117	30.17	17.57	**41.76**	Commenced two-staged system operations on April 18
5-May	128	28.12	19.56	**30.47**	
7-May	130	34.27	23.28	**32.06**	
9-May	132	22.88	11.68	**48.96**	
17-May	140	12.96	3.74	**71.15**	Commenced RAS to 2-staged system operations on May 22
1-Jun	155	17.64	2.56	**85.51**	
2-Jun	156	14.25	1.99	**86.05**	

APPENDIX A3 – BAF DATA

Summary of results for 1st and 2nd stage effluent wastewater characteristics in BAF.

Parameter		Effluent	
		First stage BAF	Second stage BAF
$TBOD_5$, mg/L	Average	36	16
	Std. Dev. (SD)	22	8.17
	Count (N)	9	39
$CBOD_5$, mg/L	Average	30	11
	Std. Dev. (SD)	18	7.72
	Count (N)	8	25
$SBOD_5$, mg/L	Average	-	15
	Std. Dev. (SD)	-	8.4
	Count (N)	-	10
TSS, mg/L	Average	-	18
	Std. Dev. (SD)	-	22
	Count (N)	-	45
NH_3-N, mg/L	Average	15.4	3.63
	Std. Dev. (SD)	10.3	6.78
	Count (N)	9	21
NO_3-N, mg/L	Average	0.11	4.21
	Std. Dev. (SD)	0.14	1.24
	Count (N)	4	4

First column BAF $TBOD_5$

FIRST COLUMN BAF - $TBOD_5$

Feed date	Nth day of study	Influent BOD mg/L	1st stage effl. BOD	$TBOD_5$ VLR (g/m^3-day)	$TBOD_5$ RR (g/m3-day)
29-Dec	1	187.5			
3-Jan	6	152.5			
5-Jan	8	247.5			
10-Jan	13	136.1			
11-Jan	14	165.0			
15-Jan	18	342.1			
17-Jan	20	215.9			
19-Jan	22	163.6			
21-Jan	24	259.6			
23-Jan	26	324.8			
28-Jan	31	301.9			
29-Jan	32	241.2			
30-Jan	33	239.4			
31-Jan	34	246.0			
1-Feb	35	249.1			
4-Feb	38	157.2			
6-Feb	40	245.6			
7-Feb	41	242.5			
11-Feb	45	235.5			
12-Feb	46	243.2			
13-Feb	47	325.3			
14-Feb	48	305.6			
18-Feb	52	223.575			
19-Feb	53	235.225			
20-Feb	54	324.838			
21-Feb	55	314.8			
22-Feb	56	312.35			
25-Feb	59	270.825			
27-Feb	61	364.663			
28-Feb	62	331.45			
1-Mar	63	255.913			
4-Mar	66	251.388			
5-Mar	67	208.00			
6-Mar	68	254.413			
7-Mar	69	267.1			
11-Mar	73	275.675			
18-Mar	80	428.25			
19-Mar	81	316.00	**23.19**	4186.4	3879.1
27-Mar	89	287.46	73.55	3808.3	2833.9
5-Apr	98	268.48	14.04	3556.8	3370.8
24-Apr	117	204.53	17.50	2709.5	2477.8
29-Apr	122	254.88	20.88	3376.6	3100.0
4-May	127	280.73	**48.99**	3719.0	3070.1
17-May	140	170.38	14.40	2257.2	2066.4
20-May	143	256.50	49.97	3398.1	2736.1
27-May	150	192.75	57.23	2553.6	1795.4

First and Second Column BAF $TBOD_5$

FIRST & SECOND COLUMN BAF $TBOD_5$

Feed date	Nth day of study	Influent BOD mg/L	Effluent BOD mg/L	$TBOD_5$ VLR (g/m^3-day)	$TBOD_5$ RR (g/m3-day)	**% Red in BOD-BAF 1&2**
29-Dec	1	187.5				
3-Jan	6	152.5				
5-Jan	8	247.5				
10-Jan	13	136.1				
11-Jan	14	165.0				
15-Jan	18	342.1				
17-Jan	20	215.9				
19-Jan	22	163.6				
21-Jan	24	259.6				
23-Jan	26	324.8				
28-Jan	31	301.9				
29-Jan	32	241.2				
30-Jan	33	239.4				
31-Jan	34	246.0				
1-Feb	35	249.1	*16.10*	3300.4	3087.1	**93.54**
4-Feb	38	157.2	*17.83*	2082.9	1846.7	**88.66**
6-Feb	40	245.6	21.02	3254.0	2975.6	**91.44**
7-Feb	41	242.5	*6.73*	3212.3	3123.1	**97.22**
11-Feb	45	235.5	15.75	3119.2	2910.6	**93.31**
12-Feb	46	243.2	*12.4*	3221.4	3057.2	**94.90**
18-Feb	52	223.575	44.9	2961.9	2366.6	**79.90**
19-Feb	53	235.225	12.37	3116.3	2952.4	**94.74**
20-Feb	54	324.838	*12.12*	4303.4	4142.9	**96.27**
21-Feb	55	314.8	*5.83*	4170.5	4093.3	**98.15**
22-Feb	56	312.35	25.36	4138.0	3802.0	**91.88**
25-Feb	59	270.825	27.78	3587.9	3219.8	**89.74**
27-Feb	61	364.663	14.08	4831.0	4644.5	**96.14**
28-Feb	62	331.45	9.8	4391.0	4260.8	**97.03**
1-Mar	63	255.913	*7.47*	3390.3	3291.4	**97.08**
4-Mar	66	251.388	21.2417	3330.4	3049.0	**91.55**
5-Mar	67	208.00	9.96	2755.6	2623.7	**95.21**
6-Mar	68	254.413	9.20	3370.5	3248.6	**96.38**
7-Mar	69	267.1	9.83	3538.5	3408.4	**96.32**
11-Mar	73	275.675	19.95	3652.1	3387.8	**92.76**
15-Mar	77	182.2	21.825	2413.8	2124.6	**88.02**
18-Mar	80	428.25	*27.0979*	5673.5	5314.5	**93.67**
19-Mar	81	316.00	**11.70**	4186.4	4031.4	**96.30**
25-Mar	87	330.73	12.22	4381.4	4219.6	**96.31**
27-Mar	89	287.46	20.19	3808.3	3540.9	**92.98**
29-Mar	91	331.76	20.08	4395.2	4129.2	**93.95**
4-Apr	97	319.75	16.71	4236.0	4014.6	**94.77**
5-Apr	98	268.48	*7.49*	3556.8	3457.5	**97.21**
24-Apr	117	204.53	12.45	2709.5	2544.7	**93.91**
26-Apr	119	283.88	11.50	3760.8	3608.4	**95.95**
29-Apr	122	254.88	**26.83**	3376.6	3021.1	**89.47**

4-May	127	280.73	18.43	3719.0	3474.9	**93.44**
5-May	128	255.94	*7.35*	3390.7	3293.3	**97.13**
11-May	134	293.66	11.11	3890.4	3743.2	**96.22**
17-May	140	170.38	8.55	2257.2	2143.9	**94.98**
18-May	141	210.26	27.42	2785.6	2422.3	**86.96**
20-May	143	256.50	10.19	3398.1	3263.1	**96.03**
22-May	145	191.25	8.96	2533.7	2414.9	**95.31**
23-May	146	149.63	5.00	1982.2	1916.1	**96.66**

First and Second Column BAF TSS DATA

1 & 2 BAF – TSS DATA

Feed date	Nth day of study	Influent TSS, mg/L	Effluent TSS, mg/L	**% Red. in TSS**	Influent VSS, mg/L	Effluent VSS, mg/L	**% Red. in VSS**	Solids in effluent g/day	Backwash Solids g/day	**Total Solids g/day**
29-Dec	1	-			-					
3-Jan	6	193.0			145.0					
5-Jan	8	241.0			196.0					
10-Jan	13	85.0								
11-Jan	14	165.0			159.0					
15-Jan	18	-			-					
17-Jan	20	228.0			196.0					
19-Jan	22	120.0			109.0					
21-Jan	24	136.0			129.0					
23-Jan	26	193.0			168.0					
28-Jan	31	165.0			155.0					
29-Jan	32	195.0			-					
30-Jan	33	47.5			47.5					
31-Jan	34	49.5	9.50	**80.8**	43.5	9.50	**78.2**	1.09		
1-Feb	35	88.0	7.50	**91.5**	77.5	6.00	**92.3**	0.86		
4-Feb	38	67.0	25.50	**61.9**	53.0	15.00	**71.7**	2.94		
5-Feb	39	137.5	52.33	**61.9**	118.5	51.00	**57.0**	6.03		
6-Feb	40	52.0	26.80	**48.5**	48.8	24.80	**49.2**	3.09		
11-Feb	45	73.8	6.20	**91.6**	65.3	6.20	**90.5**	0.71		
12-Feb	46	210.0	2.8	**98.7**	200.0	1.6	**99.2**	0.32		
18-Feb	52	83.7	8.1	**90.3**	75.8	6.9	**90.9**	0.93		
20-Feb	54	162.0	1.17	**99.3**	149.8	-	**-**	0.13		
25-Feb	59	130.7	4.00	**96.9**	118.9	3.71	**96.9**	0.46		
26-Feb	60	62.2	3.42	**94.5**	59.5	3.42	**94.3**	0.39		
4-Mar	66	112.3	2.75	**97.6**	103.7	-		0.32	10.49	**10.8**
6-Mar	68	87.1	2.0	**97.7**	82.5	-		0.23		
11-Mar	73	158.6	5.2	**96.7**	144.1	4.4	**96.9**	0.60		
15-Mar	77	55.0	36.8	**33.1**	55.3	34.6	**37.5**			
18-Mar	80	210.2	19.3	**90.8**	184.3	-		2.23	83.04	**85.3**
19-Mar	81	72.24	12.40	**82.8**	66.27	10.80	**83.7**			
1-Apr	94	238.03	32.56	**86.3**	187.69	24.75	**86.8**	3.75	11.52	**15.3**
2-Apr	95	288.87	18.48	**93.6**	249.16	16.67	**93.3**			
3-Apr	96	451.77	42.00	**90.7**	407.92	35.00	**91.4**	4.84	11.40	**16.2**
4-Apr	97	225.66	18.00	**92.0**	201.53	15.50	**92.3**			
5-Apr	98	94.50	5.75	**93.9**	90.50	-	**-**	0.66	14.20	**14.9**
9-Apr	102	300.23	120.71	**59.8**	252.91	109.94	**56.5**			
11-Apr	104	154.50	9.00	**94.2**	153.00	-	**-**	1.04	16.29	**17.3**
12-Apr	105	182.83	15.13	**91.7**	163.02	14.09	**91.4**			
24-Apr	117	89.33	1.75	**98.0**	88.00	-	**-**			
26-Apr	119	59.50	3.60	**93.9**	0.00	-	**-**			
29-Apr	122	143.86	12.67	**91.2**	135.96	11.90	**91.2**			
4-May	127	126.24	22.06	**82.5**	111.29	20.28	**81.8**			
5-May	128	128.00	17.00	**86.7**	122.37	16.50	**86.5**			
9-May	132	209.43	3.40	**98.4**	198.00	-	**-**			

17-May	140	121.72	8.80	**92.8**	114.22	-	**-**		
20-May	143	77.50	11.00	**85.8**	76.00	-			
22-May	145	82.67	36.45	**55.9**	-				

First and Second Column BAF AMMONIA DATA

BAF REACTOR AMMONIA DATA

Feed date	Nth day of study	Influent NH_3-N mg/L	Effluent NH_3-N mg/L	**% Red. in NH_3**	1st stage BAF effl. NH_3-N	**% Red. in NH_3**
29-Dec	1	14.36				
3-Jan	6	31.53				
5-Jan	8	16.88				
10-Jan	13	15.15				
15-Jan	18	22.96				
17-Jan	20	29.76				
19-Jan	22	17.50				
28-Jan	31	18.87				
5-Feb	39	23.58	17.95	**23.87**		
7-Feb	41	28.78	0.34	**98.83**		
19-Feb	53	30.91	5.66	**81.70**		
21-Feb	55	29.18	0.18	**99.38**		
26-Feb	60	27.47	0.28	**98.98**		
28-Feb	62	29.34	0.36	**98.76**		
5-Mar	67	18.03	0.22	**98.76**		
12-Mar	74	26.04	0.06	**99.78**		
14-Mar	76	29.18	13.33	**54.32**	19.85	**31.96**
19-Mar	81	26.18	4.42	**83.10**	4.09	**84.39**
26-Mar	88	22.21	0.47	**97.88**		
28-Mar	90	27.85	0.27	**99.03**	15.52	**44.27**
2-Apr	95	27.48	0.57	**97.92**		
4-Apr	97	25.30	0.37	**98.54**		
9-Apr	102	29.40	0.47	**98.40**	18.58	**36.80**
11-Apr	104	24.86	1.68	**93.24**	5.81	**76.62**
24-Apr	117	30.17	0.10	**99.67**	14.72	**51.22**
5-May	128	28.12	25.10	**10.75**		
7-May	130	34.27	2.99	**91.27**	34.27	**0.00**
9-May	132	22.88	0.80	**96.49**	23.30	**-1.84**
17-May	140	12.96	0.50	**96.11**	2.32	**82.07**

APPENDIX A4 - BAF/ MBBR NITRATE DATA

Date	BAF-1, mg/L- NO3-N	BAF-2,mg/L- NO3-N	MBBR,mg/L- NO3-N
5/8/2001	0.14	5.77	-
5/10/2001	0.00	2.97	-
5/11/2001	0.30	3.51	-
5/17/2001	0.00	4.61	0.21
5/18/2001	-	-	
5/19/2001	-	-	0
average	0.11	4.21	0.11
stdev	0.14	1.24	0.15
N	4	4	2

APPENDIX A5 - KINETIC AND STOICHIOMETRIC EXPERIMENT DATA

Yield Data

Sample	Total COD (mg/L)	Soluble COD (mg/L)	Biomass COD (mg/L)
1	282.49	257.31	25.18
2	234.49	188.07	46.43
3	180.98	85.77	95.21

Substrate Utilization Rate Data

Experiment 1		Experiment 2	
Time (min)	DOC (mg-C/L)	Time (min)	DOC (mg-C/L)
0	126.8	0	131.3
10	134.6	10	122
20	144.4	20	132.1
30	133.6	30	117.1
40	124.1	40	110.7
50	120.1	50	104.9
60	121.7	60	104.2
70	112.6	70	97.96
80	106.4	80	94.53
90	101.3	90	93.58
100	105.2	100	85.35
110	121.8	110	79.86
120	111.7	120	77.78
130	121.7	130	72.91
140	98	140	71.11
150	105.8	150	70.21
160	101.6	160	61.98
170	101.4	170	60.39
180		180	61.45
		190	55.34
		200	53.7

SECTION II - SULFIDE-INDUCED CORROSION IN ANAEROBIC DIGESTER GAS PIPING

CHAPTER 1.0 INTRODUCTION AND BACKGROUND

1.1 Background

The research project described in this report was conducted during summer 2000, and was supported by are Edna Bailey Sussman Foundation fellowship and sponsored by Peppers Ferry Regional Wastewater Treatment Authority (PFRWTA).

The purpose of the fellowship was to provide the researcher with a better perspective of environmental problems encountered by an industry and to work with the industry to develop a solution to a selected problem. Some problems identified with the PFRWTF operations included foaming in the activated sludge basins and accumulation of deposits in the anaerobic digester gas piping. With the time frame available for the fellowship and other considerations, the gas piping was selected for study. The goal of the research project was to:

- investigate the possible causes for deposition in the gas piping,
- investigate the composition of the deposit, and
- recommend options for abating the formation of the deposit.

Both laboratory experiments and field analysis were done to address these objectives.

A laboratory scale reactor similar to the anaerobic digesters at the PFRWTF plant was set up to study the effect of digester gas on iron pipe and shavings, composed of the same material used to convey digester gas in the field. The study was carried out for 62 days and the following features were investigated: the composition of the digester gas, and sulfate, sulfide and sulfite concentrations in both the influent and effluent water phases of sludge samples. Mixed liquor suspended solid (MLSS) and mixed liquor volatile suspended solids (MLVSS), pH and gas production were monitored to ensure reactor performance.

Results of the study showed approximately 90% removal of soluble sulfate across the laboratory reactor. Soluble sulfite and sulfide concentrations were extremely low to undetectable in the effluent. Given the reduced redox conditions in the digester, this suggests that sulfate was reduced to sulfide. By assuming chemical equilibrium, the majority of the sulfide probably remained as an insoluble sulfide precipitate in the sludge.

The metal iron pieces exposed to the digester gas showed increasing weight trends and formation of a black film on the metal surfaces. The shaving became brittle after being exposed to the digester gas for 62 days. The concentration of hydrogen sulfide in the gas that had been in contact with the iron pieces was consistently lower than the concentration found in the gas not exposed to the iron pieces. These results indicate that the gas was reacting with the iron pieces and causing corrosion.

Field data revealed increasing levels of hydrogen sulfide gas in the PFRWTF digester gas, but the concentrations were far less than concentrations that are reported to cause inhibition of methanogenesis in anaerobic digesters. Analysis of the deposit found in the PFRWTF gas piping revealed the presence of elemental oxygen, organic carbon, iron and sulfur at 57%, 25%, 15% and 3% by weight. The ratio of iron to sulfur mass suggests that approximately 6 % of iron was needed to react with sulfur if sulfur was a limiting reactant in the digester under oxygen free environment. Iron was 15 % in the deposit, suggesting that other forms of iron compounds such as ferrous hydroxide, which is also formed during the formation iron sulfide by sulfate reducing bacteria in an oxygen free environment, were present.

It was recommended that further analysis be carried out to determine the compounds formed by these elements in the deposit. It was also suggested that the major cations associated with the particulate sulfide in the sludge should be determined. The presence of cations such as iron in the wastewater going to the PFRWTF may be protecting methanogens in the digester from free hydrogen sulfide gas exposure by enabling sulfide precipitation in the digester. At the same time, precipitated counter cations may become nutrient limiting for the methanogens if not provided in excess.

1.2 Introduction

The Peppers ferry regional wastewater treatment facility (PFRWTF) is a secondary treatment plant, which treats wastewater from the City of Radford, Town of Pulaski, Town of Dublin and portions of Pulaski and Montgomery counties. The treatment plant consists of primary clarifiers, activated sludge aeration basins, secondary clarifiers, chlorine contact basins and anaerobic digestion for waste stabilization and production of methane gas as a by product. The industrial waste input of the plant was originally about 25%, but has increased in recent times to about 40% due to the growth of industries in the region and the requirement of these industries to meet USEPA disposal regulations and standards. The constituents of concern in the industrial wastewater include sulfates, sodium, zinc, cadmium, nickel ions. The wastewater has been monitored to determine levels of these ions in the wastewater and to determine if the ions adversely impact the operation of the plant and safety of the plant employees.

In 1994, an industry in Pulaski, Magnox Inc. which generates very high concentration of sodium sulfate (about 70,000 mg/L) from its industrial activities, wanted to introduce its waste into the collection system of the PFRWTF. An activated sludge pilot study was conducted at the treatment plant in 1995 by Olver, Inc. (1995) to investigate the effect of sodium and sulfate ions on the overall performance of the activated sludge process. Due to difficulties involved in carrying out a pilot study for the anaerobic digester, no experiments were performed to evaluate that process. However, data collected from the pilot study for the activated sludge process and information from the technical literature were used to evaluate the impact of the industrial input from Magnox on the anaerobic digester. The study was carried out for 3 months and the conclusions are summarized below:

- *Aerobic Wastewater Treatment.* The pilot treatment process effectively treated the blend of domestic, industrial and Magnox wastewater discharged to PFRWTF without adversely affecting the pilot plant's performance or the overall treatment efficiency.
- *Anaerobic Digester.* Waste Activated Sludge (WAS) analyses indicated that the proposed Magnox discharge would have minimal impact on the general quality of the

WAS fed to the anaerobic digester other than to increase the concentration of sodium and sulfate ions. At high loadings of sulfate to an anaerobic digester, it is possible to inhibit the methanogenic process by generating toxic concentrations of aqueous sulfide in the digester. The PFRWTF anaerobic digester was operating at a sulfide concentration of 126 mg/L on average at the time of study. Sulfide concentration levels in the digester were expected to rise above the accepted inhibition value of 200 mg/L (Kroiss et al., 1983; Olver Inc., 1995).

CHAPTER 2.0 OBJECTIVES

Recent inspection of a section of the methane gas piping adjacent to PFRWTF primary digesters A revealed a 2-inch thick black lining in the 6-inch gas pipe. PFRWTF staff sponsored a study of this problem. The objectives of the research project were to:

1. investigate possible causes for deposition in the gas piping
2. determine composition of the deposit, and
3. recommend options for abating the formation of the deposit.

CHAPTER 3.0 LITERATURE REVIEW

3.1 Fate of sulfur in Bioreactors

Sulfur in wastewater can exist in one or more of the following forms: sulfate, sulfite, sulfide, sulfur, or in association with organic matter. Under anaerobic conditions, sulfur as sulfate (+6 oxidation state) can be reduced to sulfite (+4), sulfur (0) or sulfide (-2). The following generalized reactions are typical.

$$\text{Organic matter} + SO_4^{-2} \xrightarrow{bacteria} S^{-2} + H_2O + CO_2 \qquad \text{(Equation 2.1)}$$

$$S^{-2} + 2H^{+} \rightarrow H_2S \qquad \text{(Equation 2.2)}$$

$$H_2S \leftrightarrow HS^{-} + H^{+} \qquad \text{(Equation 2.3)}$$

$$HS^{-} \leftrightarrow S^{-} + H^{+} \qquad \text{(Equation 2.4)}$$

At the neutral pH required for anaerobic treatment, only dissociation of H_2S in equation 2.3 and biologically mediated sulfate reduction (equation 2.1) are important.

Sulfate reduced biologically under anaerobic conditions to sulfide can cause problems in anaerobic digesters at high concentrations. Problems usually of concern in anaerobic treatment processes that are associated with sulfur include competition between sulfate reducing bacteria (SRB) and methane producing bacteria (MPB) for electron donors, sulfide inhibition of methanogenesis and corrosion of concrete and metals caused by sulfides (Mara and Williams, 1971; Harada et al. 1994).

In anaerobic digesters where methanogens have been inhibited in the presence of sulfur, SRBs are normally the dominant species. It was reported that SRB could utilize between 53 to 92 % of available substrate electrons (Isa et al., 1985). The kinetics of competition for the available electron donors between SRBs and MPBs revealed that SRBs have higher affinity (lower K_s, half-saturation coefficient) and competitive maximum specific substrate utilization rates (q_{max}) for hydrogen and acetate, which are the major methane precursors for MPBs. Ks reported by Isa et al. (1985) for SRBs and MPBs for acetate were 0.2 mM and 3.0 mM of

substrate and for hydrogen 0.001 mM and 0.006 mM of substrate, repectively. The q_{max} for SRBs and MPBs for acetate were 74 and 45 mM substrate removed/ g VSS per day respectively and for hydrogen were 112 and 123 mM substrate removed/ g VSS per day respectively.

Thermodynamically, the reduction of sulfate is energetically slightly more favorable than the reduction of bicarbonate (Khosrovi et al., 1971). Table 2.1 summarizes the free energy changes of the reactions that are pertinent to sulfate reduction and methanogenesis. The values suggest that hydrogen and acetate are preferentially used as electron donors for sulfate reduction relative to methanogenesis under standard conditions. These findings indicate that SRB have a tendency to out-compete MPB for both acetate and hydrogen under the conditions found in anaerobic digesters.

Table 2.1 Free energy changes for biologically-mediated sulfate reduction and methane production.

Reactions				Free Energy (KJ/reaction)
Hydrogen -Consuming				
SRB:	$4H_2 + SO_4^{2-} + H^+$	$\rightarrow$	$HS^- + 4H_2O$	-152.2*
MPB:	$4H_2 + HCO_3- + H^+$	$\rightarrow$	$CH_4 + 3H_2O$	-135.6+
Acetate-Consuming				
SRB:	$CH_3COO^- + SO_4^{2-} + H^+$	$\rightarrow$	$HS^- + 2HCO_3^-$	-47.6*
MPB:	$CH_3COO^- + H_20$	$\rightarrow$	$CH_3 + HCO_3^-$	-31.0+

* Data from Widdel, 1988 (16) + Data from Thauer et al., 1977 (17)

Kroiss and Wabnegg (1983) related methanogenesis inhibition to the level of free H_2S in solution, which they reported is toxic to MPBs. Sulfide can exist in the forms H_2S, HS^-, S^{2-} in solution and H_2S in biogas. The concentration of the various forms of sulfide in solutions depends on the pH of the solution as shown in Table 2.2. At pH values normally found in anaerobic digesters, the dominant species in solution is free H_2S. Kroiss and Wabnegg (1983) found that a free H_2S level of 50 mg/L inhibited acetoclastic MPB by about 50% while complete inhibition occurred at a free H_2S level of approximately 200 mg/L.

Table 2.2 Percentages of hydrogen sulfide and HS- and solubility of hydrogen sulfide as a function of pH.

pH	% of H_2S	% of HS^-	Solubility (mg/L)
4	99.9	0.1	3470
5	98.9	1.1	3510
6	90.1	9.9	3840
7	47.2	52.3	7270
7.5	22.5	77.5	15400
8	8.3	91.7	41800[a]
8.5	2.8	97.2	124000[a]
9	0.89.	99.11	290000[a]
10	0.09	99.91	

[a]Solubility, neglecting ionic strength correction, at 25° C

from ASCE (1989)

High H_2S levels in biogas are problematic for both bacteria and humans and its removal from the gas could be quite expensive. H_2S gas is corrosive to piping. The Instrumentation Society of America has classified reactive environments based on hydrogen sulfide concentrations, ranging from mild to severe. A description of the conditions and concentrations is presented in Appendix B1. Additionally, H_2S can cause death when present in concentrations above 50 ppm (Edwards et al., 1997). If H_2S containing gas is burned in heat exchangers, the products of combustion can damage the unit and severely corrode exhaust-gas heat- recovery equipment, especially if allowed to cool below the dew point (Metcalf and Eddy, Inc., 1991). Sulfides are known to precipitate the non-alkali metals in digesters, which may reduce the availability of these metal nutrients for the microorganisms (Isa et al., 1985). The absence of the metals will affect the growth of these microorganisms, which could result in reduced methane production from a digester. In human H_2S can cause eye and respiratory irritation, headaches and dizziness, unconsciousness and even death (Isa et al., 1985). Finally, H_2S can be biologically oxidized under aerobic conditions to sulfuric acid, which is corrosive to construction materials. This reaction is shown in equation 2.5 below:

$$H_2S + 2O_2 \xrightarrow[\text{bacteria}]{\text{sulfur}} H_2SO_4 \qquad \text{(Equation 2.5)}$$

3.2 Corrosion mechanism

Both oxidized and reduced forms of sulfur play a significant role in metal corrosion. Hydrogen sulfide corrodes metal by direct reaction under anaerobic conditions, or after it is biologically oxidized to sulfuric acid under aerobic conditions. The chemical reactions which occur were presented previously (equation 2.1 through 2 5).

The mechanism for sulfuric acid-based corrosion in sewers is illustrated in Figure 2.1 below.

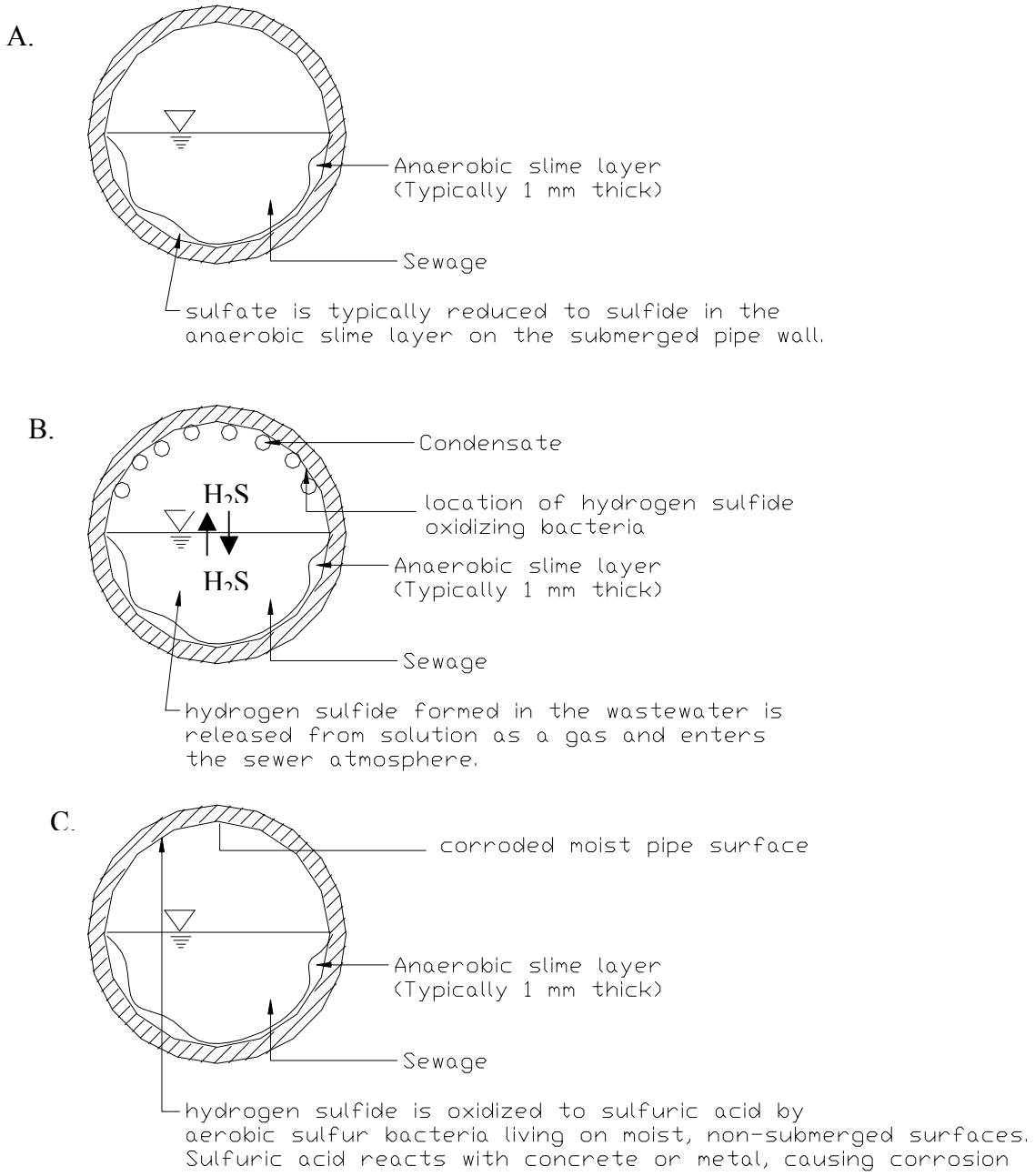


Figure 2.1 Mechanism of sulfide generation and corrosion in sewers.

From USEPA (1992).

Corrosion can also occur anaerobically when $H_2S_{(g)}$ reacts directly with metals, such as copper, iron, cadmium and silver. It was reported that electrical and instrumentation systems are particularly vulnerable to low levels of hydrogen sulfide gas (USEPA, 1992). For example, $H_2S_{(g)}$ reacts readily with copper to form copper sulfide, cadmium to form CdS, and with cast and ductile iron to form black FeS deposits (USEPA, 1992). It was reported that continued exposure of metals to $H_2S_{(g)}$ would destroy the metals (ASCE, 1989). $H_2S(g)$ reacts with iron, steel, copper and galvanized piping to form "black water" when oxygen is absent. The reaction of $H_2S(g)$ with the metals was reported to be often complex. The results from such a reaction may be readily apparent or may not appear for months before suddenly becoming severe (AWWA, 1999)

Mara and Williams, 1971; Wolzogen Kuhr and van der Vlught, 1934 reported that iron corroded in oxygen free environment mediated by SRBs with the enzyme hydrogenase. The chemical reactions which occur are presented below:

$$4Fe \rightarrow 4Fe^{2+} + 8e^- \quad \text{(Equation 2.6)}$$

$$8H_2O \rightarrow 8H^+ + 8OH^- \quad \text{(Equation 2.7)}$$

$$8H^+ + 8e^- \rightarrow 8H \quad \text{(Equation 2.8)}$$

$$SO_4^{2-} + 8H \rightarrow S^{2-} + 4H_2O \quad \text{(Equation 2.9)}$$

$$4Fe + S^{2-} + 6OH^- \rightarrow FeS + 3Fe(OH)_2 \quad \text{(Equation 2.10)}$$

This mode of corrosion is more likely to occur in anaerobic digester gas piping where oxygen is virtually absent.

3.3 Corrosion Measurement

Various methods are used to measure the extent of corrosion. The two most common methods are weight-loss and electrochemical. Visual inspection has also been used in preliminary investigations to assess potential corrosion areas in piping (AWWA, 1999). The weight-loss methods measure corrosion over a period of time. The electrochemical methods measure either instantaneous corrosion rates or rates over a period of time, depending on the method used (AWWA,1999).

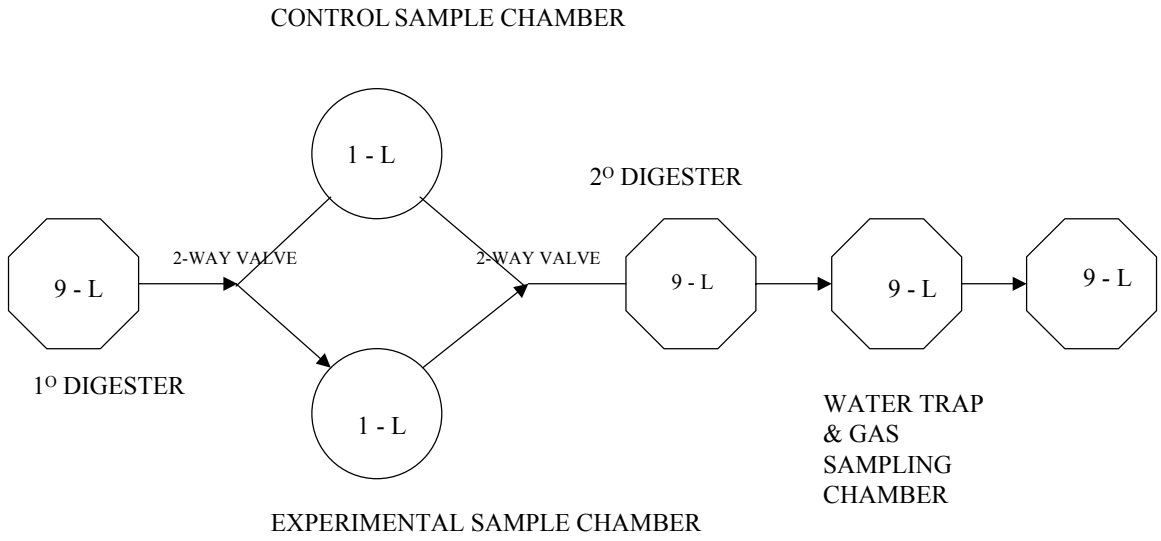
The oldest of the weight loss methods used is the flat coupon approach in which a thin rectangular piece of metal is inserted on an insulating stem into the flowing stream of a distribution network. Oxide films eventually form on the metal surface. The ultimate loss of weight due to erosion of oxide films from the coupon and also the pipe walls was used to estimate the extent of corrosion. A minimum of 120 days is recommended for the experiment. The test protocol is cumbersome due to the process of preparing the coupon surface for the test, and surface restoration upon exposure to corrosion. The method pertains only to water flowing in distribution networks. During the coupon testing process, molecule from the coupon are lost to the flowing stream; therefore corrosion can be measured based on extent of weight loss after the oxide coating is cleared from the surface and the original metal surface is restored.

CHAPTER 4.0 MATERIALS AND METHODS

4.1 Reactor setup

Two experimental set ups, I and II, were used to investigate the possible sulfide gas induced corrosion of metal iron. Experimental set up II was a modification of I. For experiment I, a configuration similar to PFRWTF two-stage anaerobic digester (AD) was used for the investigation. Design parameters for the laboratory scale ADs were based on the design and operating conditions of the PFRWTF AD. However, a batch system was used due to the small size of the laboratory set up and difficulty involved in pumping sludge continuously at the design flow rate. The laboratory-scale AD consisted of two 9000-ml reagent bottles as primary digester and secondary digester respectively (Figures 2.2 and 2.3).

(A).



(B).

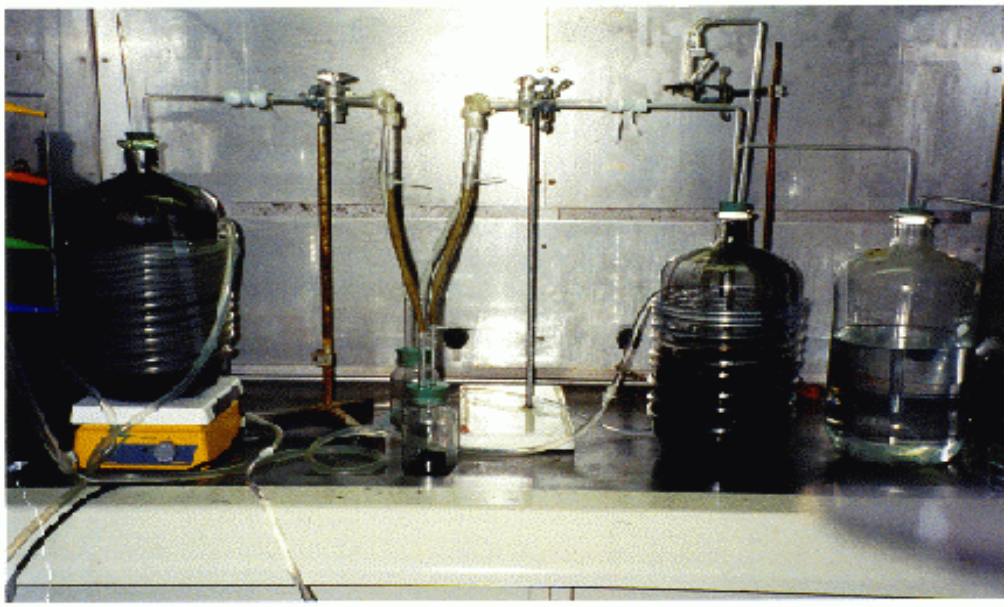
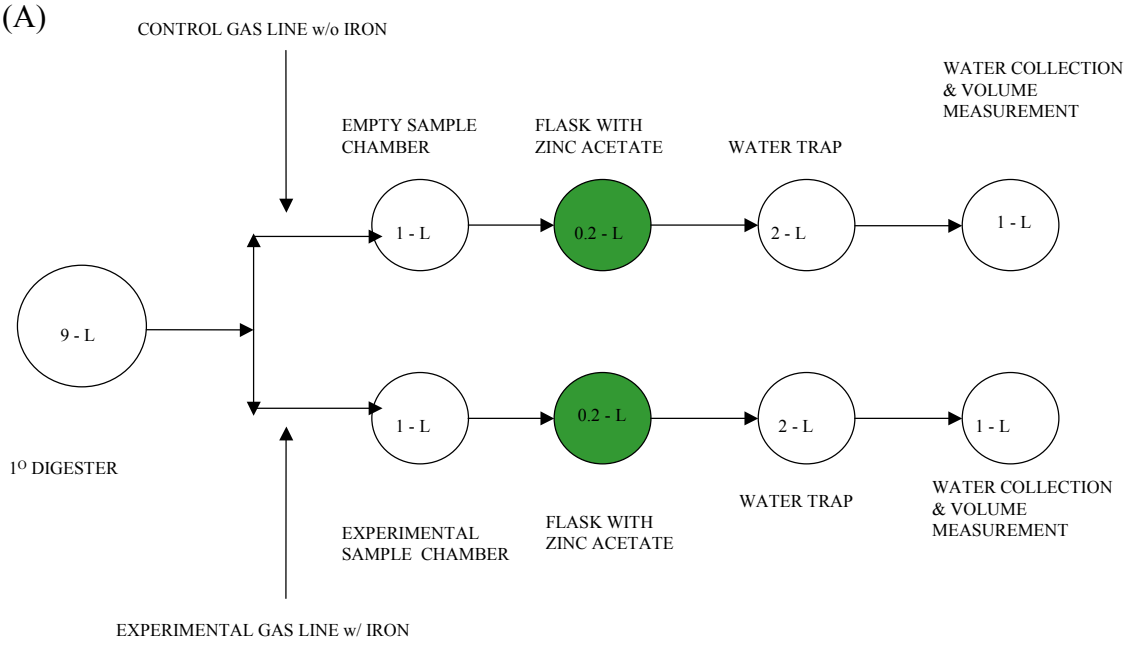


Figure 2.2. Schematic (A) and photograph (B) of experimental setup I

(A)



(B).

Figure 2.3 Schematic (A) and photograph (B) of experimental setup II

The gas produced was trapped and stored in another 9000-ml bottle, which served as a water trap as well. The water trap controlled pressure build up in the system and maintained the system at atmospheric pressure. Gas volume measurements were made by measuring the displacement of water contained in the water trap bottle. The procedure was similar to one used by Harada et al.,1994., to investigate the interaction between sulfate reducing bacteria and methane producing bacteria in upflow anaerobic sludge blanket reactors (UASBR). One quarter inch (¼ in.) OD glass tubes were specially bent to improve on properties of the glass at the joints and connected the digesters, experimental and control sample chambers and the water trap bottle (Figures 2.2B and 2.3B). Chemically resistant rubber stoppers with vents were used to connect the glass tubes to the digesters and the bottles. Two-way valves, one at the influent and the other at the effluent end, controlled gas flow through the sample chambers, which contained metal pieces (described below). The valves were turned to enable gas flow through the experimental sample chamber for the entire period of the experiment. The valves were closed during weighing of the metal pieces or maintenance of the reactors to keep the digesters anaerobic and to prevent exposing the metal pieces to the atmosphere. Weighing was performed at the same period when the digesters were being maintained. To ensure anaerobic conditions, the entire system was purged with oxygen-free 100% pure nitrogen gas for 15 minutes each day after maintaining (wasting and feeding) the digesters and weighing the samples. The digesters were maintained at a constant temperature of 35^{o} C by means of PVC tubing wrapped around the digester bottles through which water from a constant temperature water-circulating bath flowed. The content of the reactor was kept mixed using a magnetic stirring bar and mixer.

The samples for the corrosion investigation consisted of iron pipes and shavings. A piece of three-quarter (¾ in.) diameter iron pipe (same material as that used for gas piping by PFRWTF, but not previously exposed to digester gas) was threaded and cut into 2 in. lengths. Shavings from the threading of the pipe were also cut approximately into 2 in. lengths. These were degreased, washed and cleaned with detergent (dishwashing liquid) and dried in an oven at 104^{o}C to remove water. Each of the sample chambers (experimental and control) contained a known mass of iron pipe and shaving. The experimental samples were exposed to digester gas for 13 days and re-weighed. Subsequent weight measurements were taken

weekly to determine the extent of corrosion. The iron pieces were weighed by removing the pieces from the glass exposure chambers, and rapidly weighing them with a balance (XE Series, model 100A, Denver Instrument Co.) under atmospheric conditions.

The purpose of experimental setup II was to investigate if hydrogen sulfide in the digester gas reacted with the iron pipe and shaving. For this experiment, the 2-way valves in set up I were removed. The secondary digester was also removed since the location where corrosion was observed in the field indicated that it was not contributing to the corrosion of the iron. Two gas lines, experimental and control, were introduced by way of a tee (replaced the influent 2-way valve) which connected the glass tube from the primary digester. The experimental line consisted of an exposure chamber containing the iron pieces exposed to biogas previously during experiment I, a conical flask with zinc acetate solution, a water trap and a graduated bottle for measuring gas volume. The control line was identical to the experimental line except there were no iron pieces in the sample chamber. With this design, it was possible to remove and compare the hydrogen sulfide gas remaining in each line, and to determine the degree to which the gas reacted with the iron pipe and shaving. The system was constructed to ensure that pressure losses across the two lines were hydraulically similar. Both lines had PVC rubber tubing connections to enable control of gas through the lines. (See Figure 2.3).

4.2 Operation and Maintenance

The digesters were operated as batch reactors and were held at a volume of 6000 ml. They were operated at a hydraulic retention time (HRT) and solids retention time (SRT) of 15 days. Seed for the primary digester was collected from PFRWTF primary anaerobic digester. 5600 ml of the seed was initially fed into the primary digester. Feed for the digester consisted of sludge collected from PFRWTF dissolved air floatation (DAF) unit and primary clarifier. The sludges were mixed at a ratio of 3 to 1 (same as used in the field), respectively, and fed once daily into the primary digester at a rate of 400 ml/day. For experiment I, waste sludge from the primary digester was used as feed for the secondary digester. Wastage of sludge from the secondary digester began after 15 days.

Gas generated by the primary reactor and collected in the water trap was used to measure bioactivity according to gas volume, produced and gas content. Gas samples were collected using a gas tight syringe, and analyzed as described below through a septum provided in the rubber stopper to the water trap bottle.

The primary reactor was operated as a high rate anaerobic digester by completely mixing the reactor with a stirring bar and a magnetic plate on which the reactor was placed. The secondary digester was a standard anaerobic digester with no intentional mixing (similar to how the secondary unit is operated at the PFRWTF). It was mixed once a day prior to wasting to ensure that the designed HRT and SRT were maintained. The gas lines were maintained at room temperature to simulate the operating conditions at PFRWTF.

4.3 Sample preparation and testing

Routine maintenance of the reactors included measuring suspended solids concentrations and pH. The digested sludge was analyzed for mixed liquor suspended solids (MLSS) and mixed liquor volatile suspended solids (MLVSS) using Standard Methods procedure 209 E. (APHA, 1995). The pH of the digested sludge was tested daily to ensure that the reactors were operating within the desired pH range of 6.6 to 7.6.

The influent sludge sulfate concentration was determined once for each batch that was collected from the PFRWTF plant, and was assumed to remain constant for the period the sludge was used for feeding. The influent sludge was collected every four days except for the second batch, which was collected seven days after the commencement of the study. The influent sludge was stored at 4^{o}C. Procedures 427 B and D and 428 in Standard Methods (APHA, 1995) were followed to preserve and analyze the liquid phase of sludge samples from the digester for sulfide and sulfite, respectively. The sludge samples from the digester were centrifuged at 8000 rpm for 15 minutes using the Beckman, model J-21C centrifuge. The clear liquid from the centrifuged sludge was titrated daily to determine sulfide and sulfite concentrations. The influent and digested effluent sulfate concentrations were determined weekly. The clear liquid from the centrifuged sludge was stored at 4^{o}C. The liquid was filtered with 1.5 um glass microfibre filters and analyzed using an ion chromatography (Dionex 2010i with conductivity detector, gradient pump and AS40 autosampler with Peaknet 6.0 Chromeleon software) according to method 4500 SO_4^{2-}B in standard methods (APHA, 1995).

The hydrogen sulfide concentration in the digester gas was routinely determined by a titrimetric method during experiment II. Known volumes of a 1.2 M zinc acetate solution were used to remove hydrogen sulfide from the digester gas over a 24-hour period. The solution was titrated to determine the H_2S concentration per liter of zinc acetate solution. The mass of hydrogen sulfide was determined from the calculated concentration and the volume of zinc acetate used. The mass of H_2S and the total volume of gas measured in each gas line were used to determine the hydrogen sulfide gas concentration in the experimental and control digester gases.

Gas collected from the water trap bottle with a gas tight syringe gas was analyzed for methane, carbon dioxide and hydrogen sulfide using a Gow-Mac Series gas chromatograph with thermal conductivity detector (TCD). A gas mixture standard of known concentration (by moles) of methane (4.03%), carbon dioxide (5%), carbon monoxide (5.02%), hydrogen (4%), nitrogen (5.03%), and oxygen (5%) in helium was analyzed each time a gas sample was analyzed. The chromatograms obtained from the standard and the gas sample were compared to determine the methane, carbon dioxide and hydrogen concentrations by mass in the gas sample. A separate standard for hydrogen sulfide (100 ppm) was used.

The deposits found in the anaerobic digester gas piping of PFRWTF were analyzed for constituents using a Perkin Elmer PHI-100 X-ray Photoelectron Spectrosmeter (XPS). The technique measures elemental composition of a surface to depths of up to 100 angstroms. XPS can detect all elements, except for hydrogen and helium, by measuring the kinetic energies of electrons emitted from surface atoms during bombardment by X-ray radiation and computing binding energies. The electron binding energies from different elements are distinct enough to allow accurate identification not only of the element, but its oxidation state as well. The binding energy intensities can be used to quantify the relative amounts of various elements on a surface. Prior to analysis, a MLVSS test was performed on samples from PFRWTF digester A and B gas piping to determine the organic fraction of the deposit. This was to ensure that volatile organics which could affect the testing procedure or results were not excessive. Sample preparation for the test involved drying samples at 104^{o} C. The samples were then pulverized and subjected to vacuum conditions at or below approximately 5×10^{-7} torr to remove any volatile compounds. The prepared samples were stuck on a double sided clear tape which enabled the sample to be exposed to the X-rays using the equipment above (Jackson, 1999).

4.4 Precision of Corrosion Measurement

It was observed during the study that the experimental iron pieces (iron exposed to digester gas) lost weight each time they were exposed to the atmosphere for a relatively long period of time (compared to the short duration needed to weigh pieces during the experiment). The reductions in weight were more pronounced in the shavings than in the pipes. On the contrary, such observations were not found with the control iron pieces when they were also exposed to the atmosphere. In order to ensure that weight losses were minimized during weighing of the iron pieces, the exposure chambers (bottles containing the iron pieces) were immediately capped after the bottles were removed from the system. Weighing was performed rapidly on a balance (XE Series, model 100A, Denver Instrument Co.). The samples were returned to the bottles immediately after weighing,

The reaction resulting in the weight losses in the experimental samples is not yet known. However, it could be stated that errors due to weighing of the samples were minimal since the control samples did not show significant variability in their weights during the entire period of the experiment. Reiber et al., 1996 reported that ASTM, in contrast to most of its analytical standards, does not estimate the precision of measurement in corrosion-rate measurement using the machined-nippled test described in ASTM D2688-82. Rather, ASTM recommends that precision in measurements be considered a function of the individual system. They also mentioned that the ASTM standard rarely mentioned confidence intervals or other quantitative indicators of precision with measuring corrosion rate. It is therefore believed that measures taken regarding the weighing of the iron pieces resulted in reliable data within the margin of acceptable experimental errors.

CHAPTER 5.0 RESULTS AND DISCUSSION

5.1 Laboratory Experiment I Reactor Performance

The experiment I configuration was operated for 30 days. During this period, pH, MLSS, MLVSS, gas production rate and methane/CO_2 production were monitored to ensure reactor performance. The average methane and carbon dioxide production yields were 25.5 % (SD = 13.6, N = 20) and 31.8 % (SD = 11.2 and N = 20), respectively. The rest of the gas was composed mainly of nitrogen, which was used to purge the reactors daily. From Figure 2.4, it can be seen that initial (8 to 11 days after the start of the study) concentrations of methane were around 50 %, which was relatively higher than the average concentration of carbon dioxide (25%). However, a decrease in methane concentration and increase in carbon dioxide concentration were observed after day 11. The carbon dioxide and methane concentrations remained fairly constant at about 26 % and 24 %, respectively, after 23 days of the study. Low levels of methane (10%) measured between the 17^{th} and 21^{st} days could be attributed to low concentrations of solid in the influent sludge collected from the treatment plant, or the possibility that SRB activities increased relative to MPB. Carbon dioxide concentration during this period averaged about 50%. Digester gas production was about 250 ml per hour during experiment I when both the primary and secondary digesters were operated. The percentage of the methane and carbon dioxide gas components obtained in the laboratory were 50% lower for methane than that measured in the field, while CO_2 levels measured in the laboratory were comparable to the field data. Field values of methane and CO_2 were about 56% and 36%, respectively, in 1998.

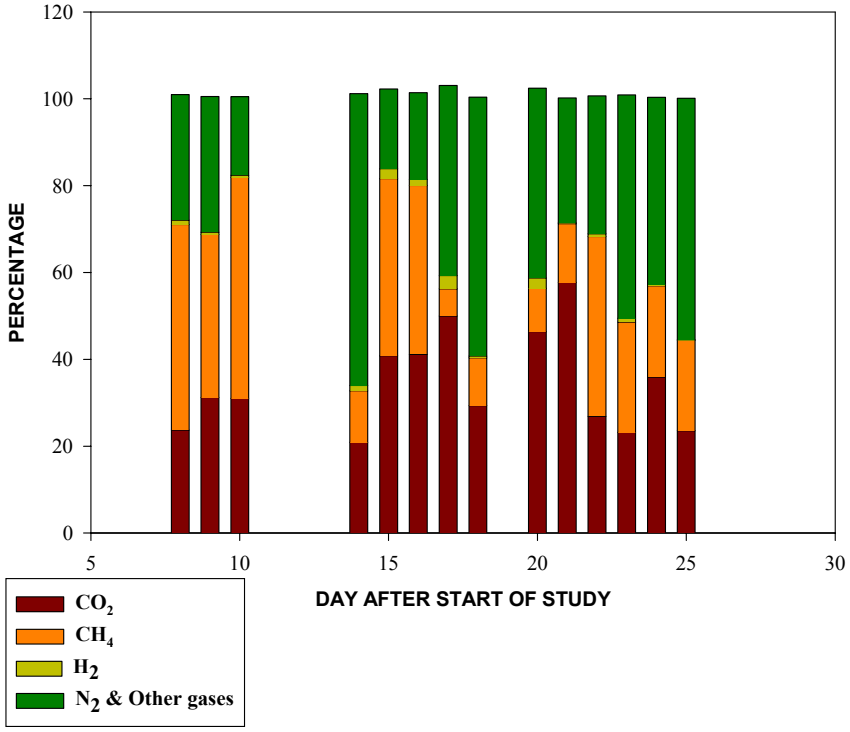


Figure 2.4 Percentage composition of digester gas components.

The MLSS and MLVSS for the primary digester sludge were determined for first 30 days of the study. MLSS and MLVSS averaged 38,400 mg/L (SD = 7,329, N = 25) and 25,800 mg/L (SD = 5391, N = 25), respectively (see Figure 2.5). Influent MLSS was determined for only one of the influent sludge batches, and was about 50,000 mg/L. Assuming this influent concentration was typical, the laboratory primary digester was achieving approximately 24 % of suspended solids. Raw data are presented in Appendix B2.

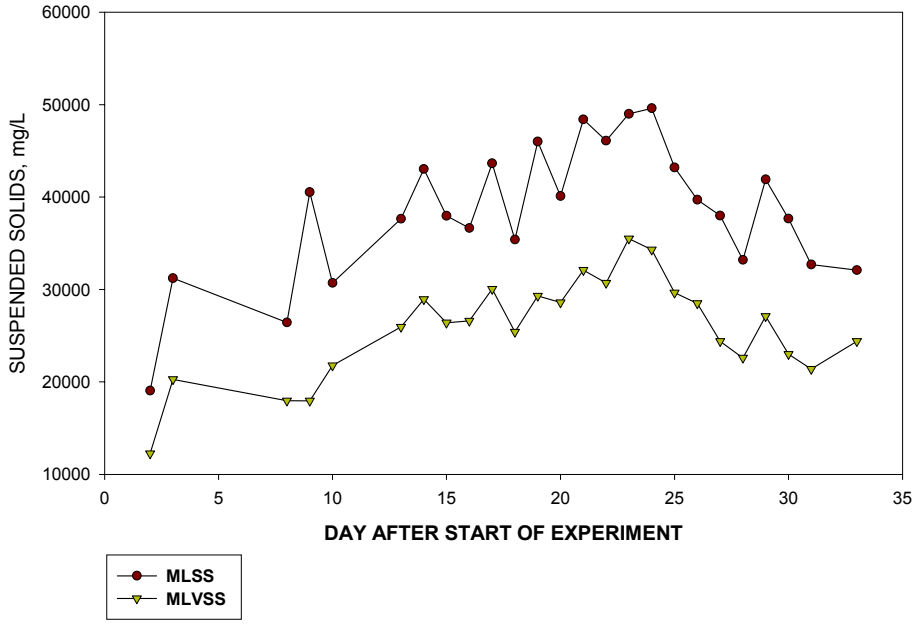


Figure 2.5 MLSS / MLVSS trends in primary digester effluent.

The primary digester effluent pH values, presented in Figure 2.6, were generally within the recommended range for efficient performance of anaerobic digesters, which is between pH 6.6 and 7.6. No pH correcting chemicals were added to the digesters during the entire experiment. Monitoring pH was essential to ensure that the reactors were operating efficiently under steady state conditions.

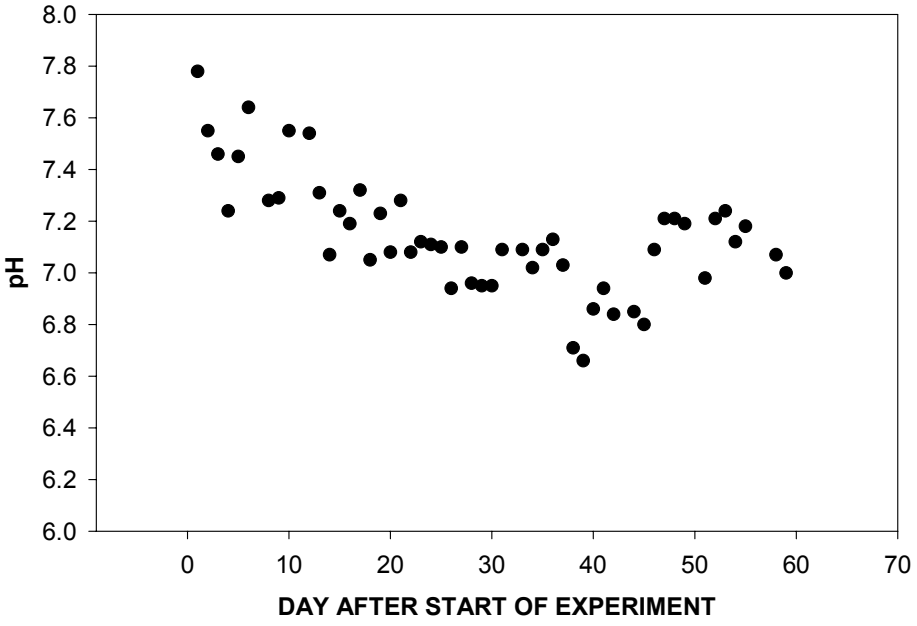


Figure 2.6 pH trends in primary digester effluent.

5.2 Corrosion

Significant weight changes were observed in both the experimental metal pipe (EMP) and shaving (EMS). The trends for the EMP were more defined than for the EMS (See Figures 2.10 and 2.11). However, an overall increase in weight of 5.7 % in the EMS was determined. The overall change in EMP weight was 1.074 %. Comparable changes in the pipe and shavings in the control exposure chamber, which was not exposed to digester gases, were 0.002% and –0.155%, respectively. Among the factors influencing the overall weight changes observed with in the samples are the size and, more importantly, the surface area of each sample. However, these characteristics were not determined due to the nature of the samples. The EMP had a regular shape (hollow and cylindrical) with inner and outer surfaces. The outer surface was threaded whilst the inner surface had a galvanized finish. Any reaction in the metal was more likely to occur on the unprotected threaded surface rather than on the galvanized surface. The EMS was relatively small with an irregular (curly) shape. However, both sides of the shavings were exposed ungalvanized iron surfaces and, therefore, reactions were possibly occurring at a faster rate than would have occurred on a galvanized iron surface.

The overall changes in weight of the metal pieces were determined to assess whether reactions were occurring between the iron and biogas, and to compare the relative rate or extent of reaction for each iron sample. Visual inspection of the metals a few days (2 days) after exposure to the gas revealed a readily detectable black film on the EMP and shaving EMS surfaces. Figure 2.7 shows the black film on the experimental pieces 13 days after the start of experiment. There weight increased by 1 % within 13 days after exposure to digester gas while there was no significant change in weight of the EMS (Figure 2.7). In contrast, there were no observed weight and visual changes in the control metal pipe (CMP) and shaving (CMS) after 13 days (Figures 2.7). The rate of change in weight of EMP decreased with time. The final weight of EMP after 62 days was almost the same as that determined 13 days after start of the experiment. It could be inferred from the observation that reaction of hydrogen sulfide gas with the metal surface was almost complete within the first 2 weeks of the experiment.

The trend for the shaving was quite different. Although changes were observed in EMP 13 days after the beginning of experiment I, no changes occurred in EMS by that time. Increases in weight of EMS were observed after 27 days. However, this weight was lost when the shaving was removed from the system and exposed to the atmosphere at the end of phase one. Reintroduction (beginning of phase II) and exposure of EMS to the digester gas followed an increasing trend in weight with decreasing trend in the percentage weight change, as was observed with EMP (Figure 2.11).

Hydrogen sulfide gas is suspected to be the cause of the corrosion observed. The black film observed in Figure 2.7 could be due to the formation of ferrous sulfide, FeS which is black in color and could have formed rapidly upon exposure of ferrous iron to hydrogen sulfide gas. Mara and Williams (1971) reported of formation of loose black gelatinous film on iron exposed to anaerobic continuous culture fed with lactate-sulfate-mineral salts.

A.

EXPERIMENTAL CONTROL

B.

EXPERIMENTAL CONTROL

Figure 2.7 Appearance of experimental and control metals 13 days (A) and 62 (B) days after the start of the experiment.

Over 90% reduction in influent soluble sulfate concentration was observed in the primary digester throughout the study Figure 2.8. The reduction in sulfate concentration was possibly due to SRB out competing MPB in the anaerobic process by accepting electrons to degrade organic material. This results in the formation of hydrogen sulfide. In an anaerobic digester,

activities of SRB competing with methane producing bacteria, (MPB) have been reported (Isa et al., 1985). The gaseous component of hydrogen sulfide mixes with other components of digester gas, namely carbon dioxide and methane and reacts with the exposed metal surfaces. The reaction of metals with hydrogen sulfide occurs either due to direct reaction of hydrogen sulfide gas with metals (copper, iron, silver, and cadmium) or through formation of sulfuric acid which corrodes metals, or both, discussed previously. The direct reaction of hydrogen sulfide gas is more likely to be the mode of corrosion of EMP and EMS, since the metals were exposed to digester gas in the absence of oxygen Therefore it is concluded that the black film observed on the metal surfaces were formed as a direct reaction of H_2S gas with the iron pipe and shaving. It was reported that continued exposure of metals to hydrogen sulfide gas would destroy the metals (AWWA, 1999). This was observed in EMS, which turned brittle after 62 days of exposure to the gas (Figure 2.7B). The brittle nature might have resulted in the loss of material during handling in the weighing process and might explain why EMS had sometimes weighed less than its initial weight.

It was determined through experiment II that the concentration of the hydrogen sulfide gas downstream of where the gas was exposed to the EMP and EMS was consistently less than in the control when the gas was not exposed to any metal (Figure 2.9). This result indicates that a reaction occurred between the metal pieces and the hydrogen sulfide gas, and provides supporting evidence for the notion that ferrous sulfide-like precipitates were forming on the surfaces of the metal pieces.

It was also observed that the overall percentage change in weight of the metal pipe was almost the same as that when it was initially exposed to the gas for 13 days (See Figure 2.10). Loss of weight in the metal pieces was observed each time the metals were exposed to the atmosphere for few days (day 13 to day 17 and day 34 to day 41 of start of experiment). It was not very clear why EMP and EMS would lose weight when they were exposed to the atmosphere. However, each time that the metal pieces were reintroduced and exposed to the digester gas, weight increase trends were observed.

Examination of Figure 2.12 shows that, there was a consistent decreasing trend in percentage change in weight for both EMP and EMS and the rate of corrosion reduced approximately to zero 6 weeks after the start of the experiment. Mara and Williams (1971) reported that initial corrosion rate of iron going sulfur corrosion by SRBs gradually reduced to zero after 3 to 6 weeks. The decreasing trends in the rate of weight increase of the corroding metals should be expected since the amount of metal surface available for reaction with hydrogen sulfide gas would be used up with time. This result suggests that corrosion happens quickly in the pipes and then slows to a stable but slow rate of corrosion with time.

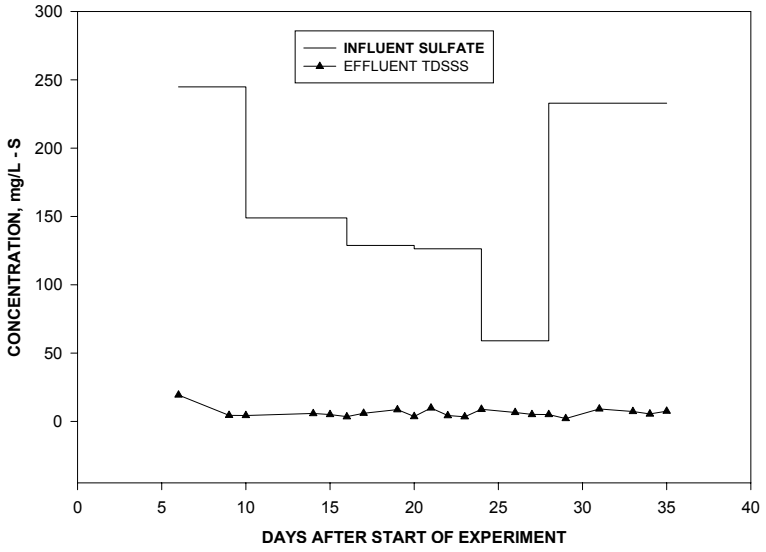


Figure 2.8 Influent sulfate concentration and effluent total dissolved sulfate, sulfite and sulfide (TDSSS).

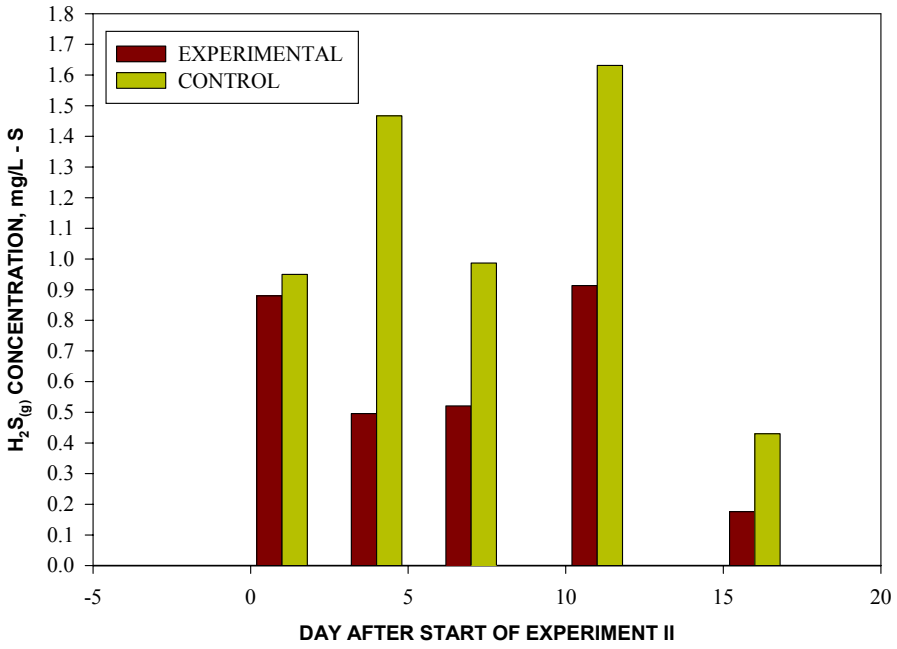


Figure 2.9 Hydrogen sulfide gas concentrations in experimental and control lines during experiment II.

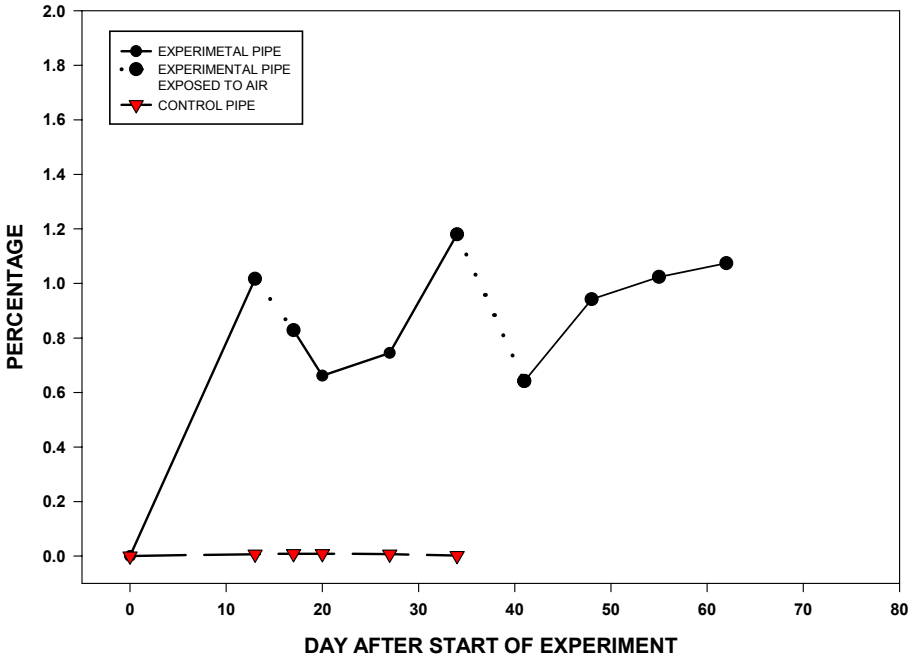


Figure 2.10 Cumulative percentage change in weight of pipes during experiments I and II

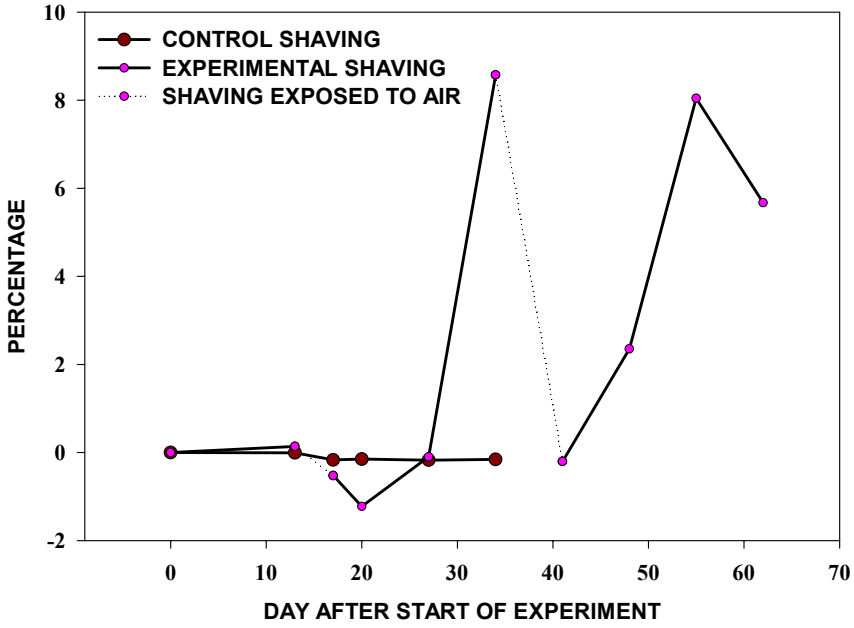


Figure 2.11 Cumulative percentage change in weight of iron shavings during experiments I and II

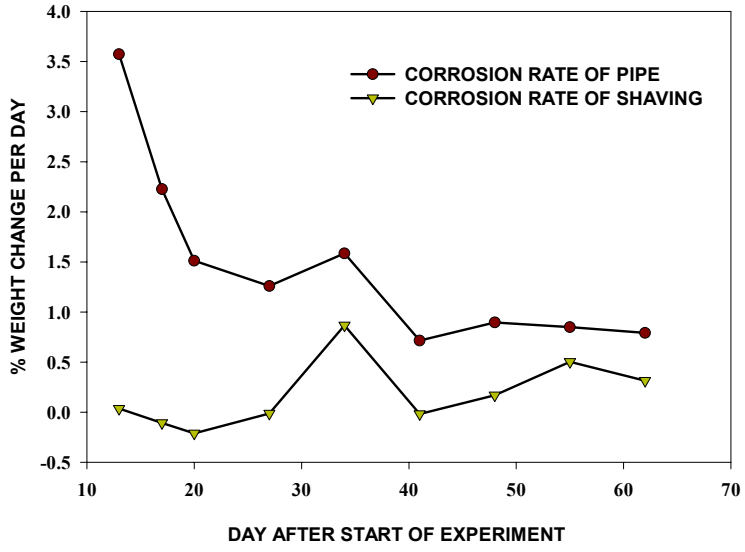


Figure 2.12 Rate of Corrosion in experimental metal iron and shaving.

An attempt was made to conduct a mass balance on sulfur in the system. Sources of sulfur in the influent were mainly in the form of dissolved sulfate with some sulfur present in the organic compounds in the sludge. In anaerobic digesters, sulfates are used by sulfate reducing bacteria (SRB) as electron acceptors. It is also possible for sulfide oxidizing bacteria to convert sulfide to sulfur or sulfate with ferric ion acting as an electron acceptor under anaerobic conditions (Brock et al., 1984). Sulfide and sulfite were virtually absent from the influent. However, since dissolved sulfide was not detected to a significant degree at any point in the system, this reaction would not be expected to occur to a significant extent in the reactor. Average influent sulfate concentrations were very high (about 245 mg/L as S), while typical effluent sulfate concentrations were less than 3 mg/L sulfate as S (See Figure 2.8). This suggests that sulfate ions are reduced to sulfide by SRBs. The sulfide formed was presumed largely insoluble since low effluent total dissolved sulfate, sulfite and sulfite (TDSSS) and low H_2S gas concentrations were measured. In addition, dissolved sulfide is slightly insoluble with metals, including iron (Table 2.3). Due to the difficulty involved with analyzing for particulate sulfur, only dissolved gaseous sulfur forms were monitored. During the analysis of the digester gas the GC barely detected any hydrogen sulfide gas when compared with a standard of 100 ppm. Therefore $H_2S_{(g)}$ concentration in the biogas were far

below 100 ppm. The average $H_2S_{(g)}$ concentration in PFRWTF biogas between June and October of 1998 was about 730 ppm with the minimum concentration being 150 ppm (Olver Inc., 2000). The $H_2S_{(g)}$ concentration in the field digester gas has shown an increasing trend lately (Apendix B3). The observations suggest that the $H_2S_{(g)}$ generating potential was much greater in the field than in the laboratory system, and the corresponding potential for gas pipe corrosion was greater as well.

During experiment II, hydrogen sulfide gas concentrations were less than 2.0 $mg/L_{(g)}$ as S as shown in Figure 2.9. This constitutes approximately 0.82% of sulfur added on a mass basis during experiment II. Figure 2.8 shows that over 90% of the influent sulfate was not detected as sulfate or reduced sulfide in the soluble phase. Combined with the fact that H_2S in the gas phase was low, these results strongly suggest that most of the sulfur ended up in the insoluble particulate phase in the effluent sludge, most likely as sulfide. Nickel, cadmium, copper and iron have been reported to be present in the wastewater of PFRWTF (Olver Inc., 1995). Sulfide removal might have occurred by precipitation with these metals in the digester. The reported solubilities of sulfide in association with these metal sulfides and their solubility products at 18^oC are shown in Table 2.5 and indicate that these metals will precipitate readily with sulfide. It is possible these metal ions are protecting the digesters from inhibitory effects of dissolved hydrogen sulfide in the field. Their absence from the influent sludge could affect the performance of the digesters. It would therefore be imperative to monitor the concentrations of the metal ions in the soluble phase of the digester sludge, since some of the metals are required for growth of bacteria as well.

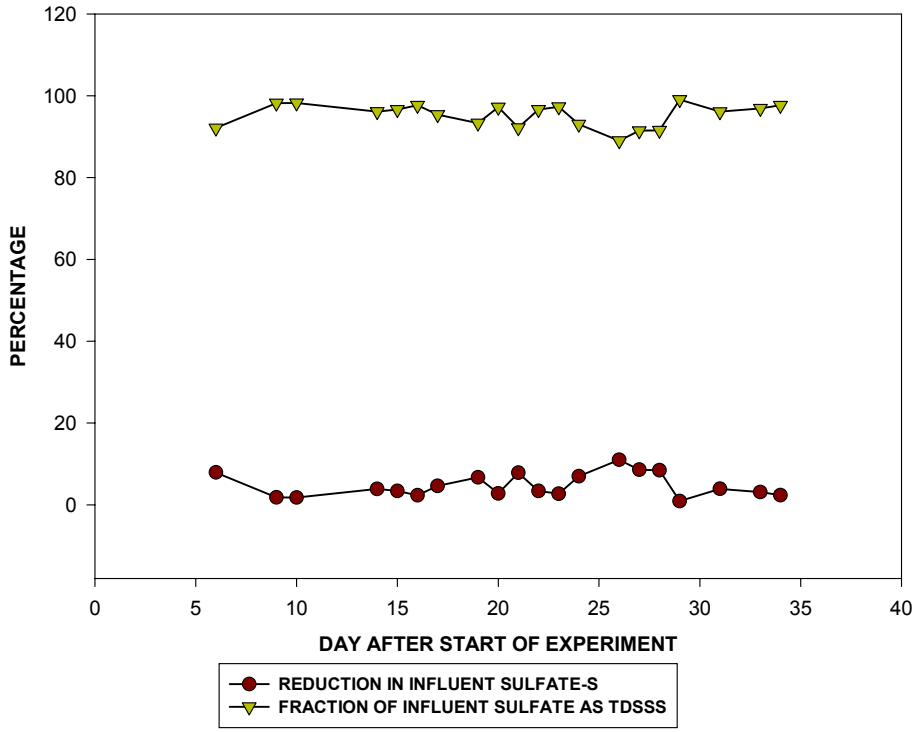


Figure 2.13 The percentage reduction in influent sludge sulfate concentration and total dissolved sulfate, sulfite and sulfide (TDSSS) as a fraction of influent sulfate concentration

Table 2.3 Solubilty Data for Selected Metal Sulfides.
Solubility Data for Selected Metal Sulfides.

Metal sulfide	Solubility product (18° C to 25°C)	Sulfide concentration, in moles/liter
Mangamous sulfide	1.40E-15	3.70E-08
Ferrous sulfide	3.70E-19	6.10E-10
Zinc sulfide	1.20E-23	3.50E-12
Nickel sulfide	1.40E-24	1.20E-12
Stannous sulfide	1.00E-25	3.20E-13
Cobalt sulfide	3.00E-26	1.70E-13
Lead sulfide	3.40E-28	1.80E-14
Cadmium sulfide	3.60E-29	6.00E-15
Silver sulfide	1.60E-49	3.40E-17
Copper sulfide	8.50E-45	9.20E-23
Mercuric sulfide	2.00E-49	4.50E-25

Culled from ASCE, Design manual No. 69.

It was reported that sulfide precipitation is especially important in industrial wastewater treatment and in anaerobic sludge digesters where metals have been concentrated in the sludge. Of the metals shown in Table 2.3, only iron has been used as a chemical additive to intentionally remove sulfide, where ferrous ions precipitate sulfide as ferrous sulfide (ASCE, 1989). Ferric ions were reported to precipitate sulfide through reduction of the iron to the ferrous form when coupled with sulfide oxidation to sulfur, shown in equation (2.11) and (2.12) below. The ferrous ion is then available for direct precipitation with sulfide. It was reported that because the solubility product of ferrous sulfide is 3.7×10^{-19} at 18° C, equation (2.12) is very effective in precipitating sulfides.

$$2Fe^{+++} + S^{2-} \rightarrow 2Fe^{++} + S^{o} \qquad \text{(Equation 2.11)}$$

$$Fe^{++} + S^{2-} \rightarrow FeS_{(s)} \downarrow \qquad \text{(Equation 2.12)}$$

The theoretical reaction to precipitate sulfide with ferrous and ferric ions is assumed to take the form shown in equation (2.13) below:

$$Fe^{2+} + 2Fe^{3+} + 4HS^{-} \rightarrow Fe_3S_4 + 4H^{+} \quad \text{(Equation 2.13)}$$

Improved sulfide removal by addition of both ferric and ferrous salts results in dissolved sulfide removal to levels of 0.2 mg/L (ASCE, 1989). Formations of other insoluble iron and sulfur forms were reported to be possible. These include ferric sulfide (Fe_2S_3), smythite (Fe_3S_4) and marcasite (FeS_2) (ASCE, 1989).

5.3 Field Analysis of solid precipitate from Digester Gas Pipe at PFRWTF

The results of the XPS test performed on the deposits found in the PFRWTF anaerobic gas piping revealed the composition of the black precipitate. The test indicated that the following elements were present at the given percent of mass: 24.19% carbon, 58.12% oxygen, 2.69% sulfur and 15.00% Iron (ferrous and ferric) in digester A deposit. The deposit in digester B had the following components and percent mass: 25.42 % carbon, 56.61% oxygen, 3.6% sulfur and 14.37% iron. It was possible that the oxygen was picked up during sampling of the deposit for testing. This is because the high percentage could not possibly have been part of the deposit under anaerobic conditions in the piping. Details of the results of the experiment are presented in Appendix B4.

The organic fraction determined using procedures in Standard Methods (APHA et al., 1995) yielded an average of 25% by mass for both digester deposits. These compared very well with results of the XPS technique and confirm the carbon present in the deposit was organic carbon. Further analysis would be required to determine the various forms of compounds in which the elements exist. The ratio of iron and sulfur in the different iron sulfur compounds (Fe_2S_3, Fe_3S_4, FeS_2 and FeS) indicate that the amount of iron in the field deposit, assuming that only iron sulfide was originally formed in the deposit and sulfur was limiting, should not exceed 6 %. Iron in the deposit was 15 % suggesting that sulfide iron was not the only iron compound present in the deposit but ferrous hydroxide, which is formed as product with iron

sulfide via equations (2.6) through (2.10) presented earlier (Mara and Williams, 1971; Wolzogen Kuhr and van der Vlught, 1934) could be present and could contribute to oxygen in the deposit.

CHAPTER 6.0 SUMMARY AND CONCLUSIONS

- Corrosion of iron piping could be attributed to the direct reaction of hydrogen sulfide gas in the digester gas. Possible iron-sulfur corrosion is evident from the field analysis of the black deposit found in the piping. However, the exact forms of the compound could not be easily determined from the test due to the numerous forms of iron (both ferrous and ferric) compounds that exist.
- Over 90% of sulfide produced as a result of sulfate reduction possibly remains in the digester sludge as particulate. It is possible that the presence of ferrous ions in the sludge is protecting methanogens from the toxic or inhibitory effect of free dissolved hydrogen sulfide by reacting with the sulfide produced in the digesters and removing it from solution.
- Corrosion of metal iron due to hydrogen sulfide gas is rapid at first, then slows to a low constant corrosion rate. Results from experiment I suggested that corrosion of the exposed surface of the metal pipe was almost complete within 13 days.

CHAPTER 7.0 RECOMMENDATIONS

- It is recommended that a detailed study on corrosion be carried out in the field to estimate corrosion rates in the gas piping itself and to analyze any deposition, which might be formed on the corroding coupons that will be used.
- Iron salts in the form of ferrous chloride could be used to scavenge hydrogen sulfide especially when the digester is kept above pH 7. At pH above 7, the dominant sulfide species is HS^- which is less toxic.
- It is recommended that the particulate biosolids in the effluent from the first digester be analyzed for sulfide to determine whether or not large quantities of it remain in the sludge.
- Iron filings or chippings can be introduced in the digester gas line to remove hydrogen sulfide gas that may be present in the digester gas.

CHAPTER 8.0 APPENDICES

Appendix B1 Classification of reactive environment based on $H_2S_{(g)}$ concentration.

INSTRUMENTATION SOCIETY OF AMERICA CLASSIFICATION OF REACTIVE ENVIRONMENTS BASED ON $H_2S_{(g)}$ CONCENTRATIONS

Severity Level G1($H_2S_{(g)}$ < 3ppb)
Mild – Environment sufficiently well controlled such that corrosion is not a factor in determining equipment reliability.

Severity Level G2 (H_2S) < 10 ppb)
Moderate – An environment in which the effects of corrosion are measurable and may be a factor in determining equipment reliability.

Severity Level G3 (H_2S) < 50 ppb)
Harsh – An environment in which there is high probability that corrosive attack will occur. These harsh levels should prompt further evaluation resulting in environmental controls or specially designed and packaged equipment.

Severity Level GX (H_2S) > 50 ppb)
Severe – An environment in which only specially designed and packaged equipment would be expected to survive. Specifications for equipment in this class are a matter of negotiation between user and supplier.

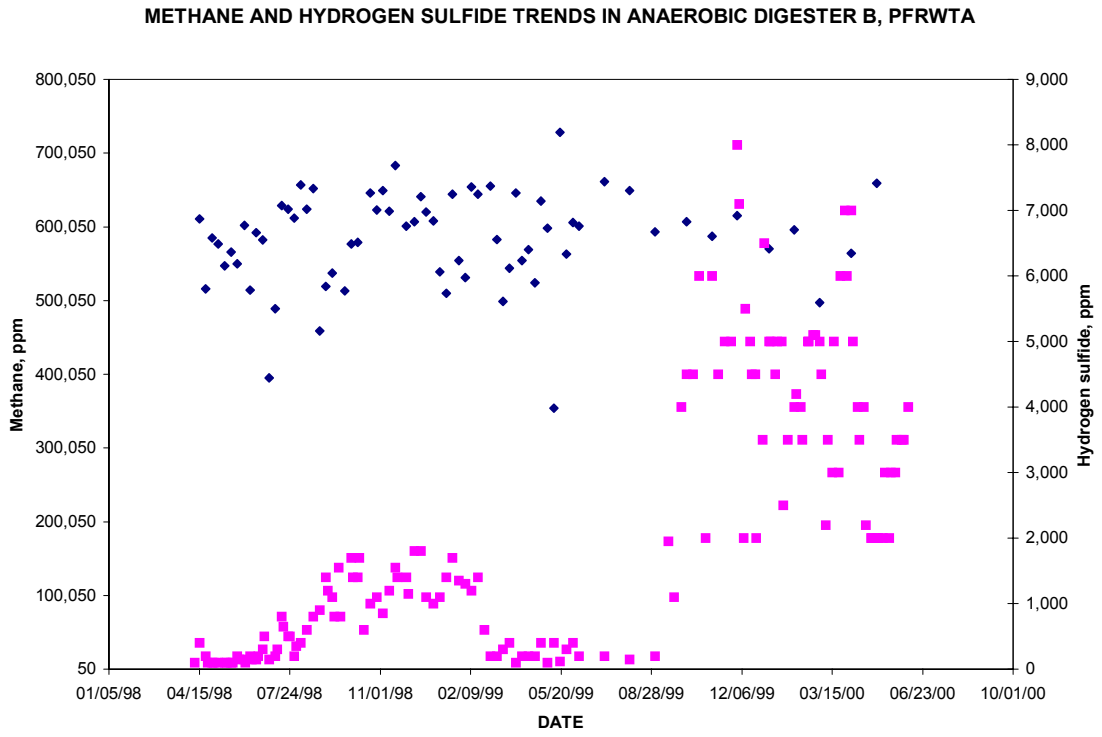
(From USEPA, Doc.EP 2.2 D 48, 1992)

Appendix B2. MLSS/ MLVSS Data.

TSS, VSS AND TSS/VSS RATIO DATA OF PRIMARY DIGESTER EFFLUENT

Date	Day of study	TSS mg/L	VSS mg/L	TSS/VSS
7/19/2000	1	-	-	-
7/20/2000	2	19050	12250	1.56
7/21/2000	3	31225	20275	1.54
7/26/2000	8	26433	17966	1.47
7/27/2000	9	40533	17950	2.26
7/28/2000	10	30716	21767	1.41
7/31/2000	13	37650	25950	1.45
8/1/2000	14	43,033	28,933	1.49
8/2/2000	15	37,967	26400	1.44
8/3/2000	16	36650	26600	1.38
8/4/2000	17	43650	30050	1.45
8/5/2000	18	35400	25400	1.39
8/6/2000	19	46000	29300	1.57
8/7/2000	20	40100	28600	1.40
8/8/2000	21	48400	32100	1.51
8/9/2000	22	46100	30700	1.50
8/10/2000	23	49000	35500	1.38
8/11/2000	24	49600	34300	1.45
8/12/2000	25	43200	29650	1.46
8/13/2000	26	39700	28500	1.39
8/14/2000	27	38000	24400	1.56
8/15/2000	28	33200	22600	1.47
8/16/2000	29	41900	27100	1.55
8/17/2000	30	37,667	23000	1.64
8/18/2000	31	32700	21400	1.53
8/20/2000	33	32100	24400	1.32

Appendix B3. Field gas composition trends



Appendix B4. Primary Digester Influent , Effluent Sulfate, Sulfite and Sulfide Data.

CONCENTRATION OF INFLUENT SULFATE AND EFFLUENT SULFATE, SULFITE AND SULFIDE A

		Influent	Effluent supernatant					
Date	Day of study	sulfate in mg/L as S	sulfate, mg/L as S	sulfite, mg/L as S	sulfide, mg/L as S	TDSSS in effluent in mg/L as S	TDSSS in effluent as a fraction of influent SO4- - S	% reduction in influent SO4- - S
7/24/2000	6	244.86	17.04	2.3	0	19.32	7.89	92.11
7/27/2000	9	244.86	1.58	2.9	0	4.43	1.81	98.19
7/28/2000	10	244.86	2.49	1.2	0.64	4.33	1.77	98.23
8/1/2000	14	148.87	2.48	1.32	1.973	5.78	3.88	96.12
8/2/2000	15	148.87	1.35	1.68	1.973	5.00	3.36	96.64
8/3/2000	16	148.87	1.83	1.00	0.64	3.47	2.33	97.67
8/4/2000	17	128.8	1.96	2.00	1.973	5.94	4.61	95.39
8/6/2000	19	128.8	1.87	2.00	4.736	8.61	6.68	93.32
8/7/2000	20	128.8	1.45	1.50	0.64	3.59	2.79	97.21
8/8/2000	21	126.3	1.31	2.50	6.06	9.87	7.81	92.19
8/9/2000	22	126.3	1.25	3.00	0	4.25	3.36	96.64
8/10/2000	23	126.3	1.26	1.50	0.64	3.40	2.69	97.31
8/11/2000	24	126.3	1.58	2.50	4.736	8.82	6.98	93.02
8/13/2000	26	58.99	1.29	2.52	2.688	6.50	11.01	88.99
8/14/2000	27	58.99	1.36	1.00	2.688	5.05	8.56	91.44
8/15/2000	28	58.99	1.30	1.00	2.688	4.99	8.46	91.54
8/16/2000	29	232.95	1.14	1.00	0	2.14	0.92	99.08
8/18/2000	31	232.95	1.91	3.48	3.712	9.11	3.91	96.09
8/20/2000	33	232.95	1.92	1.00	4.326	7.25	3.11	96.89
8/21/2000	34	232.95	1.74	0.20	3.507	5.45	2.34	97.66
8/22/2000	35	232.95	2.25	1.32	3.917	7.49	3.21	96.79

Appendix B5. pH Data.

pH

Date	DAY OF STUDY	pH	Date	DAY OF STUDY	pH
7/19/2000	1	7.78	8/19/2000	32	-
7/20/2000	2	7.55	8/20/2000	33	7.09
7/21/2000	3	7.46	8/21/2000	34	7.02
7/22/2000	4	7.24	8/22/2000	35	7.09
7/23/2000	5	7.45	8/23/2000	36	7.13
7/24/2000	6	7.64	8/24/2000	37	7.03
7/25/2000	7	-	8/25/2000	38	6.71
7/26/2000	8	7.28	8/26/2000	39	6.66
7/27/2000	9	7.29	8/27/2000	40	6.86
7/28/2000	10	7.55	8/28/2000	41	6.94
7/29/2000	11	-	8/29/2000	42	6.84
7/30/2000	12	7.54	8/30/2000	43	-
7/31/2000	13	7.31	8/31/2000	44	6.85
8/1/2000	14	7.07	9/1/2000	45	6.8
8/2/2000	15	7.24	9/2/2000	46	7.09
8/3/2000	16	7.19	9/3/2000	47	7.21
8/4/2000	17	7.32	9/4/2000	48	7.21
8/5/2000	18	7.05	9/5/2000	49	7.19
8/6/2000	19	7.23	9/6/2000	50	-
8/7/2000	20	7.08	9/7/2000	51	6.98
8/8/2000	21	7.28	9/8/2000	52	7.21
8/9/2000	22	7.08	9/9/2000	53	7.24
8/10/2000	23	7.12	9/10/2000	54	7.12
8/11/2000	24	7.11	9/11/2000	55	7.18
8/12/2000	25	7.1	9/12/2000	56	-
8/13/2000	26	6.94	9/13/2000	57	-
8/14/2000	27	7.1	9/14/2000	58	7.07
8/15/2000	28	6.96	9/15/2000	59	7.00
8/16/2000	29	6.95	9/16/2000	60	-
8/17/2000	30	6.95	9/17/2000	61	-
8/18/2000	31	7.09	9/18/2000	62	-

Appendix B6. Corrosion Data.

CORROSION OF IRON PIPE

Date	Time, days	IRON PIPE				
		Expt. Wt. (g)	Control Wt. (g)	Expt. - % weight change relative to initial wt.	Control - % weight change relative to initial wt.	Rate of corrosion- Experimental pipe
7/18/2000	0	45.6626	46.3906	0	0	-
7/31/2000	13	46.1272	46.3937	1.017	0.007	3.57
8/4/2000	17	46.041	46.3950	0.829	0.009	2.23
8/7/2000	20	45.9649	46.3948	0.662	0.009	1.51
8/14/2000	27	46.0026	46.3939	0.745	0.007	1.26
8/21/2000	34	46.2016	46.3916	1.180	0.002	1.59
8/28/2000	41	45.9556	-	0.642	-	0.71
9/4/2000	48	46.0929	-	0.942	-	0.90
9/11/2000	55	46.1302	-	1.024	-	0.85
9/18/2000	62	46.1530	-	1.074	-	0.79

CORROSION OF IRON SHAVINGS

Date	Time, days	IRON SHAVINGS				
		Expt. Wt. (g)	Control Wt. (g)	Experimental - % weight change relative to initial wt.	Control - % weight change relative to initial wt.	Rate of corrosion - Experimental shaving
7/18/2000	0	3.4411	3.4184	0	0	-
7/31/2000	13	3.4458	3.4182	0.137	-0.006	0.04
8/4/2000	17	3.4231	3.4127	-0.523	-0.167	-0.11
8/7/2000	20	3.399	3.4133	-1.223	-0.149	-0.21
8/14/2000	27	3.4377	3.4125	-0.099	-0.173	-0.01
8/21/2000	34	3.7362	3.4131	8.576	-0.155	0.87
8/28/2000	41	3.4341	-	-0.203	-	-0.02
9/4/2000	48	3.5221	-	2.354	-	0.17
9/11/2000	55	3.718	-	8.047	-	0.50
9/18/2000	62	3.6363	-	5.673	-	0.31

OVERALL PERCENTAGE CHANGE IN WEIGHT OF IRON

Description	PIPE		SHAVINGS	
	Experimental - % Change in weight.	Control - % Change in weight.	Experimental - % Change in weight.	Control - % Change in weight.
Overall percentage change in weight	1.074	0.002	5.673	-0.155

CHAPTER 9.0 REFERENCES:

1. APHA-AWWA-WPCF., 1995. Standard methods for the examination of water and wastewater. American Public Health Association, 1015 fifteenth street NW, Washington, DC 20005.
2. ASCE , 1989. Sulfide in wastewater collection and treatment systems, manuals and reports on engineering practice No. 69.
3. AWWA, Water quality and treatment, a handbook of community water supplies. (5^{th} ed.). McGraw-Hill, Inc., New York St. Louis San Francisco Auckland Bogata Tokyo Toronto.
4. Borgerding, J., 1972. "Phosphate deposits in digestion systems." JWPCF, 44.
5. Brock, T. D., D. W. Smith, Madigan M. T. (1984). Biology of Microorganisms, 4th Ed. Prentice-Hall, Inc. Englewood Cliff, NJ.
6. Gudmundsdottir, S. (1993). "Struvite formation in anaerobically digested sludge at the Renton wastewater treatment plant," MS thesis, University of Washington.
7. Harada, H., S. Uemura, K. Momonoi. 1994. Interaction between sulfate reducing bacteria amd methane producing bacteria iin UASB reactors fed with low strength wastes containing different levels of sulfate. Water Research 28: (2) pp. 355-367.
8. Ingvorsen, K., A. J. B. Zehnder, nad B. B. Jorgensen. 1984. Kinetics of sulfate and acetic uptake by *Desulfobacter postgatei.* Appl. Environ. Microbiol. 47:403-408.
9. Isa Z., Grusenmeyer S., and Versatraete W. 1985. Sulfate reduction relative to methane in high-rate anaerobic digestion.(Technical aspect). Applied and environmental microbiology, Vol. 51, No 3.
10. Jackson, M. L. 1999. Investigation of Adhesive and electrical performance of waterborne expoxies for interlayer dielectric material. PhD dissertation, Department of Material engineering science, Virginia Tech.
11. Khosrovi, B., R. Macpherson, and J.D.A. Miller. 1971. Some observation of growth and hydrogen uptake by Desulfovibrio vulgaris. Arch. Mikrobiol. 80: 324 – 337.
12. Kristjansson, J. K., P. Schonheit, Thauer R. K. 1982. Different Ks values for hydrogen of methanogenic bacteria and sulfate reducing bacteria: an explanation for the apparent inhibition of methanogenesis by sulfate. Arch. Microbiol. 131: 278 – 282.

13. Kroiss H. and Wabnegg F. P. 1983 Sulfide toxicity with anaerobic wastewater treatment. Proc. Eur. Symp. On Anaerobic Treatment, pp. 72 – 85. The Hague, the Netherlands.
14. Mara, D. D. and D. J. A. Williams .1971. The mechanism of sulfide corrosion by sulfate reducing becteria.
15. Edwards, M., B. Courtney, B., P. S. Heppler, M. Hernandez. 1997. Beneficial discharge of iron coagulation sludge to sewers., Journal of Environmental Engineering, pp 1027-1032.
16. Metcalf and Eddy, Inc. Wastewater Engineering, treatment, disposal, and reuse (3rd Ed.). McGraw-Hill, Inc., New York St. Louis San Francisco Auckland Bogata Tokyo Toronto.
17. Olver Inc., 1995. Sodium Sulfate Pilot Plant Treatability Study.
18. Olver Inc., (1998; 2000). Laboratory data on PFRWTA Anaerobic Digesters.
19. Padival, N. A., Kimbell, W. A., and Redner, J. A., 1995. " The use of iron salts to control dissolved sulfide in district" trunk sewers." Journal of Environmental Engineering, ASCE, 121, 824.
20. Pitt, P., Mamais, D., Cheng, Y. W., Loiacono, J., and Jenkins, D. 1992. " Struvite control using ferric chloride at the southeast water pollution control plant, San Francisco." Proc., Water Envir. Fed. 65th Annu. Conf. And Exposition, water Envir. Fed., Alexandria, VA.
21. Reiber, S., R. A. Ryder, and I. Wagner.1996. Corrosion assessment technoloies. Internal corrosion of water distribution systems, 2nd ed., 445-486. Denver, CO: AWWA Research foundation/TZW.
22. Robinson, J. A., and J. M. Tiedge. 1984. Competition between sulfate reducing and methanogenic bacteria for H_2 under resting and growing conditions. Arch. Microbiol. 137:26-32.
23. Singley J. E. et al., 1985. "Corrosion of iron and steel", internal corrosion of water distribution systems. Denver, CO: AWWA Research foundation/ DVGW Forschungsstelle.
24. Schonheit, P., J. K. Kristjansson, and R. K. Thauer. 1982. Kinetic mechanism for the ability of sulfate reducers to outcompete methanogens for acetate. Arch. Microbiol. 132: 285 – 288.

25. Speece, R. E., and G. F. Parkin. 1983. The response of methane bacteria to toxicity, p. 23 – 35. Proceedings of the 3rd International Symposium on Anaerobic Digestion. Evans and Faulkner, Inc., Watertown, Mass.
26. USEPA, 1992. Detection, Control and correction of hydrogen sulfide corrosion in existing wastewater systems., Washington D.C
27. Thauer R. K., Jungermann K. and Decker K. 1977. Energy conservation in chemotrophic anaerobic bacteria. Bacteria Rev. 41, 100-180.
28. Valcke, D., and W. Verstraete. 1983. A practical method to estimate the acetoclastic methanogenic biomass in anaerobic sludge. Journal of Water Pollution Control Fed. 55:1191-1195.
29. von Wolzogen Kuhr, C. A. H., and van der Vlught, L. S. 1934. Water, 18, 147.
30. Walker, S. J., and Morgan, W. J. 1995. " The addition of ferrous and ferric chloride for hydrogen sulfide and struvite control in anaerobic digesters." Metro Wastewater Reclamation District Rep., Denver, Colorado.
31. Widdel, F. 1988. Microbiologiy and ecology of sulfate and sulfur reducing bacteria. In Biology of anaerobic microorganisms (Edited by Zehnder a. J. B.), pp. 469 – 586. Wiley – Interscience, New York.