MEMBRANE BIOREACTORS FOR ANAEROBIC TREATMENT OF WASTEWATERS

by

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ABSTRACT

Anaerobic treatment of municipal wastewaters or other wastewaters of weaker strength has conventionally been difficult, especially in regions of cooler climate. Issues with solids retention and sufficient biomass concentration and activities required for treatment are usually of primary concern in these situations. However, the recent incorporation of membranes in the anaerobic treatment of municipal wastewater has made this technology feasible through the absolute retention of biomass within the reactor. Among the many advantages this technology has over its aerobic membrane bioreactor (MBR) counterpart, the most attractive may be the absence of an energy requirement for aeration. Although the anaerobic membrane bioreactor (AnMBR) is a promising technology, limited understanding of the interactions between the reactor biomass and the filtration membranes in the AnMBR has generally confined it to the lab scale level. To this end, the subject of examination for this study is the feasibility of AnMBR treatment of municipal wastewaters, with a strong focus towards membrane operation at moderate operating temperatures.

Two experiments were conducted during this study. The first was the long-term operation of two commercially available membranes (hollow fibre and flat sheet) operated in parallel within the reactor. The achievable operational OLR range for the AnMBR was about 1.25 kg COD/m^3·d. COD removal efficiency for the AnMBR system during biomass acclimatization was 84% with the addition of supplemental acetate in the feed. This decreased to 43% after the removal of supplemental acetate at a reactor temperature of 25 °C. Biogas production of the system decreased by 20% subsequent to a temperature decrease of 5 °C, due likely to decreased methanogenic activities.

The second experiment was to assess the mechanism(s) of membrane flux decline and to compare them with mechanisms identified for the same membranes applied in an aerobic MBR system. The major difference between filtration of the aerobic and anaerobic mixed liquor was the membrane surface cake

layer fouling. The observed anaerobic mixed liquor resistance was nearly 100 times greater than the aerobic contribution. It was observed that the mechanism of flux decline of the organic membrane in an anaerobic environment was due to surface cake layer fouling.

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$$Jv = \frac{TMP}{\mu(Rm + Rc + Rp + Ra)}$$ Equation 5.1 69

$$x_g = \frac{P_T}{H} P_g$$ Equation A.1 85

$$V = \frac{nRT}{P}$$ Equation A.2 86

NOMENCLATURE

AnMBR	Anaerobic membrane bioreactor
BAP	Biomass-associated products
BOD	Biochemical oxygen demand
CFV	Cross-flow velocity
COD	Chemical oxygen demand
CSTR	Continuously stirred tank reactors
GC	Gas chromatography
HRT	Hydraulic retention time
Jv	Membrane permeate flux
MBR	Membrane bioreactor
MLSS	Mixed liquor suspended solids
MLVSS	Mixed liquor volatile suspended solids
OLR	Organic loading rate
PAO	Phosphorus accumulating organisms
PHB	Polyhydroxybutyrate
PVDF	Polyvinylidene fluoride
SRT	Solids retention time
TMP	Trans-membrane pressure
TSS	Total suspended solids
UAP	Utilization-associated products
UASB	Upflow anaerobic sludge blanket reactor
R	Resistance
Ra	Resistance due to adsorption
Rc	Resistance due to concentration polarization
Rm	Intrinsic membrane resistance
Rp	Resistance due to pore plugging
Rs	Resistance invoked by sludge
SMP	Soluble microbial products
SS	Suspended solids
VFA	Volatile fatty acids
VSS	Volatile suspended solids
WERF	Water Environmental Research Foundation

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1.0 INTRODUCTION

The treatment efficiency of any conventional suspended growth biological wastewater treatment system depends heavily on the ability of the secondary clarifier to retain solids from the discharged effluent. A system with high biomass concentration is capable of generating higher quality discharge effluent while allowing the use of smaller reactor volumes. However, treatment systems capable of sustaining high concentrations of biomass usually require very efficient downstream gravity settling tanks, which return a majority of the biomass back into the bioreactor. Unfortunately, even the most efficient settling tanks may not be able to remove all suspended solids and biomass from the effluent stream. This is of primary concern because the discharged effluent may be a considerable source of oxygen demand (usually as biochemical oxygen demand (BOD), or chemical oxygen demand (COD)) on the receiving water. The design philosophy of the wastewater bioreactor treatment system coupled with solids-returning settling tanks has remained unchanged for more than one hundred years. Due to the extensive knowledge of current bioreactor treatment kinetics, the efficiency of the solids settling tanks can often be seen as the limiting factor in terms of significant treatment advancement in the present day.

1.1 Membrane Bioreactors

In order to satisfy the need for more stringent solids-liquid separation, progress in research which started in the 1960s, has revealed the possibility of a drastic departure from current conventional treatment designs. Studies have since publicized the use of microfiltration or ultrafiltration membranes as efficient physical barriers for solids retention in aerobic bioreactors (termed membrane bioreactors, or MBR) treating municipal and industrial wastewater (Stephenson *et al.*, 2000). The driving force behind the use of membranes is the elimination of critical factors governing secondary settling tanks within the treatment train. With the membranes installed, near complete retention of biomass can be achieved in the reactor and thus the treatment efficiency and size of the reactor is no longer limited by the settling characteristics of the suspended solids. It is now possible to truly separate and independently control the hydraulic retention time (HRT) and the solids retention time (SRT) as so required. Higher organic

loading rates (OLR) and lower HRTs are achievable in a bioreactor volume that is substantially smaller than those of conventional systems.

MBR technology has been gaining popularity in recent years, especially in areas such as industrial wastewater treatment, or in places where solids settling is an issue. The SRTs required for those treatment processes are considerably longer than typical municipal wastewater treatment plants. With the decreasing cost of MBR technology and increasing membrane efficiency, options such as the treatment of municipal, or low-to-medium strength industrial wastewaters, through the use of anaerobic systems, suddenly seemed more plausible.

1.2 Anaerobic Treatment

Traditional uses of anaerobic treatment systems have commonly been for the stabilization of waste solids and the treatment of industrial strength wastewaters. The first municipal sludge digestion system was installed in 1926 by a company that later became Dorr-Oliver Incorporated. Within a few following years, the anaerobic treatment of industrial wastewater became feasible and increased in popularity thereafter. However, the same problems persist with solids retention in the anaerobic system as with the activated sludge system previously mentioned. This is especially important, as hydrolysis of solids requires a greater amount of time in an anaerobic system when compared to a similar aerobic system operating under analogous conditions. This is primarily due to the slower kinetics of anaerobic biomass (Tchobanoglous *et al.*, 2003). With respect to dilute wastewater treatment with anaerobic technology such as from municipal sources, full-scale applications have really only existed in countries with warmer climates. This is because when operating at temperatures of about 10 to 20 °C in a suspended growth reactor, substantially slower reaction rates occur. As a result, the inevitably longer SRTs and lower OLRs translate into larger reactor volumes (Banik and Dague, 1997). Despite these drawbacks, research into anaerobic wastewater treatment systems to minimize these shortcomings has been strong in the past few decades, driven primarily by key advantages over its aerobic counterpart.

In a conventional aerobic activated sludge system, energy required for aeration can easily consume nearly half of the total electrical power required by a typical secondary wastewater treatment plant (EPRI, 1994). With this figure alone, it is immediately obvious that an effective municipal anaerobic wastewater treatment system can bring tremendous benefits. Not only do anaerobic systems consume less energy, they may also be net energy producers via methane production. Yields of 0.38 m^3 CH_4 per kg COD removed have been achieved through anaerobic treatment systems operating at 20°C (Tchobanoglous *et al.*, 2003). In areas where the economics permit, the methane produced in an anaerobic system can be used to generate electricity to subsidize the energy requirements of the entire treatment plant. Additionally, due to the considerably lower biomass yield coefficients as compared to conventional activated sludge systems, disposal of excess sludge becomes a less significant issue. This is an increasingly important concern as landfill sites are becoming more difficult to find and are usually associated with strict permits for disposal of biosolids due to the presence of specific compounds such as heavy metals. Furthermore, tighter regulations are in place for many regions regarding air emission discharges from incinerators, as well as sites where biosolids are used in land applications. Lastly, loading rates of 3.2 to 32 kg COD/m^3·d have been achieved with anaerobic systems instead of the 0.5 to 3.2 kg COD/m^3·d associated usually with the aerobic process (Speece, 1996). This ultimately translates into savings in capital and operational costs due to the allowance of smaller reactors, which is a driving factor for anaerobic treatment systems coupled with membrane technology. Recent and ongoing advancements in the field of anaerobic membrane bioreactors (AnMBRs) treating municipal and dilute stream wastewaters can potentially revolutionize the future of water pollution control.

1.3 Anaerobic Membrane Bioreactor

Up until the relatively recent incorporation of membranes in anaerobic bioreactors for the treatment of municipal wastewater, research in this field was not practical due to the low biomass production and retention in the system. Despite improvements in solids separation with conventional settling tanks, solids retention is still usually a limiting step in the treatment process due to the net loss of biomass from

the system. However, the integration of membranes has since made absolute biomass retention possible, decreasing the reliance on the rate-limiting methane fermentation step governed heavily by low biomass growth rates.

One of the key benefits from the use of membranes in anaerobic systems over conventional systems is the faster reactor start-up time as a result of the complete retention of microbial seed materials. This exclusive control of the SRT provides optimal treatment capability under different conditions due to the ability to operate at nearly any reactor mixed liquor suspended solids (MLSS) concentration. With longer SRTs, retained solids such as organic particulates and soluble high molecular weight compounds can now have prolonged residence time for the process of hydrolysis, resulting in a more comprehensive treatment (Tchobanoglous *et al.*, 2003). Non-biodegradable solids can be discarded with the wasted biomass instead of in the effluent, thus further improving the quality of discharge. Additionally, post treatment disinfection can be reduced, due to the decreased amount of pathogens released from the membrane treatment process (Baek and Pagilla, 2003). The AnMBR can also be an attractive technology for existing treatment plant upgrades or expansions if only a small footprint is provided.

1.4 External versus Submerged AnMBR and Reactor Configurations

The two current approaches to AnMBR design involve the different placement and operation of the membranes. The first and more common of the two is the side-stream filtration system, better known as the external membrane system. The membranes employed in this system are pressure-driven by a relatively powerful pump that can sustain a high trans-membrane pressure (TMP) for the process of effluent permeation, as well as a high rate of liquid cross flow near the membrane surface. This membrane setup usually operates in conjunction with suspended growth reactors in the form of continuously stirred tank reactors (CSTRs) due to the relative ease of construction and operation, but other configurations are possible. The use of CSTR suspended growth systems, however, has the disadvantage of exposing the membranes to the full concentration of reactor MLSS (Tchobanoglous *et*

al., 2003). This can have a detrimental impact on the permeate flux of the membranes over time due to the constant exposure to suspended, colloidal, and dissolved solids (further described in **Chapter 2: Background and Literature Review)**.

Other configurations which have been used in conjunction with the external membrane system are the upflow anaerobic sludge blanket reactor (UASB) and the anaerobic hybrid reactor (Wen *et al.*, 1999). These reactors offer some solids-liquid separation prior to final membrane filtration by promoting sludge settlement in the case of UASB, or by providing a fixed film medium in the hybrid reactor to which anaerobic biomass can attach without circulating through the external membranes. This is important because as Brockmann and Seyfried (1996) reported, for every 1 m^3 of permeate effluent in an external AnMBR, nearly 40 to 80 m^3 of liquid must be pumped to the membranes. Their study also found that increased frequency of sludge circulation through the membranes caused losses in biomass activity within the reactor; a 50% loss after 20 circulations. Choo and Lee (1998) reported that due to the shear associated with high cross flow velocities (CFV) required for membrane surface foulant removal, the resulting mean particle size in the reactor was decreased. Colloidal particles created through this shearing process can contribute to membrane surface fouling, as later described in the subsection entitled, **Colloidal Solids** (2.2.3.2) of Chapter 2.

The second approach to AnMBR design is the use of submerged membranes that operate under vacuum. This membrane setup requires much less pumping energy than the external system, and may even operate under gravity. As the name implies, the membranes used in this system are often immersed directly in the bioreactor. Due to the lack of controlled liquid CFV like the ones used in the external membrane system, foulant cake formation must be reduced by bubbling gas around the membranes to create hydrodynamic disturbances near the membrane surface. In the case of an anaerobic system, gas sparging with biogas produced in the reactor would be preferred. Since the reactor biomass does not need to be pumped through the membrane system to achieve flux, much of the biomass floc activity

should remain intact (Hernández *et al.*, 2002). As with the external AnMBRs, different reactor configurations can also be used to reduce the amount of biomass that comes into direct contact with the membranes.

A review of the literature indicates a lack of research on the submerged AnMBR system. However, clear evidence of the advantages is there to support its potential (Hernández *et al.*, 2002). With the lure of lower power consumption and higher organic loading rates (OLR), this research is focused on the feasibility of a submerged AnMBR for the treatment of municipal wastewaters. Determining the treatment efficiencies, especially under colder conditions, requires the careful examination of the interactions between the membrane and the bioreactor under continuous operation. Membrane issues such as fouling due to reactor mixed liquor characteristics, the operating parameters of the membranes, and the physical/chemical properties of the membrane itself need to be examined. Bioreactor issues, such as treatment efficiency under different operating conditions, also need to be investigated.

2.0 BACKGROUND AND LITERATURE REVIEW

The reviewed literature in this chapter pertains to the treatment of municipal or low strength wastewater with the use of AnMBR technology. Emphases are on the mechanisms governing the efficiency of submerged organic membranes, also known as polymeric membranes, and the critical parameters of the bioreactor.

2.1 Anaerobic Treatment of Municipal Wastewaters

Although the predominant membrane technology for municipal wastewater treatment has been the aerobic MBR, research has been active with respect to the application of AnMBR on the same waste stream. Traditional anaerobic treatment of municipal wastewater has been confined to warmer climates since the rate of particulates hydrolysis is heavily dependent on temperature. This is especially critical for anaerobic treatment in more temperate climates considering the relatively high levels of particulate matter present in municipal sewage (Lew *et al.*, 2003). Nevertheless, recent reviews into non-membrane anaerobic systems conducted by Kalogo and Verstraete (1999) and Seghezzo *et al.* (1998) indicate that anaerobic treatment technology is also feasible at lower temperatures, though the COD removal rates depend heavily on the long SRTs needed for the rate-limiting step of particulate hydrolysis. These reviews report that UASBs operating at temperatures below 20 °C achieved COD removal efficiencies of no more than about 70% despite the long SRTs. Findings from the study of Singh and Viraraghavan (2003), and Lew *et al.* (2003) showed similar results. These results imply a distinct advantage for the use of membranes in the anaerobic treatment of wastewater, even under more temperate conditions.

Wen and others at the University of Tsinghua in Beijing used a hybrid UASB coupled with submerged hollow fibre membranes located at the top of the reactor to achieve treatment of domestic wastewaters from the university at ambient temperatures (Wen *et al.*, 1999). The system operated with sludge concentrations between 16 and 21.5 g/L, and HRTs as low as 4 hours. It was capable of achieving over

97% COD removal with a resulting effluent COD of less than 20 mg/L. They achieved this while operating at average OLRs of 0.5 to 12.5 kg COD/m^3·d with temperatures varying between 12 to 27 °C.

Stuckey and Hu (2003), and Baek and Pagilla (2003) conducted lab scale AnMBR treatment of municipal or municipal-like wastewater with some degree of success. Stuckey and Hu (2003) observed COD removal efficiencies of up to 93% through the addition of powdered activated carbon while operating with reactor mixed liquor volatile suspended solids (MLVSS) of about 3000 mg/L and HRTs that ranged between 3 and 32 hours. They operated two different microfiltration submerged membranes and achieved an effluent COD of 30 mg/L. Baek and Pagilla (2003) observed COD removals of only 68% but they attributed the lower treatment efficiency to biomass decay given the small reactor volume and external tubular membrane system used. Despite the fact that both Stuckey and Hu (2003) and Baek and Pagilla (2003) operated their systems at temperatures between 32 and 35 °C, their findings are still valuable for the purpose of the present study.

Although Manariotis and Grigoropoulos (2002) did not utilize membranes in their three-chamber anaerobic baffled reactor (ABR), their evaluation of low-strength synthetic wastewater (COD 300 to 400 mg/L) treatment at temperatures as low as 16 °C yielded promising results. They reported a COD removal of up to 92% which was achieved at 16 °C with an OLR of up to 0.7 kg COD/m^3·d. Despite a significant decline in operating temperature from 26 °C, the researchers observed little difference in reactor performance with the exception of a 30% decrease in biogas production. This may have been due to both the reduced methanogenic activities and the temperature effects on biogas solubility.

2.2 Mechanisms Governing Efficiency of the Membranes

Membrane flux is the main parameter by which different types of membranes are compared. However, if flux is sustained only through the use of high CFV or TMP, the system may not be economically feasible. Obtaining a thorough understanding of the mechanisms behind a sustainable operating flux is therefore

fundamental. Fortunately, extensive work has been done on the investigation of mechanisms impacting the permeate flux in aerobic MBRs. Many of the lessons learned from these investigations may be applicable to anaerobic membrane operation as a starting guide. However, just as the physical, chemical and biological characteristics of the mixed liquor in the aerobic system differ from those under anaerobic conditions, it can be expected that the mechanisms impacting the membrane flux will differ as well (Van Houten *et al.*, 2001). Additionally, because the type of wastewater and the operating conditions of bioreactors significantly influence the MLSS characteristics of the biomass (Kataoka *et al.*, 1992), the focus of this review will be research conducted on the treatment of municipal wastewater or others of similar nature.

There are three identified factors that can significantly impact the flux of an AnMBR. The first is the physical characteristics of the membrane itself. This includes the composition of the membrane material (polymeric vs. inorganic), the surface charge, the pore size, the membrane area per unit volume (packing density), and the configuration. The second is the operational parameters of the membrane. The important operating conditions include surface shear, operating TMP, operating temperature, and flux recovery. Lastly, the characteristics of the mixed liquor being filtered play a significant role in the sustainability of membrane flux. For this, parameters such as the OLR, the SRT, the HRT, and the operating temperature are important.

2.2.1 Membrane System

2.2.1.1 Membrane Material

There are several different types of organic membranes available, each with their own strengths and limitations. An intrinsic membrane property that will be discussed in the following section is the polarity of the membrane surface, which can be classified as either hydrophilic or hydrophobic in nature. When a membrane is described as hydrophilic, it can be wetted by water and has an angle of contact of less than 90°. A hydrophobic membrane is one that cannot be wetted by water and has a contact angle of more

than 90°. Hydrophobic membranes are inherently more chemically stable than hydrophilic membranes and can withstand strong cleansing agents (Van Houten, 2003). However, studies have shown that hydrophilic membranes produce less fouling than hydrophobic membranes. Many commercially available membranes are now made of hydrophobic polyvinyldifluoride (PVDF), but are also provided with a hydrophilic coating to incorporate foulant-reducing ability with good chemical resistance.

There is a general consensus on the cause of fouling of organic membranes as described by several researchers. Kang *et al.* (2002) and Choo and Lee (1996a) reported that the principal cause for membrane hydraulic resistance is the cake layer that forms on the membrane surface. Both groups indicated that the cake layer was composed of a mixture of biomass and inorganic precipitant in the form of struvite ($MgNH_4PO_4 \cdot 6H_2O$). Furthermore, Lee *et al.* (2001) reported that the resistance due to the cake layer was significantly greater than the resistance imposed by internal fouling for the organic membrane. Internal fouling is generally characterized by the adsorption of soluble and/or particulate material within the pore structure.

2.2.1.2 Hydrophobic Nature and Charge of the Membrane

Choo *et al.* (2000) and Sainbayar *et al.* (2001) found correlation between the hydrophobic nature of the membrane material and the sustainable membrane flux in an anaerobic system. They reported that membranes that are hydrophilic in nature tend to foul less and can sustain higher flux for longer periods of time. Sainbayar *et al.* reported that membranes that were hydrophilic decreased the hydrophobic interaction between microbial foulant and the membrane surface. Additionally, they concluded that the permeate flux of the hydrophobic membrane could be improved through graft polymerization with 2-hydroxyethyl methacrylate (HEMA) onto the membrane surface. This imparts an increased hydrophilic nature to the membrane by changing its surface properties. Choo *et al.* determined that a 70% degree of grafting was optimal because the pore size of the membrane decreased with extent of grafting. Continuous increase in grafting has the potential for increasing the intrinsic membrane resistance.

Membrane fouling is also affected by the membrane surface charge. Shimizu *et al.* (1989) reported that negatively charged inorganic membranes fouled less than neutral or positively charged membranes during anaerobic filtration. It was determined that the negatively charged colloids in the mixed liquor experienced strong electrical repulsion and thus had more difficulty attaching to the membrane surface. However, the membrane surface charge is strongly dependent on the pH and the ionic strength of the mixed liquor. Fane *et al.* (1989) reported that a high ionic concentration in the mixed liquor reduces the impact of the membrane surface charge.

2.2.1.3 Nominal Pore Size

There are supporting literatures which state that the nominal pore size of the membrane contributes substantially to the rate at which it fouls. Choo and Lee (1996b) reported that the membrane pore size that had the least tendency to foul was 0.1 μm while filtering anaerobic digestion broth. Elmaleh and Abdelmoumni (1997) determined that the optimum flux achieved when filtering a methanogenic broth was obtained through the use of a 0.14 μm membrane. They found this contradictory to their previous result whereby a 0.45 μm membrane achieved the highest flux when filtering a broth mixed with both acidogenic and methanogenic cells. This result implies the possible need for different pore size when filtering different types of mixed liquor. Chung *et al.* (1998) reported that the permeate flux achieved with a 0.22 μm membrane was three times greater than that of either a 0.45 or 0.6 μm membrane.

While filtering high strength wastewater using membranes with a larger pore size, He *et al.* (1999) determined that the membranes fouled more rapidly due to blockage by macro-colloids. This was consistent with the results from Elmaleh and Abdelmoumni (1997). They reported that membranes with larger pore size tend to foul quickly due to internal pore clogging. Saw *et al.* (1986) reported that although the initial flux from a membrane with larger pore size was greater than a membrane with smaller pore size, it fouled more rapidly with time. The results were consistent with work done by

Imasaka *et al.* (1989) from which they reported that the rate of fouling increases with an increase in the membrane pore size due mainly to internal fouling.

2.2.1.4 Membrane Configuration

The configuration of submerged membranes produced a difference in performance as found by Stuckey and Hu (2003). They observed that the hollow-fibre internal membrane achieved a slightly higher flux than a comparable flat-sheet membrane operating under the same conditions. Lei and Bérubé (2004) found significant flux improvements through physical contact between submerged membrane fibres. This may have been the same mechanism which could explain the outcome from the research conducted by Stuckey and Hu (2003).

2.2.2 Operational Parameters

2.2.2.1 Gas Sparging

The use of gas sparging in submerged AnMBRs is effective at reducing membrane fouling by creating high shear conditions at the membrane surface. This shear, as reported by Choo and Lee (1998), significantly reduces the resistance due to concentration polarization and cake layer formation. Concentration polarization is the reversible accumulation of solids on the retention side of the membrane, which can form a permanent cake layer over time.

Through their study with submerged ceramic membranes, Kayawake *et al.* (1991) found that a two-fold increase in permeate flux could be achieved as a direct result of head gas sparging of the membranes. As Stucky and Hu (2003) found through the sparging of head space gas, the TMP required to sustain a constant permeate flux decreased with an increase in gas sparging rate. This effect however, reached a plateau at which further TMP decrease was not possible with a continuous increase in gas sparge rate. When Imasaka *et al.* (1989) supplied nitrogen gas for sparging as well as mixed liquor circulation through tubular ceramic membranes due to gas-lift pump effect, the highest initial permeate flux was

achieved with the highest tested gas sparging rate. However, they later discovered that at higher gas sparging rates, the flux had a greater tendency to decline over time, when compared with membranes of lower sparging rates. They described this phenomenon as thinning of the "self-rejective dynamic membrane"; a situation in which when the rate of gas sparge increases, the higher shear forces thins out the cake layer. In this case, thinning of the cake layer decreases the protective layer against the passage of foulants towards the membrane pores, which eventually become plugged. It has then been found that pore plugging is the main contributor to flux decline.

2.2.2.2 Trans-Membrane Pressure

The operating TMP for the submerged membrane is generally below 80-100 kPa. Beaubien *et al.* (1996) reported that when operating below these pressures, the permeate flux is dependent on the TMP applied. It was also determined that at low TMPs and MLSS concentrations below 2.5 g/L, the permeate flux was adversely affected by any MLSS increase. Factors contributing to low membrane permeability include pore plugging, adsorption, and concentration polarization. However, MLSS increase beyond 2.5 g/L did not yield any additional negative impact on the flux, up to 25 g/L. Finally, Beaubien *et al.* (1996) did not observe any significant impact on flux with an increase in CFV at low TMPs.

2.2.2.3 Operating Flux

In general, it has been indicated that operating a membrane at maximum flux tends to increase the rate at which membrane fouling occurs, due to the greater amount of foulant material moving towards the membrane surface (Wen *et al.*, 1999). Attainable flux is not only dependent on membrane properties but also the anaerobic mixed liquor properties and other environmental conditions such as temperature.

2.2.2.4 Permeate Flux Recovery

Due to irreversible fouling over the operating life of the membrane, it is necessary to periodically stop membrane operation for cleaning to recover flux. The common method of cleaning requires the alternating use of chemical agents such as caustic and acidic solutions for foulant cake layer removal. It

is accepted that acidic solutions are effective in removal of inorganic foulant while caustic solutions are effective at removal of organic/biological foulant. While treating piggery wastewater, Lee *et al.* (2000) did not observe considerable flux recovery by using caustic solution alone. This was despite the fact that a considerable amount of the cake layer was composed of biological material. However, when acidic and caustic solutions were used in series, a significant increase was observed. This result further reinforces the fact that organic membrane foulant cake layers are often comprised of organic and inorganic constituents. Kang *et al.* (2002) and Choo *et al.* (2000) reported that it was possible to recover significant membrane flux through the exclusive use of acidic solutions. This suggests that the removal of inorganic precipitants is a key aspect of foulant cake removal on organic membranes.

Membrane relaxation has been reported to decrease the rate of fouling through the reduction of TMP to zero for a small increment of time after each permeation cycle. This would allow for more efficient foulant removal from the surface of the membrane via liquid cross-flow; a feat usually attained by gas sparging in the case of submerged membranes. Wen *et al.* (1999) identified an ideal relaxation cycle of 4 minutes permeation pumping on, and 1 minute off, to sustain membrane operation for a reasonable amount of time before cleaning.

2.2.2.5 Operating Temperature

Not only does the operating temperature have a profound impact on the microbial activities in the system and the rate of particulate hydrolysis, it also influences the permeation flux of membranes through changes in liquid viscosity. When operating at higher temperatures, an increase in permeate flux can be expected due to the lower liquid viscosity. Hogetsu *et al.* (1992) observed a 33% increase in flux of the membrane by increasing the temperature of the reactor from 40 °C to 47 °C. Zoh and Stenstrom (2002) also reported improved flux when the reactor temperature was increased from 15 °C to 40 °C.

Another important factor governing the permeate flux of membranes is the production of soluble microbial products, or SMP. As discussed in a following subsection entitled, **Soluble Products** (2.2.3.3), SMP production in the reactor has a detrimental impact on the membrane permeate flux. SMP concentration in the reactor has been linked to the operating temperature of the system. Schiener *et al.* (1998) reported a decrease in SMP concentration as the reactor temperature increased. This observation was in agreement with Barker *et al.* (2000) who observed an increase in SMP production when the reactor temperature was decreased.

2.2.3 Characteristics of the Mixed Liquor

2.2.3.1 Suspended Solids

The level of suspended solids in the mixed liquor of the reactor has been reported to impact the permeate flux of the reactor. Beaubien et al. (1996) observed a near linear decrease in the stabilized permeate flux as the concentration of the mixed liquor increased from 2.5 g/L to 22 g/L. While filtering a digested sludge, Saw *et al.* (1986) reported a log-linear decrease in steady state flux with an increase in mixed liquor suspended solids concentration. Both Harada *et al.* (1994) and Pillay *et al.* (1994) observed similar results when they increased the mixed liquor concentration during filtration. These observations can be explained by the higher rate of mass transfer towards the membrane surface as the MLSS concentration increases, while the rate of mass transfer away from the membrane decreases. Additionally, Lübbecke *et al.* (1995) reported an exponential increase in the measured viscosity of the MLSS as the biomass concentration increased in an aerobic MBR. The lower turbulence at the membrane surface caused by higher liquid viscosity is generally characterized by lower back-transport velocities.

2.2.3.2 Colloidal Solids

Choo and Lee (1996b, 1998) concluded that the constituent in the anaerobic broth most responsible for the cake layer resistance was fine colloidal solids. They indicated that this was due mainly to the fact

that smaller particle sizes tend to have greater net particle velocity toward the surface of the membrane. Under these conditions, more compact cake layers are created by smaller particles, which lead to an increase in cake resistance as a result of porosity decrease. Choo and Lee (1996b) suggested that because the polarization index of the fine colloids was substantially higher than the cells and soluble components of the anaerobic broth, flux improvement could be achieved by either degrading the colloids into soluble components or by flocculating them into bigger particles. Furthermore, Langenhoff *et al.* (2000) observed an increase in the production of SMP from reactors fed with colloidal material. As will be discussed in further detail in the next section, SMP has been well documented for its contribution to membrane fouling.

2.2.3.3 Soluble Microbial Products

There is a distinct correlation between the production of SMP in the reactor and the level of membrane fouling. Through their aerobic MBR research, Lee *et al.* (2001a) discovered unexpectedly that the rate of membrane fouling was much higher in an attached growth system than that of a suspended growth system, which leads to the conclusion that the soluble fraction of the mixed liquor contributes appreciably to membrane fouling. They determined that the formation of a dynamic membrane (deposition of surface cake layer) can adsorb or reject soluble organics which would have otherwise been adsorbed directly to the membrane surface and cause irreversible fouling. Shin and Kang (2002) determined that dissolved solids induced the majority of short term membrane resistance with their research into aerobic MBR systems. Although these results were obtained from aerobic systems, the effects of SMPs in AnMBRs may be similar. Harada *et al.* (1994) concluded that it was the formation of a gel layer through the accumulation of soluble substances in their AnMBR which played the most significant role in fouling, rather than the deposition of suspended solids.

Through a review of literature, Barker *et al.* (2000) defined SMP as compounds of both high and low molecular weight, or more simply, "the pool of organic compounds that results from substrate

metabolism (usually with biomass growth) and biomass decay". The results of their study indicate that the production of SMP increased with increased HRT, due perhaps to the elevated biomass decay from a lack of substrate. In addition, Barker *et al.* (2000) and Schiener *et al.* (1998) both observed an increase in SMP production as the operating temperature decreased. This they explained could have been due to the increased stress on the biomass and the lower rate of biodegradation of SMPs.

The production of SMPs is affected by the OLR, the SRT, the HRT, the type and strength of feed, and the concentration of biomass as reported by Barker *et al.* 2000. Furthermore, SMPs can be classified into utilization-associated products (UAPs) through the process of biomass substrate utilization, and biomass-associated products (BAPs) which occur from biomass decay. These may resolve the inconsistency in results of optimal AnMBR loading in terms of SMP production and membrane operation. In this case, Barker and Stuckey (2001) reported an increase in SMP formation with higher feed COD concentrations, while Shin and Kang (2002) observed that for an aerobic-anoxic MBR, filtration resistance due to the soluble fraction was more severe at longer SRTs.

2.2.3.4 Inorganic Precipitates/Struvite

The formation of struvite ($MgNH_4PO_4 \cdot 6H_2O$) is favoured by the higher pH conditions created by the degradation of biological organic acids in an anaerobic reactor (Yoon *et al.* 1999). Choo and Lee (1996a) reported a typical struvite formation pH range of 7.5 to 8.5. Despite the self-sustaining concentrations of soluble ammonium and phosphate due to biodegradation, the concentration of magnesium is usually only sustained through the incoming feed. Yoon *et al.* (1999) indicated that the amount of struvite precipitated on the membranes can be calculated through the difference between the influent and effluent concentrations of magnesium. Formation of the cake layer on organic membranes contributes to the majority of fouling and filtration resistance (Choo and Lee, 1999), and is comprised of microbial cells layered with struvite.

3.0 OBJECTIVES OF THE RESEARCH

Several decades of research have demonstrated the potential of AnMBR treating municipal or low strength industrial wastewaters as a potential alternative to traditional approaches. However, a significant amount of research is still required to assess the feasibility of this technology prior to full-scale application. Several key issues have been identified which require careful attention in the topics of system efficiency and economic feasibility. However, these were not the subjects of focus in this research due to the considerable amount of literature which indicates the treatability of municipal wastewater through conventional anaerobic processes. Rather, the primary focus of this investigation was the membrane component of the AnMBR. Issues with regards to physical and chemical properties of the membrane and mechanisms of membrane fouling were closely examined.

The research reported here was completed in conjunction with a parallel AnMBR study utilizing the same wastewater feed with external tubular membranes instead of the submerged type. This research was part of Water Environmental Research Foundation's (WERF) investigation (WERF project 02CTS4) into the feasibility of AnMBR treatment of municipal wastewater. The parallel setup was for the purpose of direct comparison between the two systems using different membranes, which was detailed in the report "Membrane Bioreactors for Anaerobic Treatment of Wastewaters" (WERF 02CTS4). The insights assembled in the present study were intended to generate guidelines for an AnMBR pilot study for the treatment of municipal wastewater. Furthermore, it was anticipated that membrane fouling through the treatment of municipal wastewater would be more complex than those observed from the treatment of industrial wastewaters. Several specific objectives were set to address the operational concerns regarding the membrane component of the AnMBR. They were as follows.

1. To assess the impact of selected AnMBR operating and design parameters on submerged membrane flux and MBR treatment system performance.

2. To assess the efficacy of different, commercially available submerged membrane modules when operated in an AnMBR treating municipal wastewater.
3. To assess the mechanism(s) of membrane flux decline in a submerged membrane AnMBR treating municipal wastewater and to compare these to mechanisms identified for the same membranes applied in an aerobic MBR system.
4. To identify options for optimizing long-term average membrane flux in a submerged membrane AnMBR treating municipal wastewater.

4.0 EXPERIMENTAL SETUP AND OPERATION

4.1 Experimental Setup

4.1.1 Bioreactor Setup

The schematic layout of the submerged membrane system is presented in Figure 4.1. This reactor system was made of Plexiglas and had a volume of 40 L. It was designed to incorporate two different submerged filtration membranes for simultaneous performance comparison. Both membranes were located inside the reactor and were separated by baffles to promote independent hydraulic conditions. Each membrane was operated with independent peristaltic pumps (Masterflex L/S, Cole-Parmer, Quebec, Canada) for effluent permeation. The Masterflex pumps were controlled by timing switches to allow for membrane relaxation cycles during operation. With the baffles in, the submerged membrane bioreactor was designed to accommodate different OLRs through a simple change in reactor liquid level or influent/effluent flow rate.

An external gas pump (EW 79200, Cole-Parmer, Quebec, Canada) was used for membrane gas sparging to reduce membrane fouling through surface scouring. Gas used for scouring was recycled from the reactor head gas zone through the top of the system. Gas flow meters were used to determine the amount of gas used for each membrane.

Ultrasonic level sensors (Ultrasonic Level Switch, Cole-Parmer, Quebec, Canada) controlled the depth of the mixed liquor within the reactor. Once the high level sensor was triggered, the solenoid valves controlling the discharge of permeate opened. When the liquid level decreased to below the set point, the solenoid valve closed and the membrane permeate was recycled back into the system. A constant liquid level could be maintained depending on the desired OLR.

Gas production of the reactor was recorded through the use of a submerged gas tip meter. Gas count was recorded daily, as was the volume derived from the number of tips. Gas collection for analysis utilized glass in-line collection vessels which were removable for transfer to the lab.

A pH control system was installed, but not used during the experiment due to the adequate buffering capacity of the influent feed. Unlike the pH probe setup shown in the schematic, actual reading of the reactor pH was done on mixed liquor samples and membrane effluent. This was due to the difficulty encountered in acquiring an accurate reading while the probe was in-line.

The operating temperature of the reactor was maintained through the use of a submersible heater, activated by a thermocouple and thermostat.

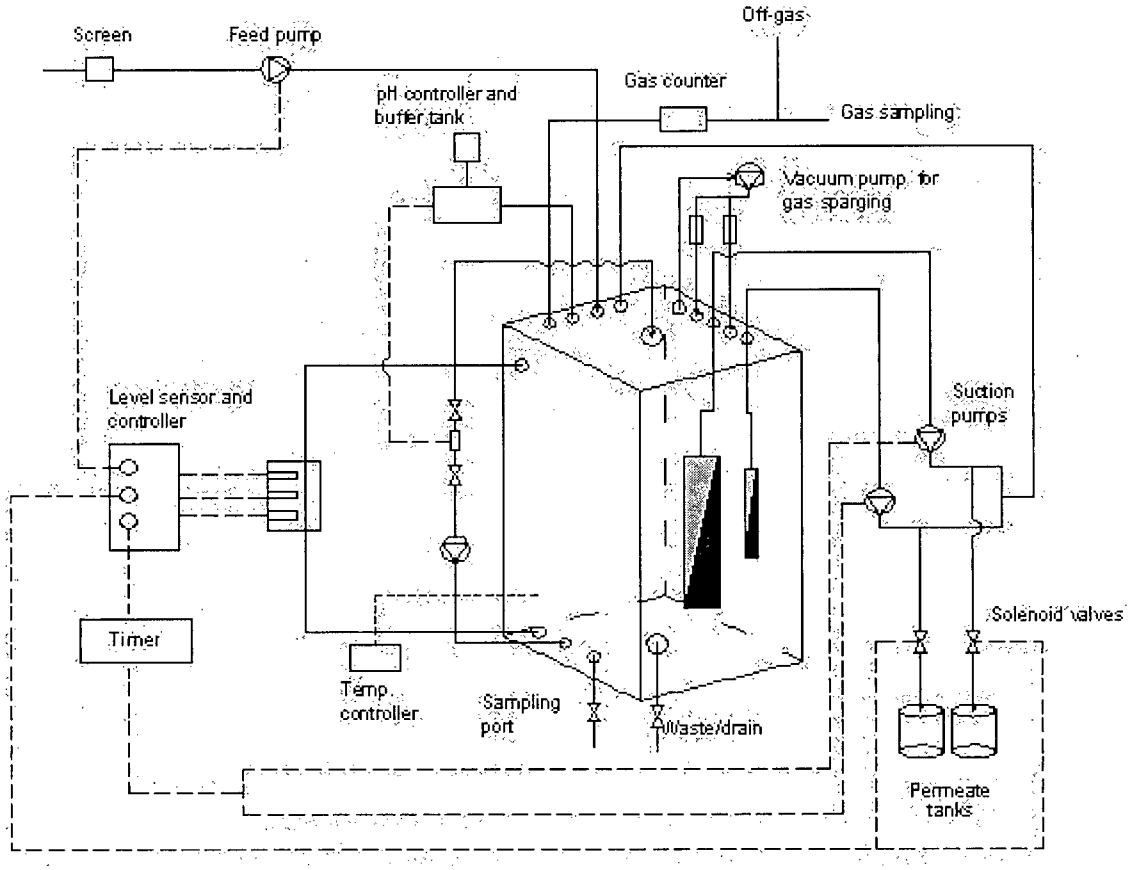

Figure 4.1 Schematic of Submerged Anaerobic Membrane Bioreactor.

4.1.2 Membrane Setup

Two different membranes from different manufacturers were used in this study. The membrane properties presented in Table 4.1 are specifications provided by the manufacturers. The hollow fibre membrane used was provided by US Filter while the flat sheet membrane was from Toray. The original membrane proposed for this study instead of the Toray flat sheet membrane was the Ionic/Mitsubishi hollow fibre membrane. However, it was later concluded that the membrane module from Ionic/Mitsubishi was too fragile for the nature of this study and was thus replaced by the flat sheet membrane. The membrane materials used by both manufacturers were polyvinylidene fluoride (PVDF), which provided relatively high stability against chemicals and good physical strength. Membrane surface properties for both systems were hydrophilic and negatively charged in nature.

Table 4.1 Submerged membrane manufacturers' specifications

Name	US Filter	Toray
Supplier Contact	Marie Rodde US Filter/JetTech 1051 Blake Edwardsville, KS 66111 913-422-7600	Dave Botwright Sanitherm Engineering Limited 431 Mountain Highway, Suite 4 North Vancouver, B.C. Canada V7J 2L1 604-986-9168
Material	PVDF	PVDF
Description	Hollow fibre, outside in flow	Sheet membrane, outside in flow
Surface Characteristic	All membranes hydrophilic/negatively charged	
Nominal Pore Size (μm)	0.08	0.08
Surface Area (m^2)	0.5	0.18
Pressure (kPa)	Max. -69	Max. -21
Water Flux (Lmh)	25	16-29
Fouling Removal	Head gas sparging 3-10 LPM	
Operating Condition	Relaxation cycles (Manufacturer Recommended)	
	12 mins on, 1 min off	8 mins on, 2 mins off

The hollow fibre membrane provided by US Filter was a pilot-scale module while the flat sheet membrane consisted of a custom module made at UBC from sections of full-scale Toray flat sheet membranes.

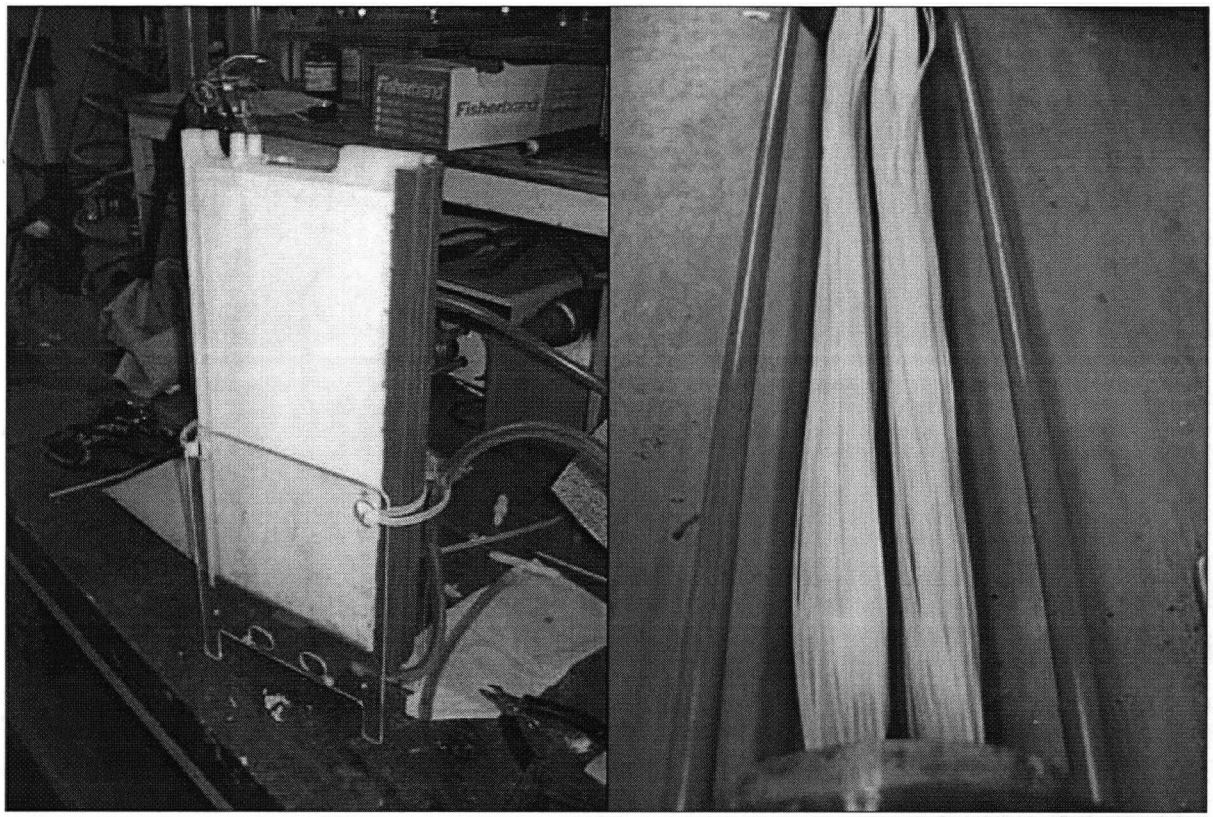

Figure 4.2 Submerged membranes used. Toray flat sheet membrane on the left and US Filter hollow fibre membrane on the right.

4.2 Experimental Description

This study investigated the feasibility of treating low strength municipal wastewater with AnMBRs at relatively low temperature conditions. The subject of investigation for this study could be broken into two sub-categories. The first issue examined was the viability of municipal wastewater treatment using acclimatized anaerobic biomass. The second issue addressed the feasibility of treatment through the use of different membranes for solids retention. The success in the overall feasibility of the AnMBR is heavily dependent on the synergy of these two key processes.

4.2.1 Biological Treatment

Addressing the need for a better understanding of anaerobic municipal wastewater treatment, a few critical bioreactor parameters were controlled. Through the review of literature, the relevant factors

selected affecting the success of an AnMBR were the bioreactor operating temperature, the OLR, and the biomass characteristics (Sutton *et al.* 2003).

The operating temperature of the system has a clear impact on the reactor treatment efficiency. Treatment at higher temperatures (high mesophilic to thermophilic range) is relatively well understood and is often associated with very high removal efficiencies. However, the focus of this study required decreased target operating temperatures that are comparable to treatment conditions found in colder climates, about 20 to 25 °C.

Bioreactor OLR (which was controlled by varying the HRT) dictates the treatment capabilities. Results from the literature review Sutton *et al.*, 2003 indicated that AnMBR systems should be capable of OLRs of 1 to 8 kg COD/m^3·day, which is made possible by the excellent solids-retaining capability of the membranes. However, the actual OLRs that could be sustained in the present study depended heavily on the performance of the selected membranes. It was thought that the likely range would be closer to 2 kg COD/m^3·day due to the low membrane flux expected. With total solids retention, a substantial biomass population can be maintained for effective treatment, especially under colder operating conditions. The target MLSS concentration for the AnMBR in the present study was about 8,000 mg/L, as described later in the subsection entitled, **AnMBR System Start-Up** (5.1).

Seed biomass was taken from the anaerobic sludge digester located at the Lulu Island Secondary Wastewater Treatment Plant located in Richmond, British Columbia. The operating temperature of that system was 37.5 °C with a pH of 7.2.

Acclimatization of the biomass required the initial dilution of the anaerobic digester sludge with raw municipal wastewater to an appropriate solids concentration suitable for use in the AnMBR. Start up of the reactor utilized supplemental acetate (in the form of sodium acetate, CH_3COONa) in the raw

wastewater feed to insure a viable methanogen population. This acclimatization process was considered to be complete after the termination of acetate supplementation at the point when the system finally reached steady state in terms of the levels of the measured parameters. More specifically, the determination of efficiency and reactor health for the AnMBR in terms of biological operation was assessed with the parameters listed in Table 4.2. Other parameters such as **OLR** (5.2.2.1) and **HRT** (5.2.2.2) were calculated based on the influent feed flow and values (i.e. COD) determined from the parameters measured in Table 4.2.

Table 4.2 Reactor parameters measured

Parameter Measured	**Location Measured**
Temperature/pH	Reactor sampling port
Chemical Oxygen Demand (COD)	Influent and effluent
Volatile Fatty Acids (VFA)	Influent and effluent
Ammonia and Nitrates	Influent and effluent
Phosphates	Influent and effluent
Reactor MLSS/MLVSS	Reactor sampling port
Biogas Production	Head gas collection
Biogas Composition	Head gas collection
Sludge Particle Size Distribution	Reactor sampling port

4.2.2 Membrane Performance

Despite improvements in conventional solids-liquid separation technologies, anaerobic biomass retention without the use of membranes is likely insufficient for efficient treatment of municipal wastewaters. Due to the membrane-based separation of SRT from HRT, absolute biomass retention is possible in the treatment of low strength wastewater during which the methane fermentation step is characterized by low biomass growth yields. However, because of the high mixed liquor suspended solids concentration and the possible release of membrane-fouling SMPs at lower operating temperatures, an adequate understanding of membrane performance and operation under these conditions is critical for success. As previously discussed, the characteristics of the mixed liquor being filtered were expected to significantly

affect the permeate flux in an AnMBR. These characteristics in turn are largely affected by the operating parameters that are specific to the biological component of an AnMBR. The impact of operating parameters such as the OLR, the HRT, and the operating temperature, on the permeate flux in an AnMBR were investigated.

Membrane performance in the reactor was evaluated through the measurement of permeate flux and TMP as functions of time. The submerged membranes were operated with gas sparging comparable to the manufacturers' suggested values for the purpose of foulant reduction at the membrane surface. Membrane relaxation cycles were based on conversations with the membrane manufacturers in conjunction with reviewed literature, which recommended cycles consisting of permeation steps ranging from 6 to 10 minutes and relaxation steps ranging from 1 to 3 minutes. Additionally, different membrane flux recovery techniques were explored to determine the optimal cleaning methods in order to discern the feasibility of the submerged membrane operation in the anaerobic environment. Membrane cleanings are required when the permeate flux can no long achieve the target HRT, or when the TMP is close to, or exceeds the manufacturers' recommended values. In the present study, cleaning methods generally followed manufacturers' guidelines with only slight variations. Membrane cleaning required the complete shutdown of the reactor, often for a few hours, depending on the type of cleaning undertaken. Table 4.3 provides an overview of the different membrane cleaning procedures used during this study along with their type designation. Cleanings were usually performed at a liquid temperature of 40 °C.

Off-line filtration tests were conducted on both anaerobic and aerobic mixed liquors at similar MLSS concentrations, using the hollow fibre membranes, to obtain a better understanding of the differences in the filtration resistance of each. Membranes used for this part of the research were PVDF Zenon (Zenon Environmental, Oakville, Ontario) hollow fibre membranes with a nominal 0.04 μm pore size. These membranes were similar to the US Filter membranes used during this research, possessing surface

characteristics that were both hydrophilic and negatively charged in nature. The filtration setup in Figure 4.2 was adopted from Geng's (Z. Geng, Dept. of Civil Engineering, UBC, Vancouver, B.C., pers. comm., 2005) work at UBC involving membrane fouling characteristics. The equipment included a 2 L reactor that supported five strands of hollow fibre membranes to be operated with nitrogen gas sparging during filtration. A digital pressure gauge was installed on the permeate line to measure pressure over time. The system was driven by a Masterflex positive displacement pump equipped with a variable speed controller. Additional details on the testing procedure and results are presented in Section 5, **Results and Discussion**.

Table 4.3 Membrane cleaning procedure

Cleaning Type	Step Details
I	1. Water rinse. 2. 300 mg/L sodium hypochlorite soak for 30 min. 3. 1000 mg/L citric acid soak for 30 min.
II	1. Water rinse. 2. 1000 mg/L sodium hypochlorite soak for 1hr. 3. 2000 mg/L citric acid for soak for 1hr. 4. 1000 mg/L sodium hypochlorite soak for 1hr.
III	1. Water rinse. 2. Sodium hydroxide at pH 10.5 with 10 min permeation and 20 min. relaxation, all with air sparge. 3. 1000 mg/L sodium hypochlorite at pH 10.5 with 30 min permeation and 30 min relaxation, all with air sparge.
IV	1. Water Rinse. 2. 1000 mg/L sodium hypochlorite at pH 10.5 with 10 min permeation and 10 min relaxation, all with air sparge. 3. Sodium hydroxide at pH 10.5 with 10 min permeation and 10 min relaxation, all with air sparge. 4. 2000 mg/L citric acid at pH 2 with 10 min permeation and 10 min relaxation, all with air sparge. 5. Repeat steps 3 and 4 twice.
V	1. Water Rinse. 2. 1000 mg/L sodium hypochlorite at pH 10.5 with 10 min permeation and 10 min relaxation, all with air sparge. 3. 2000 mg/L citric acid at pH 2 with 10 min permeation and 10 min relaxation, all with air sparge. 4. Repeat steps 2 and 3 twice.

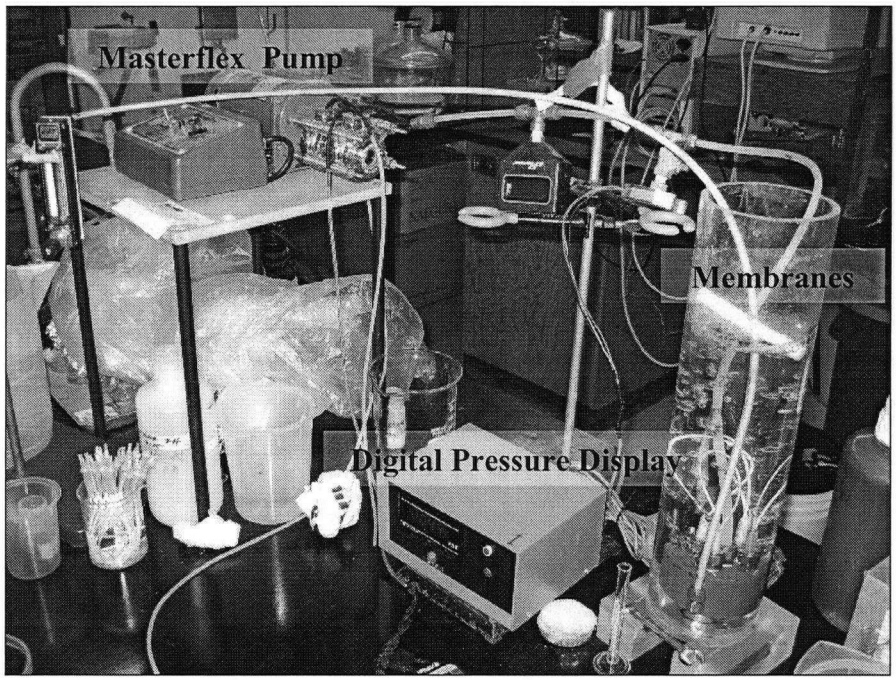

Figure 4.3 Setup for off-line membrane filtration tests.

4.2.3 Monitoring Program

The monitoring program consisted of onsite data collection as well as off-line laboratory measurements. Onsite data collection involved measurement of parameters such as reactor temperature, pH, biogas production, membrane permeate flux, and the TMP. These values were measured daily following the initiation of AnMBR operation.

The sampling port for the influent feed was located on the feed tube, just before the reactor entrance, while the effluent sampling ports were the individual membrane effluent discharge ports. The samples taken were analyzed for the parameters listed in Table 4.2. Samples taken for lab analysis were transported within 30 minutes of sample collection and preserved in accordance with the specifications of Standard Methods (APHA *et al.* 1998). Lab sampling occurred three times per week, unless otherwise specified.

4.3 Analytical Methods

4.3.1 Chemical Oxygen Demand (COD)

The COD measurements were preformed according to Standard Methods section 5220D (APHA *et al.* 1998), closed reflux colorimetric method. Samples were preserved by acidification to pH of less then 2 through the addition of H_2SO_4, and then digested in HACH block digester for 2 hours prior to measurement. For soluble COD samples, 0.45 μm cellulose nitrate membrane filters were used to remove any suspended solids prior to digestion. Sample absorbance readings were performed at 600 nm using a HACH DR-2000 spectrophotometer. The samples were then compared to a standard curve to determine their COD values.

4.3.2 Volatile Fatty Acids (VFAs)

Sample preparation for VFAs first involved filtration through 0.45 μm cellulose nitrate membrane filters, preservation of the filtrates with one drop of 5% phosphoric acid (H_3PO_4) and refrigeration at 4 °C until analysis. Analysis of samples was conducted on a Hewlett Packard 5880A series gas chromatography (GC). The carrier gas used was helium with an injection port temperature of 150 °C and 200 °C at the detector. The column used was 2 mm SUPELCO CARBOPAK C with 0.3% CARBOWAX 20M/0.1% H_3PO_4. The response peaks were compared with standards to determine the concentrations.

4.3.2 Orthophosphate (ortho-P), Nitrite/nitrate (NOx)

Sample preparation for orthophosphate and nitrite/nitrate involved filtration through 0.45 μm cellulose nitrate membrane filters, preservation of the filtrates with phenyl mercuric acetate and refrigeration at 4 °C until analysis. Measurements were conducted through the use of Lachat Quikchem 8000 Automated Ion Analyzer, where sample concentrations were determined in accordance with guidelines given in the manual.

4.3.3 Ammonia ($NH_3 + NH_4^+$)

The Lachat QuickChem 8000 Automated Ion Analyzer was used to measure the ammonia levels in the samples. The samples were first filtered through 0.45 μm cellulose nitrate membrane filters, preserved with three drops of 5% sulfuric acid (H_2SO_4) to pH 2, and stored at 4 °C until analysis. Sample concentrations were then determined in accordance with guidelines given in the Quikchem Automated Ion Analyzer manual.

4.3.4 Biogas Composition, Gas Count

Biogas volume production was recorded daily through the use of a submerged gas tip meter. The number of tips multiplied by the volume per tip would be the gas production per day. The volume of gas required per tip was determined by measuring the volume of gas injected into the meter to activate a tip. Biogas for composition analysis was collected with glass collection vessels connected to the head gas discharge port and transported to the lab for GC analysis. The GC used in this case was the Fisher-Hamilton Gas Partitioner using helium as the carrier gas.

4.3.5 Suspended Solids (SS/VSS, MLSS/MLVSS)

SS/VSS and MLSS/MLVSS concentrations for both the influent and the reactor mixed liquor were determined by the total suspended solids and volatile suspended solids procedure as prescribed by the 20th edition of Standard Methods (APHA *et al.* 1998). Known sample volumes were first filtered through 1.5 μm Fisher Brand Glass Fibre Filter Circles and then dried at 105 °C in an oven (VWR Scientific 1350 FM Forced Air Oven) for one hour to determine the SS/MLSS. The samples were then transferred to a 550 °C furnace (Lindberg Furnace) for another hour to determine the VSS/MLVSS concentration.

Particle size analysis was conducted with the Malvern Mastersizer 2000 that has a size spectrum range of 0.02 μm to 2000 μm. The sludge sample was first diluted with de-ionized water in the mixer/injector

unit and then fed through the analyzer for analysis. An estimated particle refractive index most suitable to biomass was selected. Sample readings in the analyzer were taken three times, with a resulting average presented in a graphical form. The particle size distribution is presented in terms of percent volume with respect to the particle size. Samples of AnMBR contents were obtained from sludge wasting, which was about 100 ml per week from the reactor sampling port.

4.3.6 Membrane Flux

Membrane fluxes were measured by taking the volume produced over a set period and then dividing it by the membrane surface area and the sampling time. The period of time measured was the permeating portion of the membrane on/off cycle. This was to ensure that the flux variations between the beginning and the end of the cycle were averaged out. This was needed due to the higher flux measured at the beginning of the cycle as compared to the end, due to reversible fouling.

4.3.7 Reactor pH

Reactor pH was measured using a Cole-Parmer AgCl double-junction pH electrode and meter. The meter was calibrated using standard solutions with pHs of 4.0, 7.0, and 10.0. The pH of the system was not measured in the reactor, but instead from the membrane effluent line on a daily basis.

4.3.8 Reactor Temperature

Reactor temperature was measured with a Cole-Parmer pipe-plug temperature probe coupled to a Dynapar T506 temperature/process controller. Temperature readings were taken manually once per day and calibrated against laboratory Cole-Parmer glass thermometers monthly with two different temperatures.

4.4 Wastewater Characteristics

Wastewater used in this research came from the UBC Environmental Engineering Pilot Plant. The pilot plant obtains its wastewater by pumping from a sump installed in a nearby municipal trunk sewer. The

wastewater was then stored in two 9,500 L equalization tanks with gentle vertical-shaft mixing. These tanks are at ground level and serve to equalize somewhat the fluctuating organic and hydraulic loadings of the sewer system. Intake pumping occurred four times a day at 09:00, 15:00, 21:00 and 03:00 hours. From the storage tanks, the raw wastewater was pumped to a pilot scale primary clarifier and then through a 2 mm stainless steel screen prior to entering the submerged AnMBR. Table 4.4 presents an average value for each parameter measured in the feed wastewater (with out the addition of supplemental acetate) after the primary clarifier and screens.

4.4.1 Wastewater Feed

Reactor wastewater feed rate was controlled by a timer that activated the feed pump (Masterflex L/S) with a maximum flow rate of 270 mL/min. A relatively high flow rate was used to reduce the wastewater residence time in the feed tube leading to the reactor, to minimize COD reduction by growth within the tube. The required timing for the feed system depended on the operating OLR.

Table 4.4 UBC Pilot Plant wastewater characteristics

	Range	**Average**
Total COD (mg/L)	200 – 480	300 ± 60
Soluble COD (mg/L)	50 – 293	130 ± 40
TSS (mg/L)	44 – 148	86 ± 6
Total P (mg/L)	2.82 – 5.86	4.2 ± 0.1
Ortho-P (mg/L)	1.40 – 3.82	2.8 ± 0.2
TKN (mg/L)	25.9 – 44.2	32 ± 1
Ammonia (NH_4^+) (mg/L)	20.8 – 36.1	29 ± 1
Nitrite/nitrate (NO_x) (mg/L)	0.035 – 0.309	0.09 ± 0.02
pH	6.6 – 7.6	7.2 ± 0.1
Alkalinity (mg/L as $CaCO_3$)	175 – 325	250 ± 34
Temperature (^{0}C)	16.8 - 23.1	20.3 ± 0.3

5.0 RESULTS AND DISCUSSION

This chapter presents and discusses the results of this study in three sections: 5.1 AnMBR System Start-Up; 5.2 AnMBR Acclimatization and Steady state Operation; and 5.3 Membrane Fouling.

The AnMBR system start-up discussion presents the details of system initiation. During this phase, the reactor was operated under batch conditions due mainly to mechanical hindrances (i.e. necessary equipment such as the level switch and the remaining peristaltic pumps for the continuous operation of the AnMBR were not yet in place). The reason why operation of the submerged membrane AnMBR was initiated prior to full completion was due to the fact that the external parallel tubular membrane AnMBR was fully constructed and had already been put into operation. The only way the submerged membrane reactor could operate in parallel within the same timeframe was to start under batch conditions until the necessary pumping equipment was in place. This section will discuss the initial two months of AnMBR operation.

During the first part of the AnMBR acclimatization and steady state operation phase, the objective was to acclimatize anaerobic biomass towards the treatment of municipal wastewater by providing favourable conditions in terms of temperature and substrate composition. Initial operating conditions during this phase were geared towards an environment ideal for anaerobic biomass growth. As the effect of lower operating temperatures on the system was one of the primary focuses of the study, its effect on the operation of the membranes was also examined. The system temperature was reduced near the end of the acclimatization phase in preparation for steady state operation. The AnMBR steady state operation phase examined the treatment of municipal wastewater under stabilized reactor conditions, after terminating the addition of supplemental acetate. During this phase, removal efficiencies of different measured parameters were examined following each incremental increase in the OLR. Last but most importantly, the impacts on the submerged membranes through changes in operational parameters were

carefully examined to gain a better understanding of the mechanisms governing membrane flux. In the section for membrane flux recovery, different methods of membrane cleaning were explored through the duration of the project. The type of chemicals and mechanisms used during cleaning gave some insight into the composition of foulant deposited on the membrane.

Finally, membrane fouling was examined through a series of filtration tests on anaerobic and aerobic mixed liquor. The tests helped determine the type of membrane fouling that occurred with respect to the mixed liquor filtered.

5.1 AnMBR System Start-Up

Initiation of the submerged membrane reactor operation occurred on October 19, 2004. Start-up of the submerged AnMBR required anaerobic biomass from a municipal anaerobic sludge digester as recommended by the WERF project sub-committee (PSC). The seed for the system was obtained from the Lulu Island Secondary Wastewater Treatment Plant, in Richmond, BC. Once the AnMBR system was purged of air with nitrogen, a mixture of sludge and wastewater (based on sludge MLSS concentration) were combined with sodium acetate (in powdered form, dissolved in solution for a source of acetate) to achieve a final reactor sodium acetate concentration of about 300 mg/L, and a reactor MLSS concentration of 5,000 to 6,000 mg/L. The anaerobic digester sludge obtained was first screened with a 2 mm screen and then seeded on the same day it was acquired.

The initial value suggested by the PSC for the MLSS concentration in the systems was 10,000 mg/L. However, based on the expected biomass yield for an anaerobic system and the raw wastewater characteristics, it was expected that a maximum MLSS concentration of only 7,000 to 8,000 mg/L could be achieved. Clogging of the relatively small tubing in the lab scale system was observed during start-up testing when the MLSS concentration in the system exceeded 8,000 mg/L. Therefore, the initial target operating MLSS concentration was set to 8,000 mg/L.

System feed first required daily effluent withdrawal through the membranes of about 1 L. In order to maintain robust biogas production during start-up, measured amounts of sodium acetate in the form of dry powder were mixed in with 1 L volumes of wastewater to achieve a reactor acetate concentration of 300 mg/L through fed batch operation each day. Temperature, pH, COD, and gas composition were the controlled and monitored parameters. Initial reactor operating temperature fluctuated between 25 and 30 °C despite temperature control through the use of a thermocouple and heater. This issue of temperature fluctuation was later resolved on day 154 after replacing the thermocouple that was subsequently determined to be faulty. The pH of the reactor remained steady near 7.8 without much fluctuation during this period, as discussed in the following subsection entitled, **Temperature and pH** (5.2.2.3).

Initial COD concentrations for the effluent during start-up were in the range of 3000 to 4000 mg/L, which was extremely high. The presence of a light brown tint in the membrane effluent suggested high levels of soluble solids within the reactor. This was expected due to the nature of digester sludge, and with the low hydraulic loading rates used during this phase, the system could not rapidly flush out existing or additional soluble solids present in the system. Gas production during the batch mode operation was relatively unsteady and fluctuated between 10 and 35 $L/d \cdot m^3$ reactor volume. However, much of this instability was attributed to the constant modifications associated with the reactor operation. This was also the case for the analysis of biogas composition. Preliminary results indicated a high percentage of N_2 in the gas, which was attributed to the repeated use of N_2 as a method of O_2 displacement in the reactor headspace. During the 60 days of fed batch operation, N_2 composition declined from 45% to near 15% of the total AnMBR headspace gas while CH_4 increased from 45% to 80%. Percent CO_2 in the headspace dropped from 10% to 6% over that same period. The gas data provided a good indication that the substantial population of methanogens requisite for continuous feed operation was present. Results of the gas production and composition can be seen in the following subsections entitled, **Gas Production Rate** (5.2.2.6) and **Gas Composition** (5.2.2.7). When the

necessary equipment permitting the conclusion of operation in the batch feed mode was installed, the OLR of the reactor was increased, as detailed in the next section.

5.2 AnMBR Acclimatization and Steady State Operation

5.2.1 Operation Overview

The submerged membrane system started with continuous feed on the 18th of December, 2004 (day 61). The brown tint initially observed during the fed batch operation phase cleared up once a sufficient feed volume had passed through the system in the few days following the continuous feed initiation. During this period, effluent COD concentrations decreased considerably from about 3000 mg/L to 100 mg/L in 30 days (refer to **Appendix B: Raw and Calculated Data**). Supplemental acetate was utilized for the first 175 days of operation. Once acetate supplementation was terminated on day 175, the influent feed flow was increased to compensate for the reduced COD loading.

The reactor initially started operation with only one US Filter membrane, while the Toray membrane received additional modifications allowing for its operation in the reactor. Operational parameters monitored for the US Filter membrane included the gas sparging rate, the permeation on/off cycle, and the TMP. The sparging rate of the membrane was initially set at 10 LPM (litres per minute), the permeation cycle at one minute on and one minute off, and the TMP at 6.8 kPa. The theoretical membrane operating mode was one of constant flux, at variable pressure. The gas sparging rate used was nearly twice the manufacturers' recommendation, but was chosen due to the fact that the recommended values were based on aerobic treatment and more extensive fouling was expected for an anaerobic system. The same applied to the conservative on/off cycle, which was selected to combat fouling that was expected to be more extensive under anaerobic conditions. These parameters changed over time as the study evolved to meet certain operating objectives. The impacts of these changes are presented in the following sections.

5.2.2 AnMBR Performance

5.2.2.1 Organic Loading Rates

The target values for the OLR of the AnMBR are presented in Figure 5.1 along with the actual operating values achieved. By using grab influent samples for the evaluation of total influent COD, the values for OLR were calculated based on the COD loading per unit volume reactor. During the first 61 days of operation, the influent sewage was stored in a tank and changed daily, while its COD concentration was augmented with 300 mg/L of acetate. The reactor was then fed intermittently (on a cycle determined by the desired target OLR, and thus the daily feed volume) with a timer-controlled pump, to dispense the desired influent volume each day. The conclusion of batch operation on day 61 was followed by continuous flow operation at an OLR of 0.75 kg COD/m^3·d (with a reactor volume of about 26 L), with the sodium acetate supplementation. Further increases in OLR required either a decrease in reactor liquid volume, or an increase in influent flow rate, or both. With acetate in the feed, incremental increases in OLR were easily achieved until day 175.

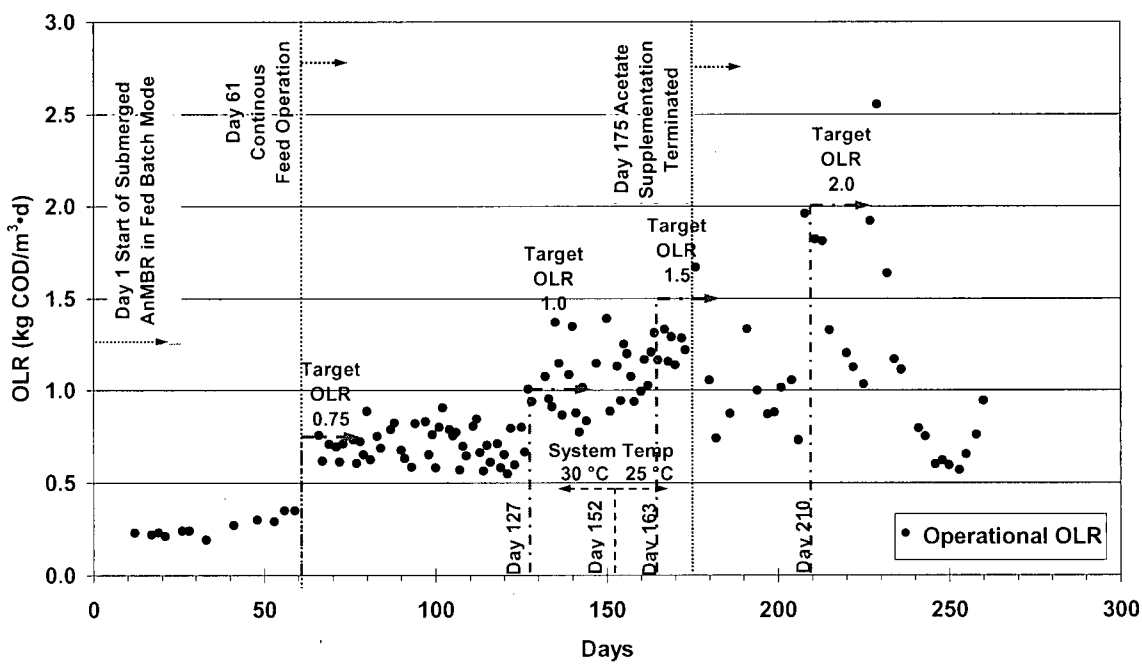

Figure 5.1 Target vs. operational OLR in the submerged AnMBR.

The termination of acetate supplementation after day 175 severely retarded the continual increase in OLR due to the withdrawal of nearly 40% of the influent COD. Consequently, nearly twice the influent flow was required to sustain the OLR of about 1.25 kg COD/m^3·d achieved prior to acetate withdrawal. However, for reasons which pertain to the performance of membranes (discussed in detail in the subsection entitled, **Membrane Fluxes** (5.2.2.4)), it was evident from Figure 5.1 that the system could not sustain an additional increase when an OLR of 1.5 kg COD/m^3·d was attempted on day 175. In this case, irreversible membrane fouling restricted the hydraulic loading rate of the system.

5.2.2.2 Hydraulic Retention Time

Figure 5.2 presents an overview of the operational HRT of the AnMBR in its entirety. The stepwise decreases correspond to the imposed increases in OLR until day 175. The treatment limits of the system were generally expressed in terms of OLR, which was highly correlated with the HRT. Despite the intended setup of the membrane system for constant flux, variable pressure operation, the system was not able to achieve this requirement. In the event of a significant foulant build-up on the membrane, flux decreased significantly in conjunction with a rise in TMP. Consequently, the TMP quickly approached the limit of the membrane manufacturers' suggested operating vacuum pressure, which prompted restorative action. Several methods of remediation were utilized with varying success, the key of which was thorough membrane cleaning. Each of the solid lines on the figure with corresponding dates indicates a modification to the membranes in the reactor. This may have been either be the installation of new membranes or a simple membrane cleaning for flux recovery; all of which are described in further detail in Table 5.1.

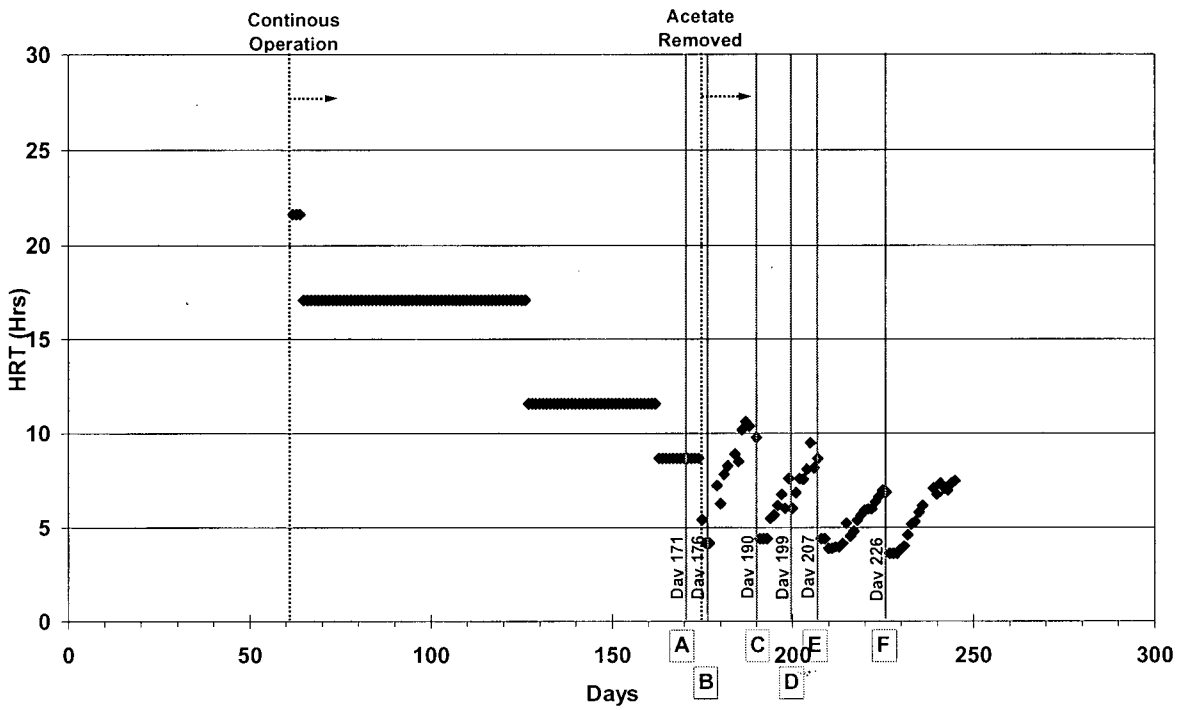

Figure 5.2 Submerged AnMBR operating hydraulic retention time.

Table 5.1 System membrane modification events

Event	Day	Details
A	170	Replaced US Filter No.2 with the previously used US Filter No.1 membrane treated with cleaning method I*. Toray membrane cleaned with cleaning method I.
B	176	Replaced US Filter No.1 with No.2 which was removed at event A and cleaned with method I* before re-installation. Toray membrane rinsed with water.
C	190	Installed US Filter No.1 so both No.1 and No.2 operating in same reactor along with Toray. Prior to installation, all membranes (US Filter No.1, 2, and Toray) treated with cleaning method II*.
D	199	All membranes removed and subjected to cleaning method III*.
E	207	All membranes removed and subjected to cleaning method IV*.
F	226	All membranes removed and subjected to cleaning method V*.

* Refer to Table 4.3

The HRT unexpectedly increased beyond the 9 hours set point on day 185. This may be the direct result of a combination of higher loading of influent suspended and/or colloidal solids, or an increased presence of SMP due to a higher colloidal content (Langenhoff *et al.*, 2000), both of which contribute to

membrane fouling. An increase in solids concentrations was observed after day 175 which seems to coincide with this explanation and is further discussed in the subsection entitled, **Suspended Solids Concentration** (5.2.2.3). The presence of colloidal solids was verified through the use of the particle size analyzer as further described in the subsection entitled, **Particle Size Distribution** (5.2.2.10). Additionally, surface characteristics of the membrane may have been altered after a cleaning cycle, which could explain the dramatic increase in the rate of fouling and thus compromising the achievable HRT. This is a very likely cause of membrane flux decrease as the subsection entitled, **Membrane Fluxes** (5.2.2.4) further describes.

The most probable cause for the flux decrease leading to an increased reactor HRT was the influence of the increased concentration of suspended/colloidal solids in the mixed liquor, and changes in the characteristics of the membrane surface after a cleaning cycle.

5.2.2.3 Temperature and pH

As presented in Fig 5.3, the system pH remained relatively constant as the feed used in this study was shared with another ongoing project at UBC which required the addition of bicarbonate buffering for the process of nitrification. Acidification through the breakdown of complex organic wastes, proteins, carbohydrates, fats, and oils by acetogenic bacteria has the potential of upsetting the system equilibrium and disrupting the delicate methanogen population through a decrease in pH. However, this was not the case as seen from Figure 5.3, as the pH did not decrease below 6.5, which is regarded as the minimum pH required for a sustainable population of methanogens. This stable pH may have been due to: (1) the presence of a large population of methanogens that was capable of utilizing the VFA produced; or (2) the dilute nature of the feed precluded high concentrations of hydrogen. Either way, the pH of the system dropped from 7.8 (±0.1) to 7.3 (±0.1) after the transition out of batch mode operation and ultimately to 6.7 (±0.2) following the termination of acetate supplementation. The higher initial pH may have been due to a greater presence of alkalinity originally established in the anaerobic digester from the Lulu

Island Secondary Wastewater Treatment Plant from which the reactor seed mixed liquor was originally obtained. This buffering capacity was eventually washed away after the initiation of the continuous feed mode.

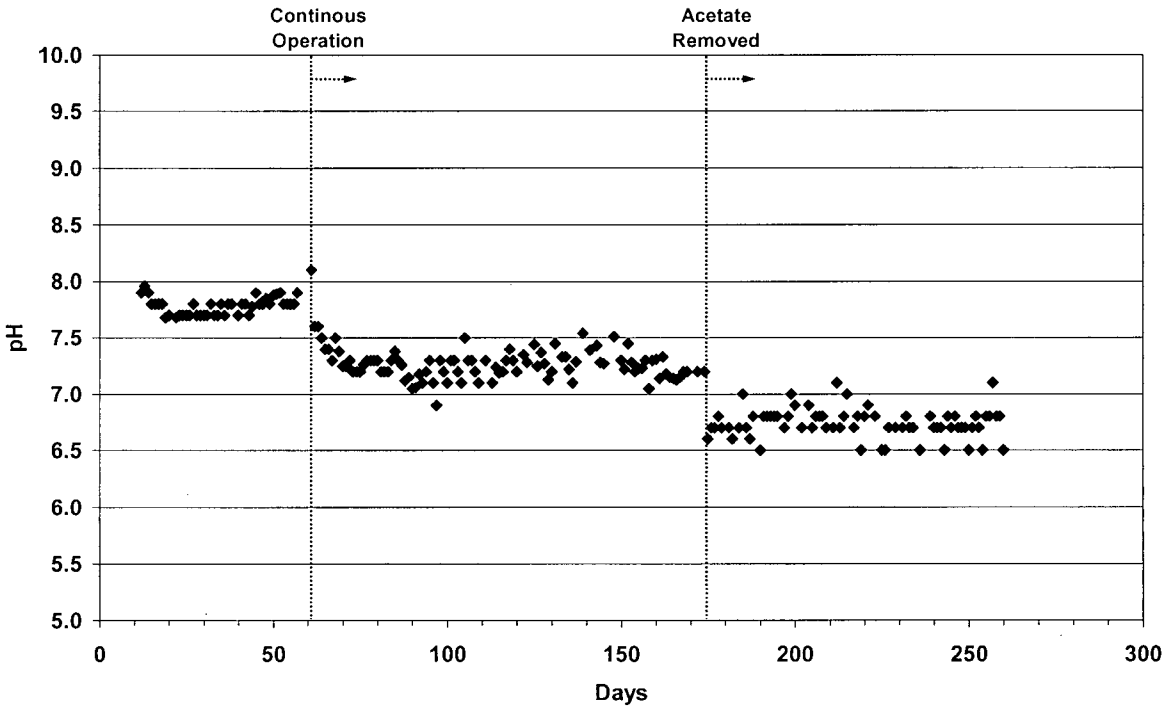

Figure 5.3 Submerged AnMBR pH readings.

As presented in Figure 5.4, a temperature decrease from 30 (±2) °C to 25 (±1) °C was imposed beginning on day 152 over the span of two days. This was the first step at reducing the reactor temperature as the focus of the study was the feasibility of wastewater treatment at lower temperatures. In retrospect, the 2 day period was much too short for proper acclimatization of the biomass. The impact of the abrupt temperature change is reflected on, among other parameters, the COD removal and the biogas production. This can be seen in the subsections entitled, **Chemical Oxygen Demand** (5.2.2.5) and **Gas Production Rate** (5.2.2.6). The effect of temperature on membrane flux is discussed in the following subsection entitled, **Membrane Fluxes** (5.2.2.4). The fluctuations in temperature experienced before day 110 were due to a faulty thermocouple while subsequent readings were verified by a

thermometer. The thermocouple was replaced on day 154 and the ensuing fluctuations were the result of the limited accuracy achievable by the thermocouple controlled heating element.

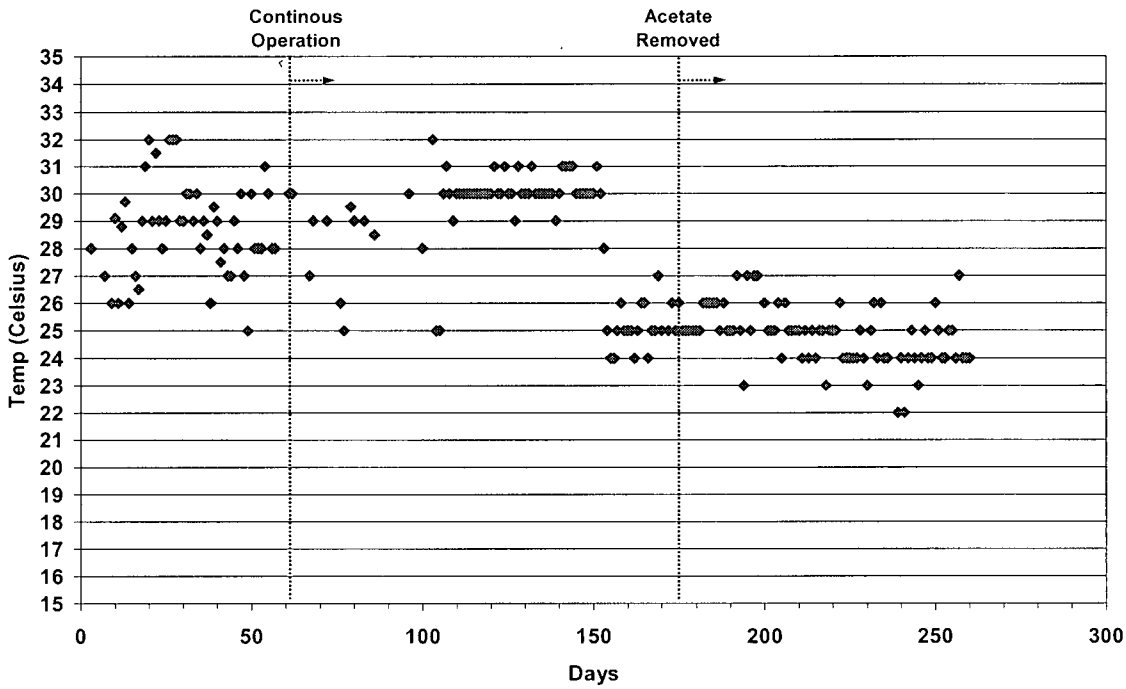

Figure 5.4 Submerged AnMBR temperature readings.

5.2.2.4 Membrane Fluxes

Figure 5.5 summarizes the performance flux of the membranes used during this study, while Figure 5.6 presents the corresponding TMP. Continuous reactor operation started with the use of only one new US Filter hollow fibre membrane module, which exhibited a starting flux of 19 Lmh at a TMP of 7 kPa. This was the second of two modules (hereby referred to as US Filter No. 2) available for this study. The first module was used during the batch period of operation, after which it was removed, and the flux was recovered by cleaning type I (Table 4.3) for later reinstallation. The membranes were operated with an on/off cycle of 1 min on, 1 min off, in conjunction with an initial gas sparging rate of 20 Lpm/m^2 (litres per min per m^2 of membrane surface area) until day 121. This sparging rate was twice the manufacturer's recommended value. Following day 121, the on/off cycle was increased to 2 min on, 1

min off, and the OLR was increased to 1.0 kg COD/m^3·d. With the exception of the installation of the Toray flat sheet membrane (day 143), no other significant changes were made to membrane operation during this period, which was concluded with the decrease in temperature on day 152.

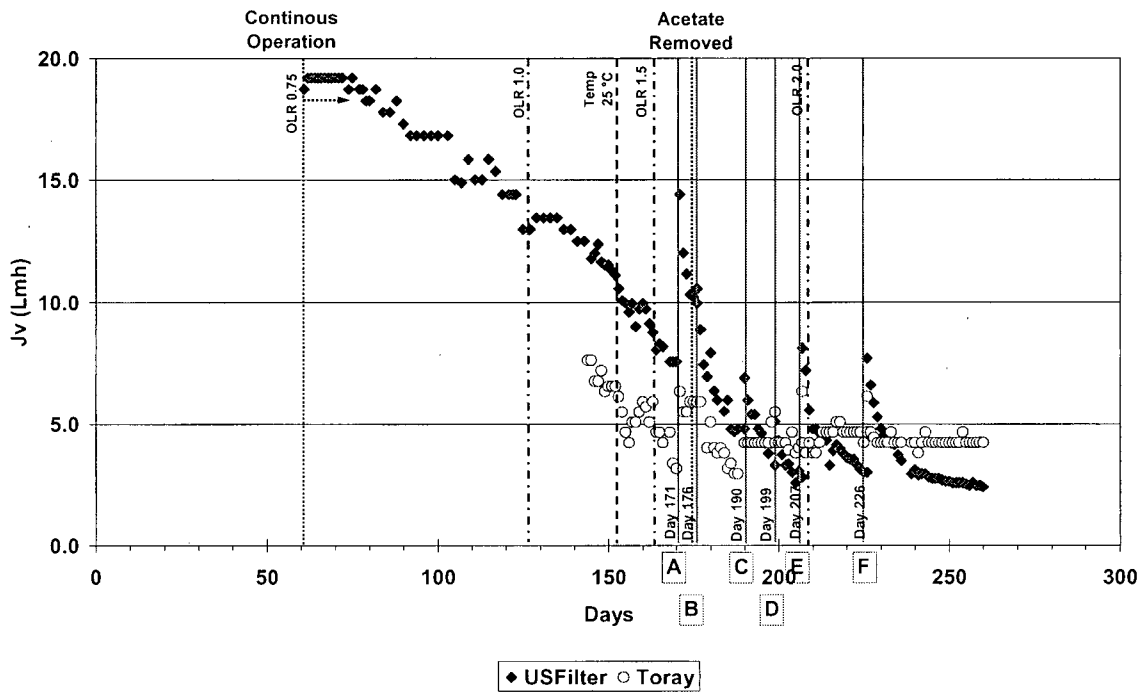

Figure 5.5 Submerged AnMBR flux for the US Filter and Toray membrane. Refer to Table 5.1 for membrane modification description.

As shown in Figure 5.5, membrane flux decreased in a linear fashion during this time, with the TMP mirroring a linear increase (Figure 5.6). The calculated rate of flux decline prior to the temperature drop was -0.09 Lmh/day. This rate of decrease doubled to -0.18 Lmh/day immediately after the decrease in temperature. This observation was similar to the findings by Hogetsu *et al.* (1992) who reported that a considerable increase in flux was observed as the system temperature was raised by a few degrees. Factors contributing to this decrease may have included a reduction in permeate viscosity as well as a probable increase in SMP production at lower temperatures, as observed by Barker *et al.* (2000) and Schiener *et al.* (1998). Similar results could not be concluded for the Toray membrane, due to insufficient data during the same period.

The on/off cycle was eventually increased to 4 min on, 1 min off (Wen *et al., 1999)* on day 163 without much measurable impact on the flux. This was to increase the membrane effluent flow and subsequently, decrease the HRT, without severely increasing membrane fouling.

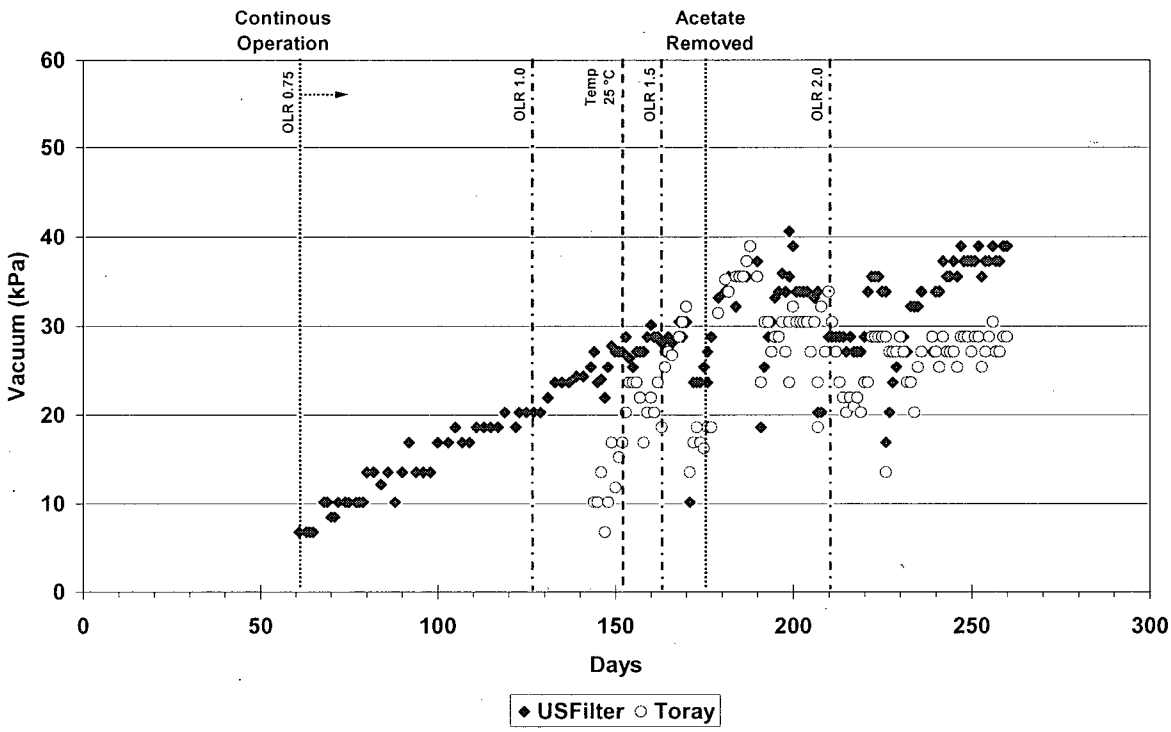

Figure 5.6 Submerged AnMBR trans-membrane pressure.

The Toray membranes started operation on day 144 with an initial flux of 7.6 Lmh at 10 kPa vacuum. The initial gas sparge rate was 22 Lpm/m^2 while utilizing the same operating on/off cycle as the US Filter membrane for the duration of the study. Though there were inadequate data to determine the effects of temperature change on the Toray flux, a slight trend similar to that observed for the hollow fibre membrane could be seen from the data that were available. Furthermore, it appears that the initial flux decline of the Toray membrane was higher than that of the US Filter membrane. The increase in flux on day 157 was due to the increase in gas sparge rate to 28 Lpm/m^2, but thereafter, the flux soon again declined. The average rate of flux reduction during the first 30 days was about -0.25 Lmh/day. This represented a 40% greater rate of flux decline relative to the US Filter membranes. This is likely

due to the fact that the gas sparged below the Toray membranes was not as evenly distributed as that sparged below the US Filter Membrane.

Each of the alphabetical designations in Figure 5.2 and 5.5 represents an alteration to the membrane(s) of the system, as detailed in Table 5.1. The membrane flux for the US Filter membrane module No.1 (used prior to day 61 for batch operation) recovered to 75% of the new membrane flux after event A (Replaced US Filter No.2 with the previously used US Filter No.1 membrane treated with cleaning method I. Toray membrane cleaned with cleaning method I.). Prior to event A, the permeate flux had reduced to 10% of its original value. The recovery value for the Toray membrane was 83% after event A, from a permeate flux reduction to 42% of its original value. Despite the significantly shorter operating time, the Toray flat sheet membrane fouled much faster than the hollow fibre membranes. However, this may again be attributed to the less efficient (non-uniform, due to a more complex design) gas sparging system on the Toray membrane.

Flux decline in the days immediately following event A were severe in both membrane systems but especially so in the US Filter module No.1. The observed flux decreased to about 23% of the new membrane flux in the first 6 days after event A. This rate of fouling would have normally taken about 55 days when the new membrane was first installed. This indicates that: (1) the cleaning procedure used was inadequate for the type of fouling encountered; or (2) the surface characteristics of the membrane had changed after the type of cleaning used; or (3) the character of the mixed liquor may have since been altered.

The purpose of event B (Replaced US Filter No.1 with No.2 which was removed at event A and cleaned with method I before re-installation. Toray membrane rinsed with water.) was to test the effectiveness of the cleaning procedure type I by reinstalling the US Filter module No.2 removed during event A. Percent flux recovery for module No.2 was only 52% when fouled to 40% of its original capacity. This

poor recovery may be attributed to the sustained run endured by module No.2 as compared to module No.1, which only ran about 55% of module No.2's duration before cleaning. This may have created more irreversible fouling though adsorption and thus made recovery difficult. The flux decline in module No.2 was severe and exhibited a "memory" like behaviour. It only took about 5 days for the flux to decline to the levels observed right before cleaning (event A) and the flux continued to decline at nearly the same rate thereafter. The rate of flux decline was calculated to be -0.36 Lmh/day, or twice that of the rate of decline after the temperature decrease. The flux recovery of the Toray and the rate of decline remained nearly the same as before following event B.

In efforts to increase the OLR of the system, US Filter module No.1 was installed in parallel with module No.2 to increase the system flow capacity at event C (Installed US Filter No.1 so both No.1 and No.2 operating in same reactor along with Toray. Prior to installation, all membranes, US Filter No.1, 2, and Toray treated with cleaning method II.). Furthermore, to improve the total system flux, the entire membrane system was treated with membrane cleaning method type II. This cleaning event was designed to coincide with the removal of acetate from the feed by increasing the flow to a level that would sustain the required OLR. Unfortunately, despite installing a second pump head to increase flow capacity, a higher flow could be sustained for only a short period of time. It was also evident from Figures 5.1 and 5.2 that the target loading rate could not be sustained. Flux for the US Filter membrane once again decreased at a relatively rapid rate. The only other change that occurred during this period was a slight decrease in gas sparging rate for the US Filter to 9 LPM each, while that for the Toray membrane remained unchanged. The gas sparging rate decrease was due to the insufficient capacity of the gas recirculation pump.

The Toray membrane, responded well to a type II cleaning (event C) from what can be observed in Figure 5.5. Despite a flux recovery to only 55% of the original capacity, the Toray membrane was able to maintain this value for the duration of the study. It appears that after 46 days of operation, the Toray

membrane seemed to have reached a steady state permeate flux. The removal of biomass on the membrane through caustic cleaning used in conjunction with acidic treatment for the removal of possible struvite deposition may have been what was required. During sustained operation, the Toray membrane out-performed the hollow fibre membranes of the US Filter after day 190. However, this conclusion cannot be firmly drawn due to the fact that the Toray membrane started operation much later than the US Filter system. Unfortunately, this flux-sustaining capability of the Toray membrane could not support the desired OLR because its overall output (i.e. surface area) was too low compared to that of the US Filter membranes. Furthermore, the packing density of the Toray membranes was also much smaller in terms of surface area to membrane volume occupied in the reactor. This limited the installation of additional Toray membranes. Consequently, the desired OLR of 2.0 kg COD/m^3·d could not be reached.

Events D, E, and F were attempts in vain to restore the flux to acceptable levels. Despite this setback, useful information was obtained during the process of experimenting with different membrane cleaning techniques. Figures 5.7 and 5.8 present the flux recovery through each incremental cleaning cycle for cleaning types IV and V. Membrane flux was measured following each cleaning cycle.

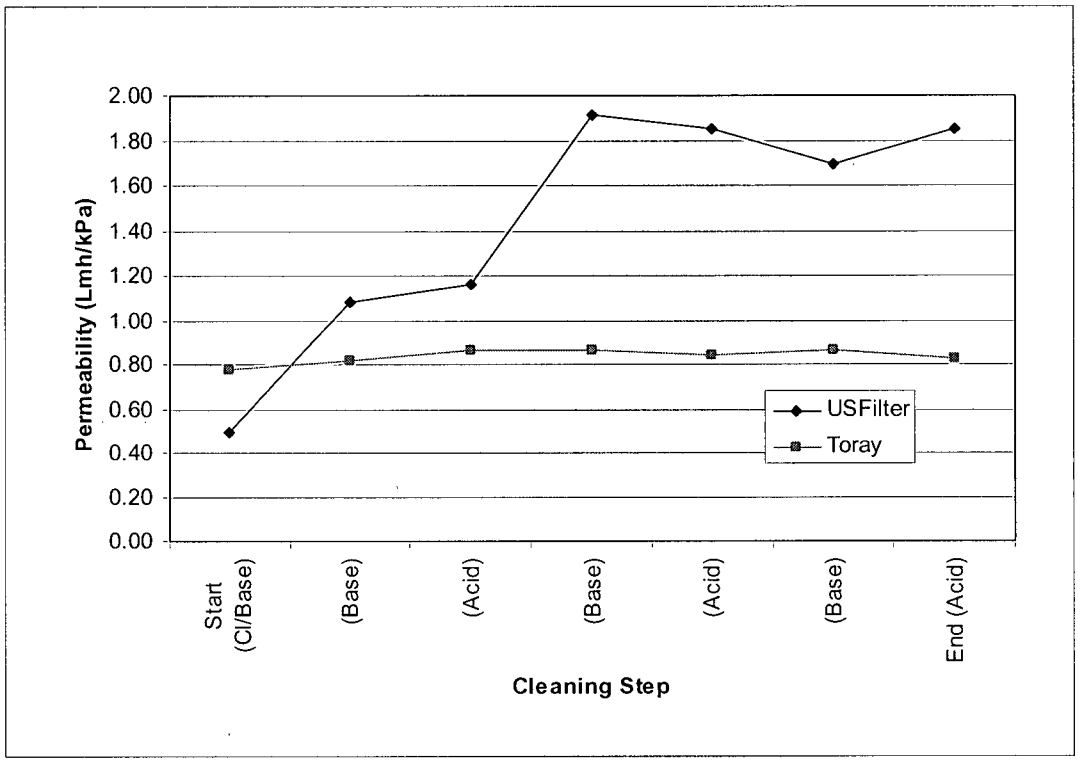

Figure 5.7 Cleaning Type IV incremental flux gain.

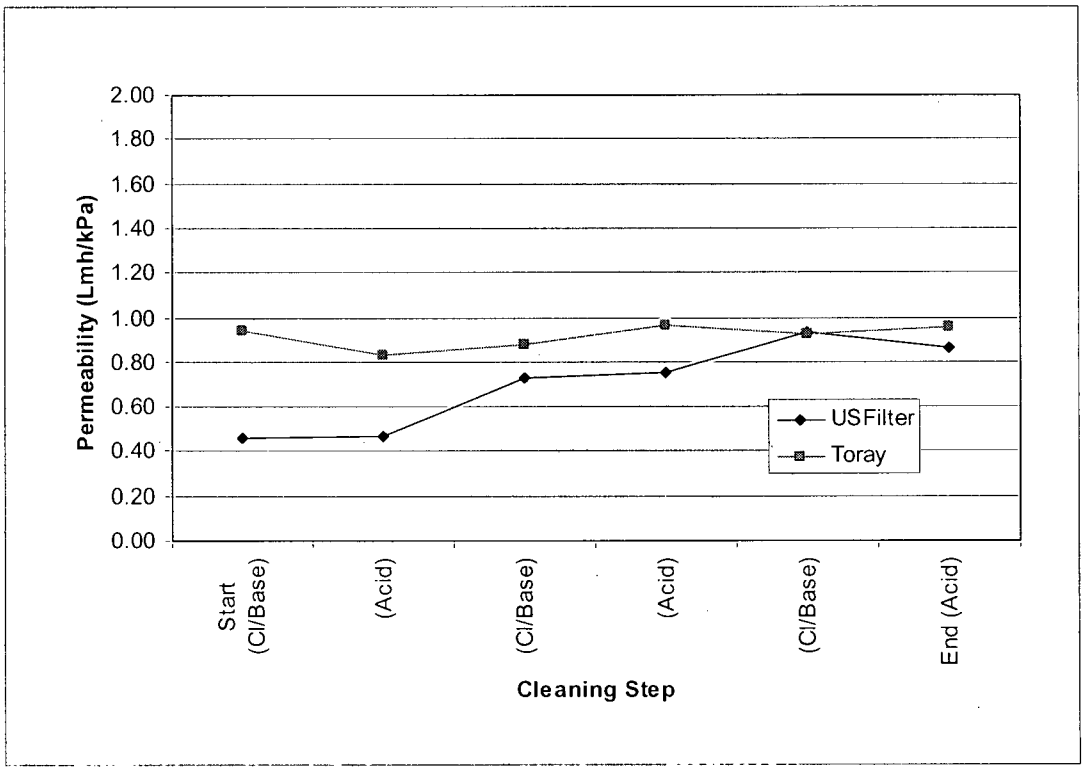

Figure 5.8 Cleaning Type V incremental flux gain.

Cleaning methods IV and V, described in Table 4.3, where the most comprehensive in terms of addressing the different issues of fouling. The figures show flux gains after each step of the cleaning procedure for the US Filter membrane, which was not the case with the Toray. This suggests that membrane fouling for the hollow fibre membrane may have resulted from layers of alternating biomass and struvite (Choo and Lee, 1996a), while the mechanism for flat sheet fouling may have been different in nature. The presence of struvite in the reactor was likely, given that the three components required for its precipitation, ammonia, phosphate, and magnesium were generally readily available in the anaerobic environment and through influent feed. It was evident that chemical flux recovery cleaning did not seem to have a significant impact on the flat sheet membrane. Instead, most of the flux recovery of the flat sheet required only mechanical removal of surface build up (i.e. use of a water jet). This was observed as the Toray membrane was mechanically cleaned with a water jet prior to cleaning type IV and V, with no significant subsequent flux recovery. This suggests that the flux of the flat sheet membrane could be effectively sustained if proper mechanical agitation was provided (i.e. uniform gas sparging). This was unfortunately not the case for the hollow fibre membrane. Despite surface foulant removal, considerable fouling still existed. Fouling due to adsorption may have been more prominent for the US Filter membrane since removal of deposited solids improved flux recovery only marginally. Flux recovery for the US Filter membrane was significantly lower in type V than IV. This may be accounted for by the different type of cleaning used (chlorine with base instead with just base). However, the more likely explanation would be that the irreversible fouling was caused by the higher SMP concentrations associated usually with increased HRTs (Barker *et al.*, 2000). Nevertheless, the different cleaning methods did not seem to impact the Toray membrane to the same extent, as its flux remained steady during both types of cleaning. Toray flux remained constant throughout the rest of the study.

The TMP of the Toray membrane (Figure 5.6) increased at a much greater rate than the US Filter membrane. This likely occurred as a result of surface foulant build-up which was not efficiently removed by the sparging gas regime applied to the Toray membrane. Interestingly, this could have

ultimately resulted in the "self-rejective dynamic membrane" observed by others in the literature (Imasaka *et al.*, 1989). However, because the Toray and US Filter membranes were operating at different permeate fluxes and sparging regimes, it is difficult to compare the rates of TMP increase. Fortunately, permeability, which incorporates both the permeate flux and the TMP, can be used to compare the performance of both membranes. The results in Figure 5.9 demonstrate the permeability of the US Filter membrane, which was initially higher than the Toray membrane, but eventually decreased below the Toray membrane beyond day 195. The permeability for both membranes remained relatively low after day 180.

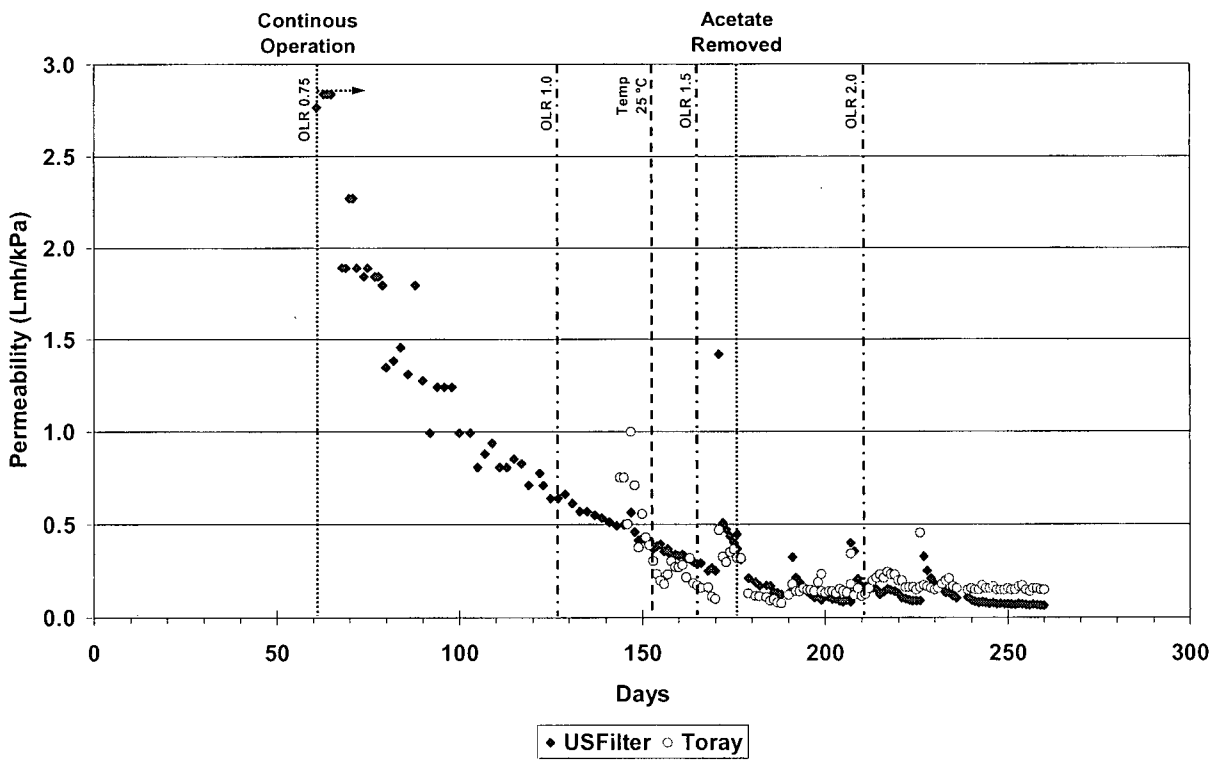

Figure 5.9 Submerged AnMBR permeability comparison.

5.2.2.5 Chemical Oxygen Demand

As presented in Figure 5.10, the average influent COD with supplemental acetate was 496 (±70) mg/L, while the average soluble influent COD was 334 (±39) mg/L during the same time. The average influent COD without supplemental acetate was 300 (±60) mg/L, while the average soluble influent COD was 130 (±40) mg/L. The effluent COD concentration of the system decreased steadily from a high of 230 mg/L on day 67 to an average of 75 (±20) mg/L between days 100 to 152. The COD data discussed in the remainder of this section are the data collected after day 100 when the system had presumably reached "steady state" in terms of COD removal.

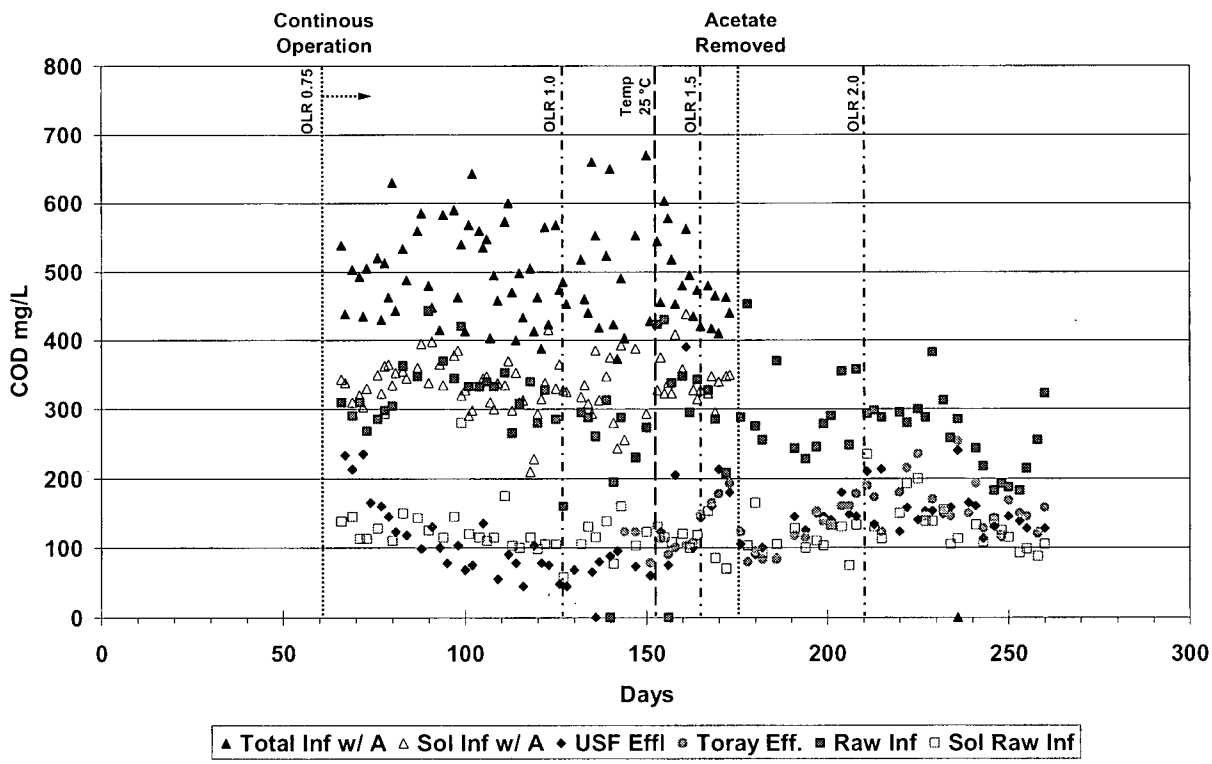

Figure 5.10 Submerged AnMBR system COD. Total Inf w/A is influent with acetate supplementation. Sol Inf w/A is soluble influent with acetate supplementation. Raw Inf is influent with no acetate supplementation. Sol Raw Inf is soluble influent with no acetate supplementation.

The overall removal of the total influent COD during the period between days 100 to 127 with acetate supplementation (where OLR is 0.75 kg COD/m^3·d) was about 84%, or an effluent COD concentration

of about 80 (±25) mg/L. The effluent may have contained a significant amount of dissolved CH_4 that could have contributed to the effluent COD concentration. It was estimated that on average, 16 mg/L of CH_4 could have been dissolved in the permeate during this period (see **Appendix A: Calculations**). Since every mole of CH_4 contributes 64 g of COD (Tchobanoglous *et al.*, 2003), the CH_4-COD concentration in the permeate, assuming no losses during sample collection and transportation, could have been as high as 64 mg/L. This calculated value is roughly equal to the average effluent COD of the system during this period. However, uncertainties lie within the solubility coefficient of CH_4 in anaerobic mixed liquor and the amount of CH_4 released from solution prior to analysis.

The effluent COD values remained below 100 mg/L for the period between day 127 and 152, after which the system temperature was decreased. The OLR during this period was about 1.0 kg COD/m^3·d, with an average percent COD removal for the US Filter membrane of about 86%, or an average effluent COD of about 72 (±16) mg/L. The initial Toray effluent COD concentrations were slightly higher (about 108 (±26) mg/L) than the US Filter but this could have been from: (1) the variability in sampling; or (2) a breach in membrane integrity; or (3) the rejective nature of the membrane itself. Regardless, effluent COD concentrations in both permeates were in fairly close agreement after day 152, with the exclusion of outliers.

As documented many times in the literature (Kalogo and Verstraete (1999) and Seghezzo *et al.* (1998)), COD removal in an anaerobic system depends on its operating temperature. This system was without exception. When the temperature was decreased to 25 °C on day 152, the effluent COD concentration increased to approximately 100 mg/L for both membranes. This represented a percent COD removal of about 81% over the 11 days till the increase of OLR to 1.5 kg COD/m^3·d. However, it was estimated that at 25 °C, the average concentration of dissolved CH_4 was about 18 mg/L in the effluent (see **Appendix A: Calculations**), which corresponds to a COD value of about 72 mg/L.

Following the increase of OLR to target 1.5 kg COD/m^3·d on day 163 (where the actual OLR was about 1.25 kg COD/m^3·d due to membrane permeation capacity issues mentioned in the previous section), a removal efficiency of about 64% was observed until the removal of supplemental acetate on day 175. Inactivation of the biomass due to a temperature shock coupled with the increase in influent COD may have played a significant role in this result. Another cause may be an increased production of SMP associated with a drop in temperature (Barker *et al.*, 2000).

Following the elimination of sodium acetate supplementation, the effluent COD decreased immediately, but slowly increased again over time. Initial COD removal after the elimination of sodium acetate was about 68%, while COD removal right before the targeted OLR increase to 2.0 kg COD/m^3·d was about 43%. The decline in COD removal efficiency of the system parallel the decline in gas production, as noted in the following subsection entitled, **Gas Production Rate** (5.2.2.6).

The effluent COD of the system after day 175 slowly increased to values that exceeded the soluble influent COD. The filter used for the soluble influent COD was 0.45 μm compared to the 0.08 μm pore size of the membranes. This suggests that the effluent of the reactor might have contained different and substantially smaller molecules than those in the soluble portion of the influent. The measured TSS of the effluent during this period was negligible.

Effluent COD after day 210 was highly variable, due likely to the non-steady state conditions of the reactor. The overall COD removal efficiency was about 40% for the remainder of the study. The effluent COD of the Toray membrane was slightly higher after modification event E due to a minor membrane surface breach. A slight tint in the effluent could be seen immediately after the membrane wash but eventually cleared up. This may have been due to foulant build-up at the site of the breach which may have sealed it.

5.2.2.6 Gas Production Rate

Figure 5.11 presents the biogas production rate measured as total gas released from the reactor headspace. The main constituents of the biogas were CH_4 and CO_2, as detailed in the subsection entitled, **Gas Composition** (5.2.2.7).

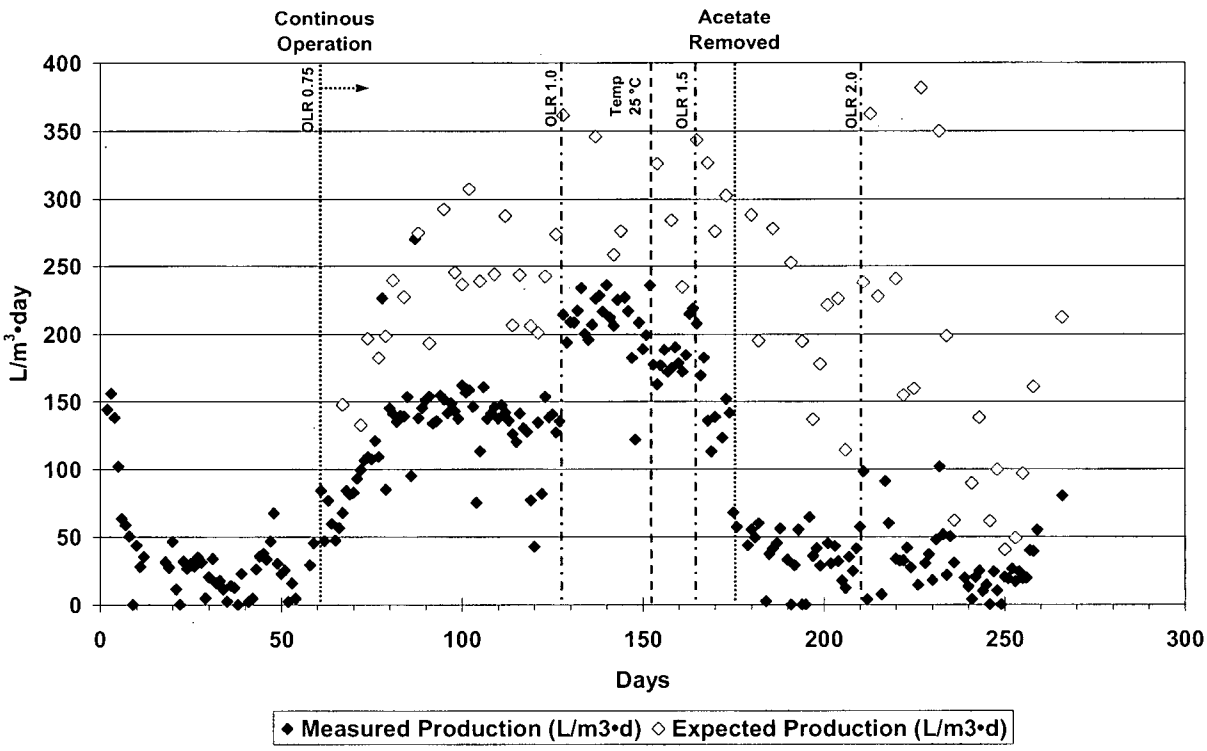

Figure 5.11 Submerged AnMBR biogas production per m^3 reactor MLSS.

Biogas production increased significantly after the start of continuous operation. Sufficient acetate in the feed made for easy substrate utilization and production of CH_4 by the aceticlastic methanogens, which over the span of 20 days (starting from day 61), reached a steady state reactor head space biogas volume production. However, biogas production during this time consistently fell short of the theoretically expected values by about 40%. The theoretically expected values were calculated by utilizing a CH_4 yield coefficient of 0.38 m^3 for each kg of COD removed (Tchobanoglous *et al.*, 2003), and the percent composition of CH_4 in the head space gas (as discussed in the following subsection entitled, **Gas**

Composition (5.2.2.7)), which was about 85%. Through subsequent investigation and calculations (**Appendix A: Calculations**), it was determined that up to about 1/2 of the biogas that was unaccounted for may have been dissolved in the effluent of the system. The other missing 1/2 may be accounted for by the method of calculation for the expected gas production, as not all of COD removed may have been converted into CH_4. In this case, total COD was used for this calculation without consideration for the COD fractions which were not converted into CH_4.

The sudden temperature decrease on day 152 was followed immediately by a decline in gas production of about 20%. This decline corresponded to a decrease in COD removal at the same time (Figure 5.10), despite a sustained OLR of about 1.25 kg COD/m^3·d. This was another indication that the biomass within the system (methanogens and acetogenic bacteria) may not have coped well with the sudden shift in operating temperature.

After the removal of acetate supplementation, gas production dropped by about 72% from the values observed immediately after the temperature change, while the effluent COD of the system continued to rise. Because the reactor tank had to be opened on numerous occasions during this time period for membrane cleaning, it was not possible to get an accurate estimate of the gas production.

5.2.2.7 Gas Composition

During operation with acetate supplementation, gas quality, as seen in Figure 5.12, quickly reached steady state within a few days following major system perturbations. Upsets usually occurred during times of maintenance when N_2 gas was injected to replace liquid removed, or to replace air that may have entered the system during reactor service (i.e. membrane cleaning). This was evident on day 61, 122, 144, and 163. Despite modifications in the operating conditions for the submerged membrane, the general composition of the biogas remained unchanged. The presence of steady levels of N_2 at about 10-15% was initially thought to be an indication of a possible air leak into the system. However, subsequent

calculations suggested that the N_2 in the biogas may have resulted from the release of dissolved nitrogen from the incoming feed (see **Appendix A: Calculations**). Gas composition analysis was a quick method used to gauge the health of the biomass in the system; or more specifically, the level of methanogenic activity.

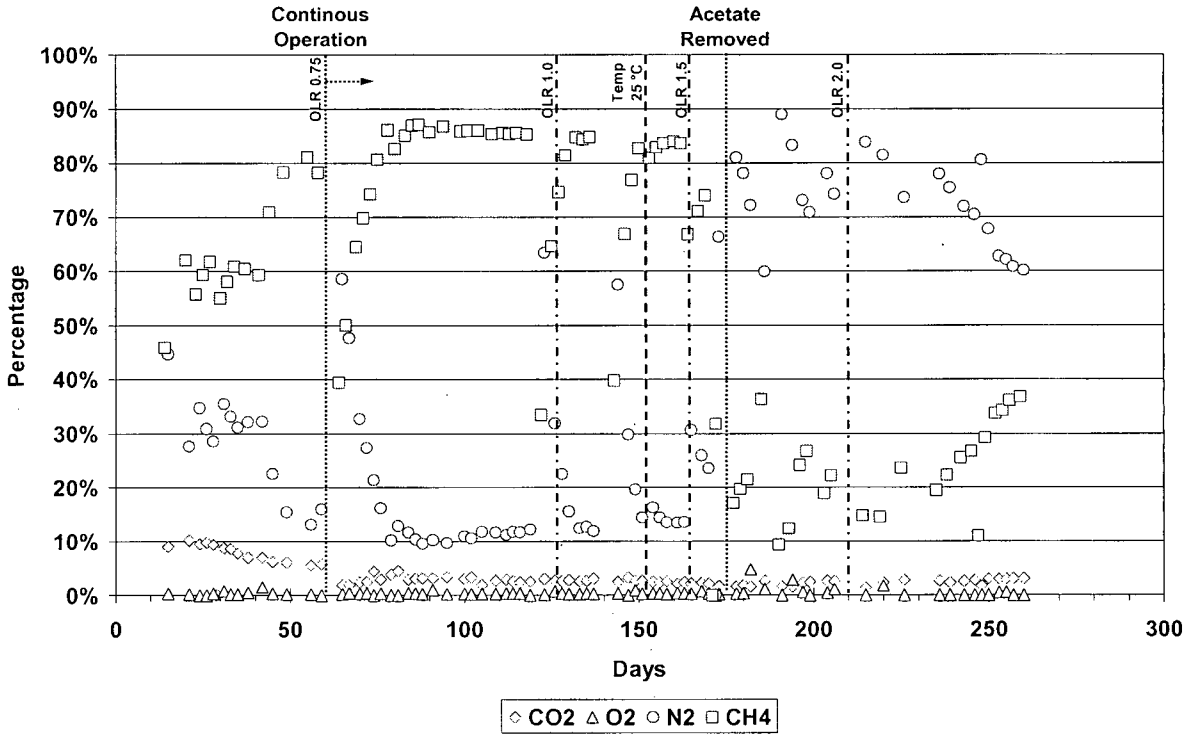

Figure 5.12 Submerged AnMBR gas composition.

Once the easily metabolizable supplemental acetate was removed, subsequently injected N_2 gas remained in the reactor head space much longer due to the low biogas production. Consequently, the 85% CH_4 levels could not be reclaimed due to the frequent membrane cleanings as outlined in the previous subsections.

The percent composition considering only CH_4 and CO_2 can be seen in Figure 5.13. During the period in between continuous operation and acetate removal (days 61 to 175), the average percentage of CH_4 was 96%. This value was higher than those typically reported in the literature. However, values reported in

the literature are usually based on the treatment of high-strength industrial/process wastewaters. The composition of the influent waste has a significant impact on the composition of biogas produced (E. Hall, Dept. of Civil Engineering, UBC, Vancouver, B.C., pers. comm., 2005). A decrease in the CH_4/CO_2 ratio after the removal of acetate was the result of a decrease in CH_4, rather than an increase in CO_2 production. During the last few operating days of the system, methane composition was on a slight increase (Figure 5.12), suggesting that the percent CH_4 could once again reach values close to the eighty percentile without the addition of acetate.

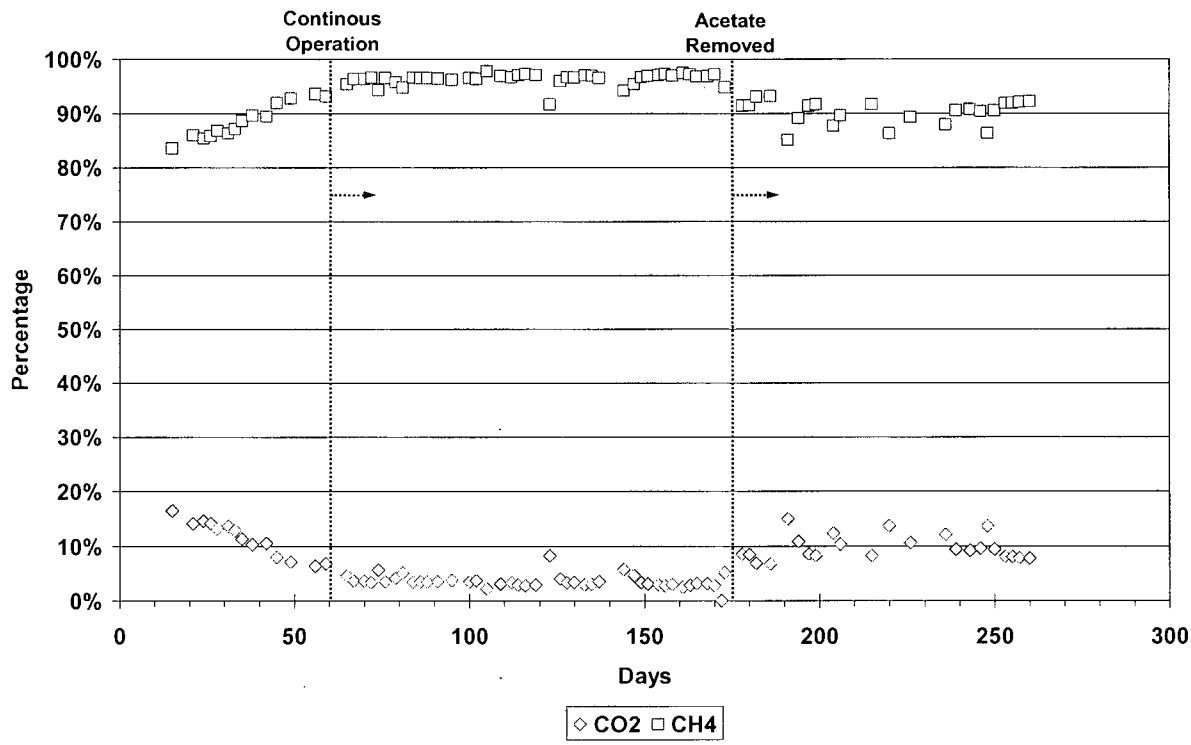

Figure 5.13 Submerged AnMBR percent CH_4 and CO_2.

5.2.2.8 Volatile Fatty Acids

Acetic acid is the major precursor of methane formation, and its concentration in the reactor was monitored and the results are presented in Figure 5.14. The average concentration of acetic acid in the raw influent supplemented with 300 mg/L of sodium acetate was about 187 (±28) mg/L, while the

concentration in the raw unsupplemented influent was about 11 (±6) mg/L. Average acetate removal in the system was near 100% from day 80 on, when the biogas production reached steady state (Figure 5.11). This removal efficiency was reduced slightly to 90% when the temperature of the system was decreased by 5 °C on day 152. The sudden change in removal efficiency was not unexpected, as methanogenesis may have been compromised due to reductions of aceticlastic methanogenic activities with a decrease in temperature. The increase in OLR to 1.5 kg COD/m^3·d through the decrease of reactor volume further reduced the removal of acetate to about 60%. The probable cause of this may have been the removal of about 25 % of the reactor mixed liquor volume, while flow and concentration of the reactor influent remained constant. The decreased biomass population may not have been able to utilize the available substrate immediately, which was thus discharged as acetate in the effluent. This was the source of increased effluent COD as available acetate was not converted to CH_4, and the resulting gas production decreased though this period.

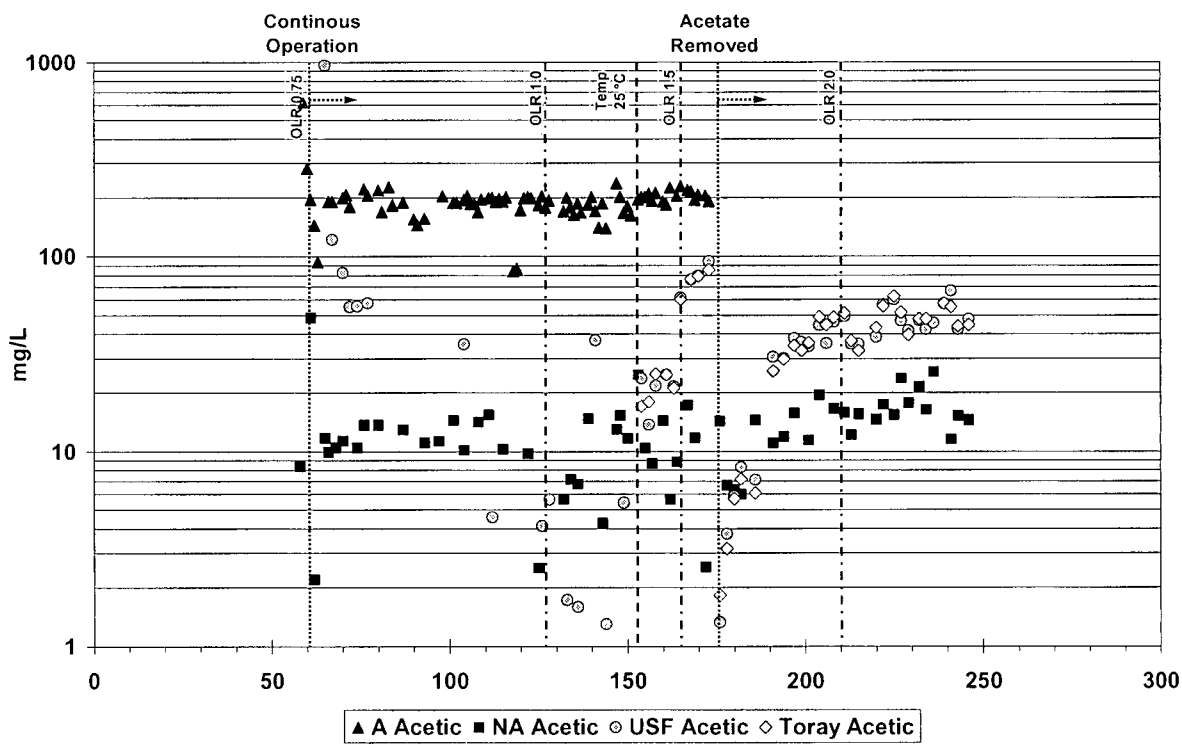

Figure 5.14 Submerged AnMBR acetic acid. A Acetic is influent with supplemental acetate while NA Acetic is influent without.

Immediately after the termination of sodium acetate addition, the effluent acetate concentration decreased to near zero as expected. However, because of the abrupt cessation of acetate supplementation, the methanogens may have been subjected to an upset in the system which could have induced endogenous respiration within the population. Over the next 15 days, as the process of hydrolysis and fermentation in the reactor continued, an increased concentration of VFA in the form of acetic acid could be seen. This was a strong indication that the syntrophic relationship between the methanogens and the acidogens in the system may have been out of balance. VFAs formed in the system were not utilized quickly enough by the methanogenic biomass. A reactor leak in the system on day 210, in which the resulting MLSS concentration in the reactor was drawn down to 4000 mg/L, did not increase the acetate concentration in the effluent significantly. Therefore, in the absence of a substantial methanogenic population, VFA concentrations in the effluent were similar to the concentrations before the biomass leak, leading to the conclusion that VFA consumption may have been negligible prior to the leak. Tchobanoglous *et al.*, (2003) reported that the anaerobic process is more stable when the VFA concentrations approach a minimal level, such as those seen prior to day 152, which could be taken as an indication that an adequate methanogenic population had been established to utilize the available hydrogen and VFA concentrations. A correlation with maximum unit gas production was also seen in Figure 5.11 during this same period.

Figure 5.15 presents the propionic acid concentration of the influent and effluent. Utilization of propionic acid via conversion to acetic acid was observed to mirror the trends of the acetic acid concentration shown in Figure 5.14. Near complete removal of propionic acid was observed prior to day 152. Subsequent to the increase in effluent acetic acid concentration after day 152, effluent propionic acid concentration between days 200 to 250 was about 300% higher than the concentration in the influent.

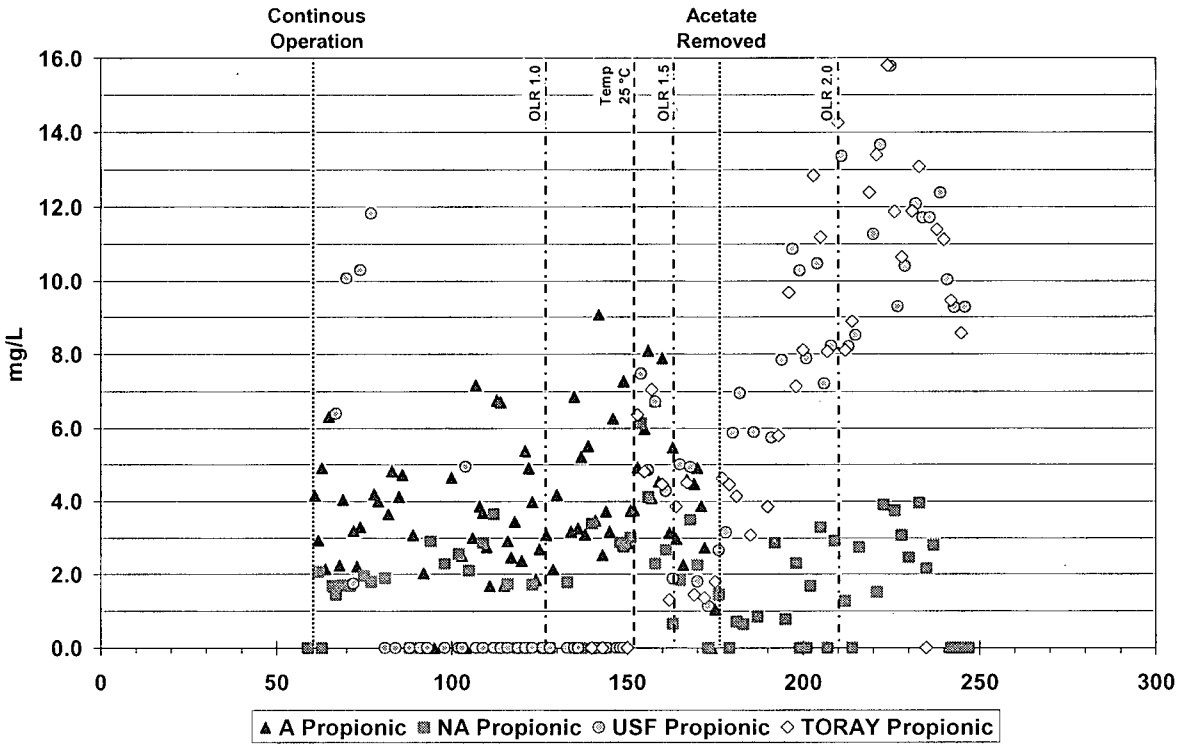

Figure 5.15 Submerged AnMBR propionic acid. A Propionic is influent with supplemental acetate while NA Propionic is influent without.

5.2.2.9 Suspended Solids Concentration

Due to the lack of sludge wasting, the reactor operated under very long SRTs (theoretical SRT of about 1700 days between days 100 and 125) which far exceeded the recommended value of 30 days (Tchobanoglous *et al.*, 2003) for complete conversion of degradable solids. Evident in Figure 5.16, MLSS equilibrium was reached at 8000 mg/L after day 130, which signified a net observable biomass yield of zero during that time (solids wasting from reactor through sampling was not significant). Calculations (**Appendix A: Calculations**) showed a 92% suspended solids hydrolization during the period immediately after acetate removal when solids loadings were high. Unfortunately, due to the rate-limiting step of methanogen growth, the majority of VFA and other by-products of fermentation were discharged through the effluent without being converted to CH_4.

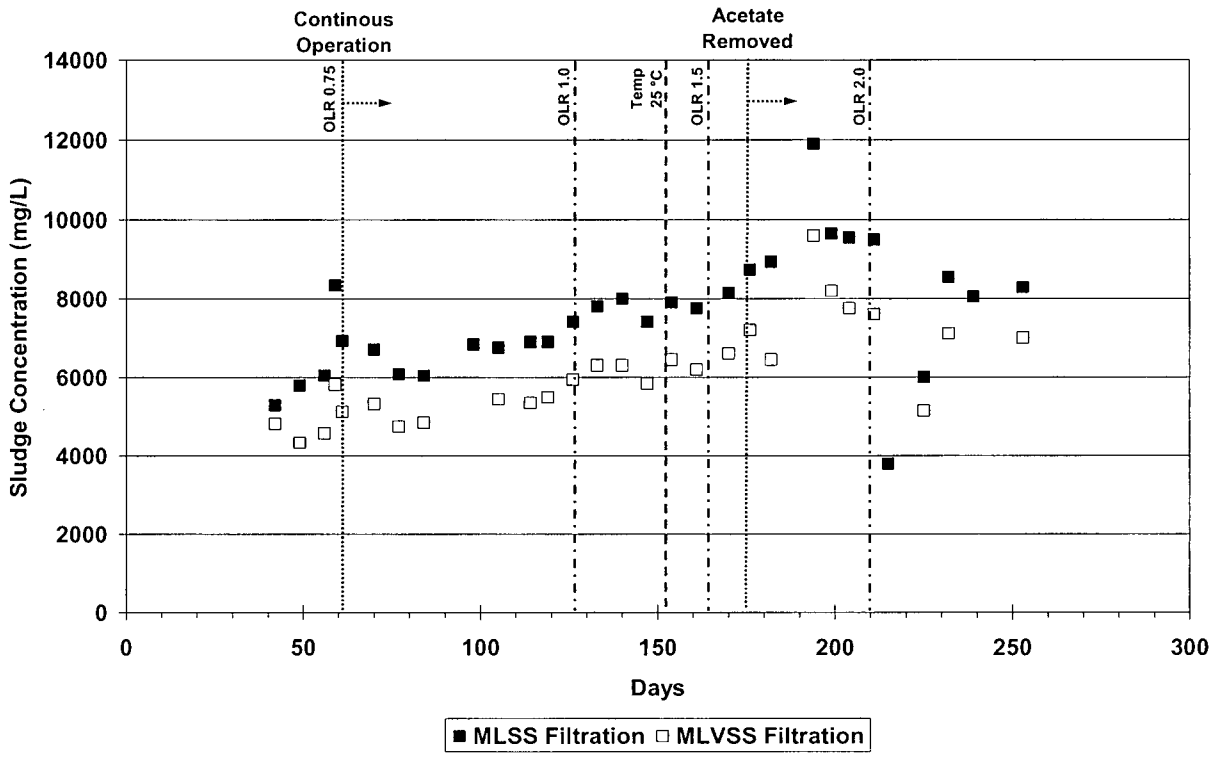

Figure 5.16 Submerged AnMBR SS/VSS concentration.

The fraction of VSS in the reactor MLSS was 80% throughout the duration of the study. Subsequent to the reactor mixed liquor leak on day 210, after which the resulting MLSS was about 4000 mg/L, the suspended solids concentration quickly climbed back up to 8500 mg/L in about 15 days. However, this increase was likely the accumulation of influent suspended solids in the system, and not of biomass growth.

A suspended solids concentration increase of 20% to 9550 mg/L resulted from an influent flow increase of about 50% following the removal of supplemental acetate. A mass flux examination of the increased suspended solids in the mixed liquor on the membranes was not conducted as it was beyond the scope of this study. However, as a result of the increased influent flow, the increase in mass flux transfer towards the membrane was a possibility. In addition, the fraction of colloidal solids and soluble products may

have played a role in the reduction of flux after the removal of acetate for the US Filter membrane. However, despite these possibilities, the Toray membrane flux remained unchanged.

The relatively constant Toray fluxes observed in Figure 5.5 suggest that the increase in the suspended solids concentration at low TMPs may not have played a significant role in flux decline. This was in agreement with Beaubien *et al.* (1996), who reported no additional negative impact on flux over an operating MLSS range of 2.5 to 25 g/L. This finding may indicate that the mechanism of membrane flux decrease for the US Filter hollow fibre membrane may be from causes other than suspended solids.

5.2.2.10 Particle Size Distribution

As defined previously in the description of the method for suspended solids determination, dissolved solids were defined in this report as anything that passed through the 1.5 μm glass fibre filter used for TSS determination. The definition of colloidal solids as considered by Tchobanoglous *et al.* (2003) includes particles within the range of 0.01 to 1.0 μm. Due to the vagueness in the published literature with regards to the definition of these terms, the values selected above were used for this report. Figure 5.17 presents the reactor mixed liquor particle size distributions from day 200 to 246. As described in detail below, particle size distributions prior to day 200 were not included as improper particle properties were used when the analyzer was initially set up. This created erroneous profiles that were not representative of the reactor mixed liquor. Figure 5.17 presents representative particle distribution profiles of the mixed liquor, but care must be taken with its interpretation as it only represent the volume and not the count of any given particle size.

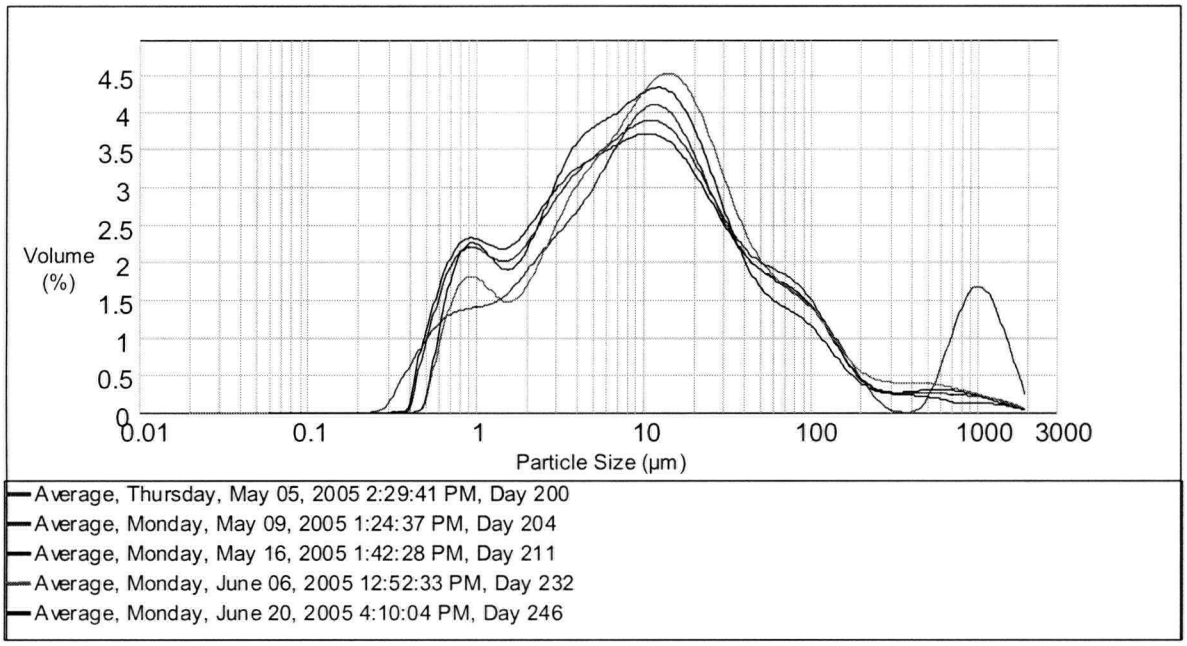

Figure 5.17 Submerged AnMBR particle size distribution (Day 200 to 246).

The technology used in the particle analyzer was based on the measurement of light scattering properties of the particles in question. For this reason, specific information regarding the type of particle analyzed must be accurately entered into the analyzer to obtain a representative result. However, due to the limited selections available from the built-in instrumentation library regarding particle refractive index, estimated values were first used when the first test was conducted on day 126. The use of the estimated properties continued until day 193, when a better fit for the particle refractive index was introduced to the system. Various parameters for the estimated properties were adjusted within the particle analyzer software to produce a model which reasonably represented the biomass analyzed. The particle size distributions for the mixed liquor after day 193 are presented in Figure 5.17. In terms of volume, the dominant particle size in anaerobic mixed liquor was approximately 10 µm as shown in Figure 5.17. This overall profile was very similar to the one reported by Park *et al.* (1999), who used a microfiltration membrane with a pore size of 0.1 µm in an anaerobic system.

Dissolved solids in the system (<1.5 μm) represented a significant portion of the MLSS as seen in Figure 5.17. As part of the dissolved solids, the colloidal solids in the system exist only between 0.4 μm to 1 μm, which is theoretically incapable of causing direct membrane pore plugging. However, this material may accumulate as a dense and compact foulant layer on the membrane surface causing considerable flux decline. If higher concentrations of colloidal solids were present, the observed membrane fouling may have been more severe. Based on percent volume, colloidal solids smaller than 0.3 μm were not present in great concentrations in the reactor mixed liqour.

5.2.2.11 Nutrients (Ammonia, Nitrate, Phosphate)

Figures 5.18, 5.19, and 5.20 show the measured nutrient concentrations of both the influent and effluent of the anaerobic submerged membrane process. Due to the low nutrient requirement generally associated with the anaerobic biomass, low to negligible removal efficiencies of nitrogen and phosphorus compounds were expected and observed. Despite the insignificant removal of these constituents from the wastewater, understanding their presence and concentrations in the reactor is important as both ammonium and phosphate are contributing compounds in the formation of struvite.

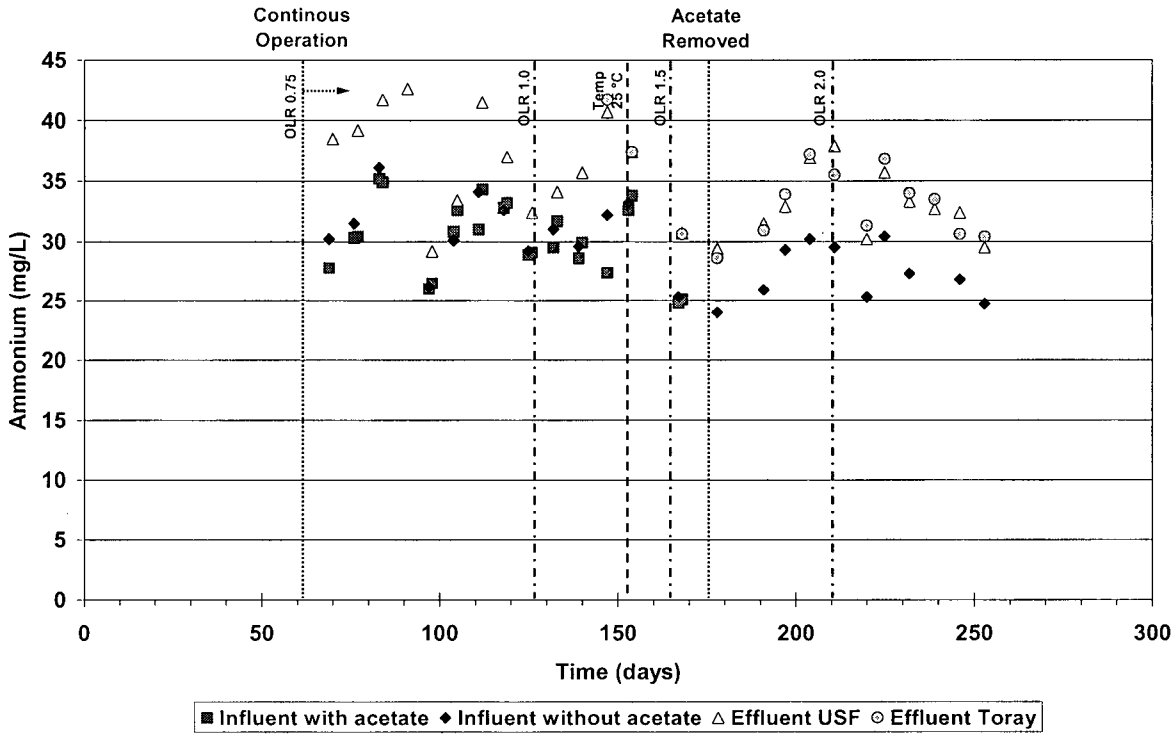

Figure 5.18 Submerged AnMBR NH_4^+-N concentrations.

Figure 5.18 is reflective of the ammonium ion concentration in the system due to sub-pH 7 operating conditions. The average observed ammonium concentration in the influent was 30 (±3) mg/L while the membrane effluent concentration was 35 (±4) mg/L. The consistently higher ammonium concentrations of about 5 mg/L in the effluent were indicative of the decomposition of organic matter in the reactor, which may have been enhanced by the long SRT. The effluent ammonium concentration remained similar to the influent concentration with only the small aforementioned difference, as no nitrification occurred within the system. Inhibition of methanogenic activity due to free ammonia was not a concern as the pH level remained slightly below 7 for the duration of the study. Additionally, the observed ammonium concentrations were not close enough to levels suggested in the literature (Tchobanoglous *et al.*, 2003) at which an inhibition threshold would be reached even if the system pH was increased significantly.

As expected, results from Figure 5.19 suggest that no nitrification took place in the anaerobic membrane bioreactor. The measured level of nitrates in the reactor effluent never exceeded 1 mg/L through the duration of the experiment.

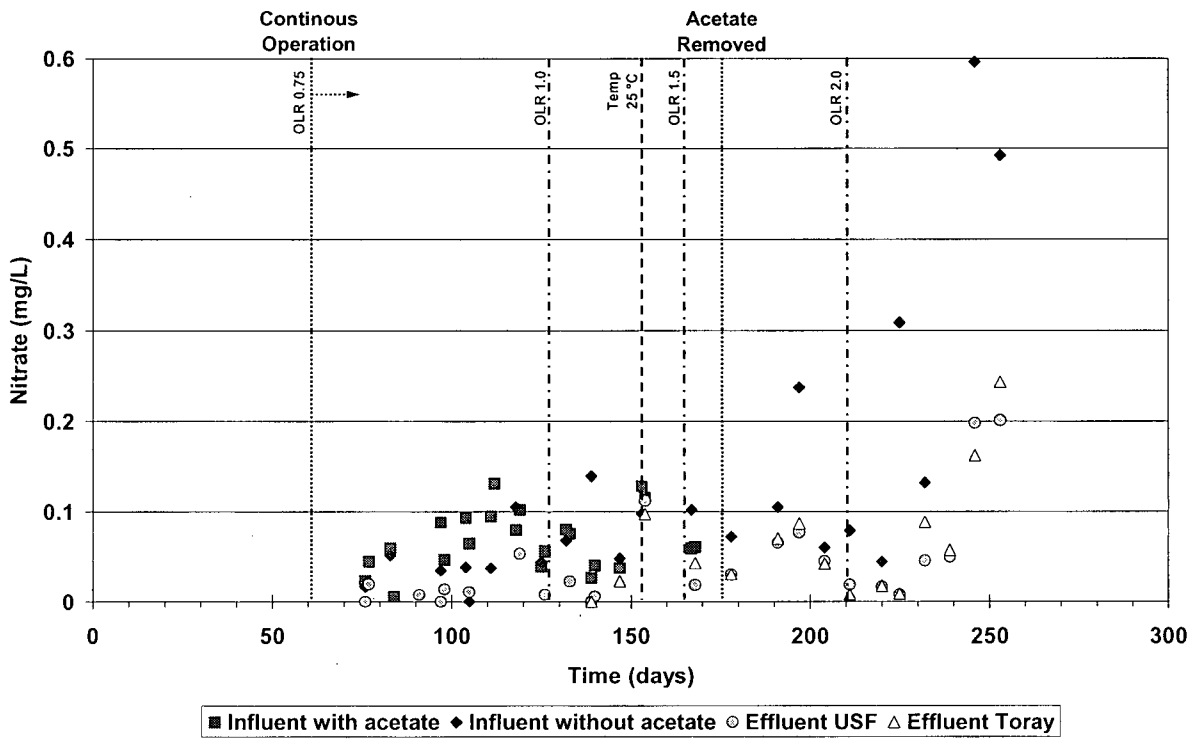

Figure 5.19 Submerged AnMBR NO_x-N concentration.

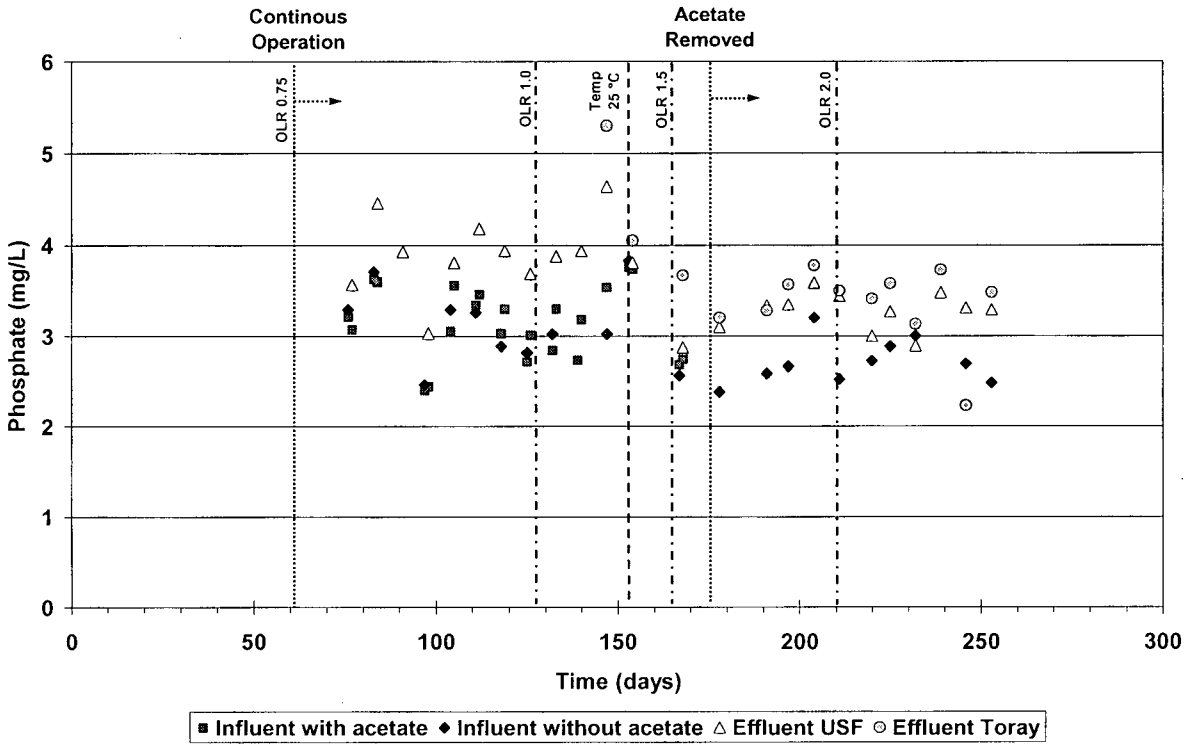

Figure 5.20 Submerged AnMBR phosphate concentration.

Orthophosphate concentrations shown in Figure 5.20 remained relatively constant for both the influent and effluent of the system. The average influent and membrane effluent phosphate concentrations were 3.0 (±0.4) mg/L and 3.6 (±0.6) mg/L, respectively. Since the determination of orthophosphate requires filtration through a 0.45 μm membrane filter, whereas the permeate is filtered through the 0.08 μm membrane units, the influent orthophosphate concentration should be higher than the effluent. This assumption is due to the passage of any orthophosphates which may be in the range of 0.08 to 0.45 μm. However, the effluent orthophosphate concentration was consistently higher than that of the influent by almost 1 mg/L. A likely source of increase for orthophosphates is through the hydrolization of organic phosphorus compounds and polyphosphates. Like the ammonium concentrations discussed previously, a small but noticeable decrease in phosphate concentration was observed after the reactor temperature decrease on day 152. This may be reflective of the decreased hydrolization of organic phosphorus

compounds or polyphosphates in the reactor. Additionally, similar to ammonium, orthophosphate concentrations in the reactor mixed liquor may aid the formation of struvite. Unfortunately, the investigation of the presence of struvite formation on membranes was not in the scope of this study.

In all cases, negligible nutrient removal was detected for the measured nutrients as expected from an anaerobic system.

5.3 Membrane Fouling

Numerous researches have been conducted on the fouling characteristics of membranes operating in aerobic MLSS, leading to relatively well understood fouling characteristics under those conditions. Thus the focus of this section is to ascertain a better understanding of the different characteristics between anaerobic and aerobic MLSS on the fouling of submerged membranes. This could also allow a better understanding of their similarities so as to take advantage of results from existing aerobic studies. Membranes used during this part of the study were new PVDF Zenon hollow fibre membranes, with a nominal pore size of 0.04 μm. These membranes are similar to the US Filter membranes in terms of material, charge, and hydrophobicity. The filtration setup in Figure 4.3 was adopted from Geng (Geng, 2006). The apparatus used was a 2 L reactor (Figure 4.2) that supported four strands of hollow fibre membranes, for a total membrane surface area of 0.004 m^2. The reactor was coupled with a nitrogen gas sparging system for the anaerobic mixed liquor (air for the aerobic mixed liquor) to provide some turbulence at the membrane surface. A digital pressure gauge connected to the permeate line was used to monitor changes in vacuum pressure over time. The system was driven by a Masterflex positive displacement pump, coupled with variable flow speed control.

Experimental operating conditions identical to Geng's were used through every test to ensure consistency. The Masterflex pump flow rate was set to 2.2 mL/min, which was equivalent to a membrane permeate flux of 33 Lmh. No additional influent was added and the increase in retentate

concentration change at the end of the test was 20%. The air/nitrogen gas sparging intensity was set at 1.8 L/min for all tests. This test was conducted three times, each with different mixed liquor concentrations, as shown in Table 5.2. Two different anaerobic reactor MLSS concentrations were used to provide information on fouling characteristics of the submerged membranes under different anaerobic mixed liquor concentrations.

Table 5.2 Filtration mixed liquor concentration

Run	Mixed liquor type and concentration
1	Aerobic mixed liquor with 6 g/L MLSS concentration.
2	Anaerobic submerged membrane MBR mixed liquor with concentration of 7 g/L MLSS.
3	Diluted anaerobic submerged membrane MBR mixed liquor with concentration of 5 g/L MLSS.

The dependence of permeate flux on applied trans-membrane pressure can be expressed by the following equation (Choo and Lee, 1996a):

$$Jv = \frac{TMP}{\mu(Rm + Rc + Rp + Ra)} \qquad \text{Equation 5.1}$$

where Jv is the permeate flux, TMP is the applied trans-membrane pressure, and is μ the dynamic viscosity of the permeate. Rm is the intrinsic membrane resistance, Rc is the resistance due to a cake layer (or more specifically, the concentration polarization layer) formed on the membrane surface, Rp is the resistance due to plugging in membrane pores, and Ra is the resistance due to adsorption of organic species within the membrane pores.

The total resistance encountered during filtration of mixed liquor is due to Rm plus the total resistance invoked by the activated sludge (Rs); which is composed of the three parameters as described above: Rc, Rp, and Ra. Rm can be obtained by subjecting new membrane modules to filtration of clean water (i.e.

distilled de-ionized water (DDW)). Rs can be obtained from the observed difference between the filtration resistance of the DDW and the mixed liquor. In a filtration process, Rc can be eliminated through the removal of bulk sludge, membrane relaxation, and gas sparging. Rp can sometimes be reduced by the process of back-flushing. Ra is hydraulically irreversible and can only be removed through chemical cleaning. The graphical relationship between these resistances is illustrated in Figure 5.21 below. Note the y-axis of Figure 5.21, which has a selected resistance of 0.20 to 0.28 kPa/Lmh (resistance) to graphically exaggerate the different types of resistances. Explanation of the experimental procedures and phases in the figure are detailed in the following section.

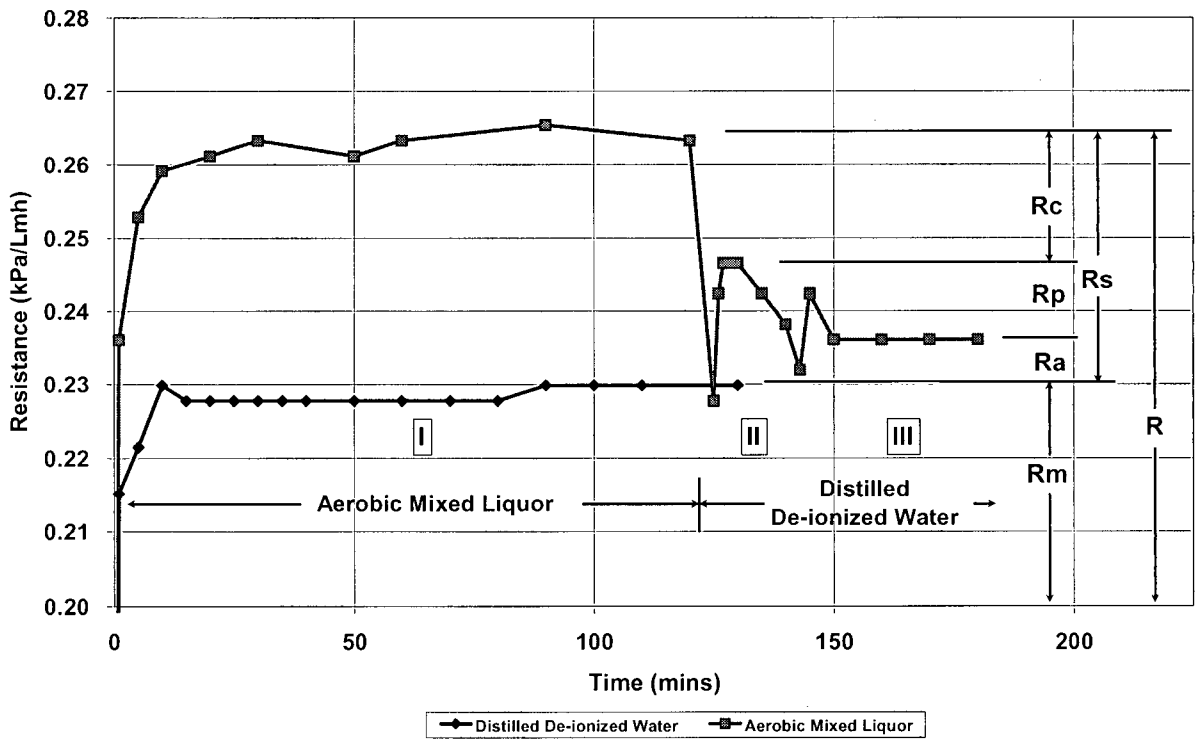

Figure 5.21 Aerobic mixed liquor 6 g/L MLSS filtration resistance.

5.3.1 Aerobic Mixed Liquor Filtration Tests

The aerobic mixed liquor used in the first test was obtained from the UBC pilot plant membrane bioreactor for the treatment of municipal wastewater with enhanced biological phosphorus removal. The

aerobic system used the same wastewater feed source as the anaerobic treatment system in this study. The result from the aerobic system mixed liquor test, presented in Figure 5.21, was then used as a benchmark to which the anaerobic mixed liquor filtration results were compared. New Zenon membrane fibres were first submerged in DDW and operated under vacuum with air sparging to estimate the intrinsic membrane filtration resistance, Rm. The DDW was then replaced with the aerobic mixed liquor for the membrane filtration tests. The results were subdivided into three phases (Figure 5.21) to help better describe the observations. Phase I produced an increase in resistance during the first 30 minutes which then levelled off and remained steady for the rest of the phase. At the end of Phase I (120min), mixed liquor in the reactor was replaced with DDW (for the remainder of the experiment) and followed immediately by a 5 min membrane relaxation period prior to re-starting permeation for Phase II. During Phase II, a significantly lower resistance was observed for the aerobic mixed liquor than in Phase I. With the conclusion of Phase II at 140 min, a quick 30 second back flush was done preceding Phase III. The observed resistance in Phase III reached steady state at an even lower level. The test was concluded at the end of 180 minutes with Phase III.

As shown in Figure 5.21, the observed Rs of the aerobic mixed liquor represented only 15% of Rm. The result of this test indicates that a surface cake layer produced the greatest contribution to filtration resistance (Rc) for the aerobic mixed liquor. However, despite Rc representing about 50% of the total Rs, its resistance value was only 0.02 kPa/Lmh of the total resistance. Finally, Ra and Rp represented only small portions of Rs, at 20% and 30%, respectively.

5.3.2 Anaerobic Mixed liquor Filtration Tests

This section presents the filtration characteristics of mixed liquor from the AnMBR from this study. The first of two filtration tests was conducted with the original mixed liquor taken from the AnMBR (7 g/L). The second was done with a diluted version of the same mixed liquor (5 g/L) in order to determine

membrane fouling characteristics under different anaerobic mixed liquor concentrations. The resulting anaerobic resistance graphs are shown in Figure 5.22 and 5.23.

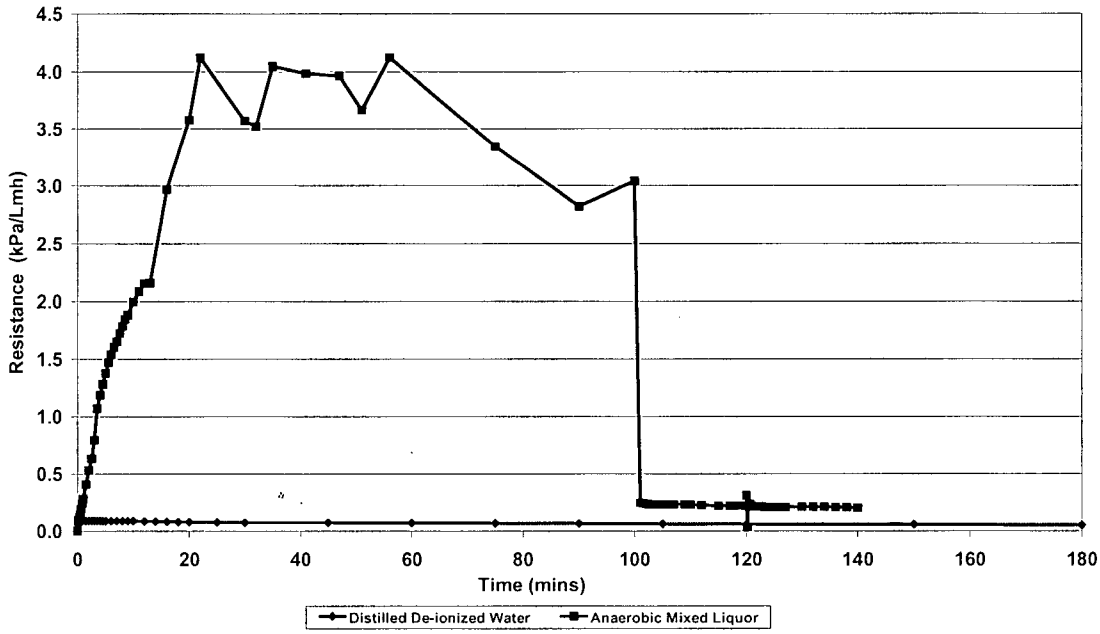

Figure 5.22 Anaerobic mixed liquor (7 g/L) filtration resistance.

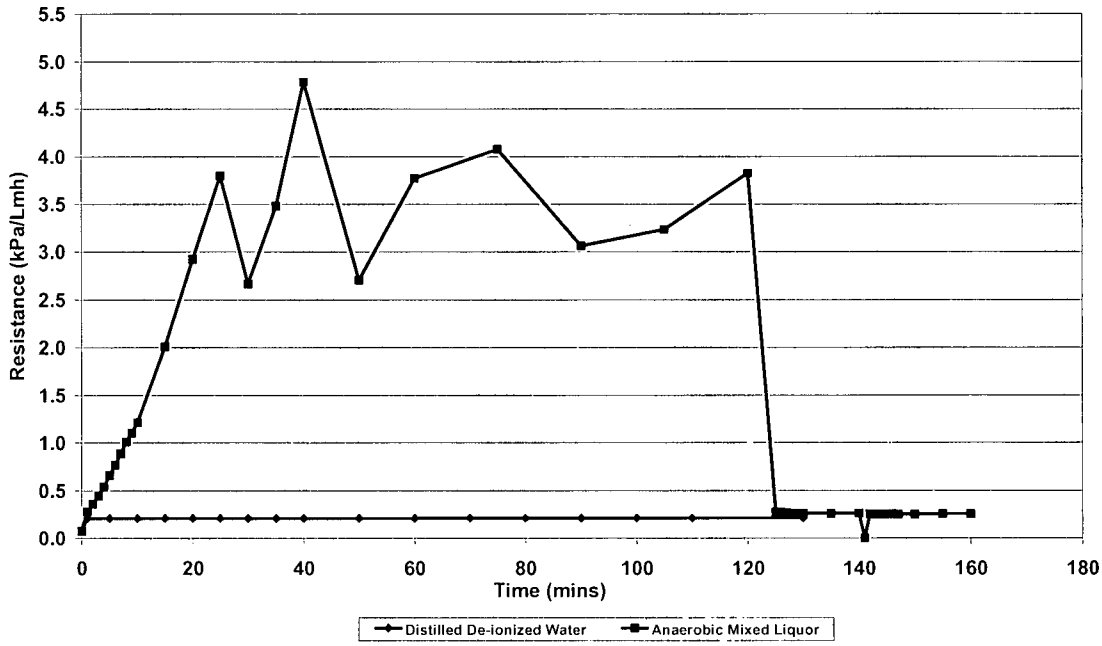

Figure 5.23 Anaerobic mixed liquor (5 g/L) filtration resistance.

Although the operating conditions were identical for both the aerobic and anaerobic filtration tests, it was established that the anaerobic mixed liquor developed a total filtration resistance (Rs) that was about 100 times greater than that of the aerobic mixed liquor at similar MLSS concentrations. Unlike the aerobic mixed liquor, the resistance of the anaerobic mixed liquor in terms of Rc was significantly greater than the intrinsic membrane resistance itself (about 15 times greater). Furthermore, anaerobic Rc was the main contributing factor to Rs. The resistance due to adsorption and pore plugging only accounted for a very small portion and was thus considered negligible for the submerged anaerobic mixed liquor, at least in the short term. The Masterflex pump used during these tests was sized primarily for aerobic filtration, and may not have been suitable for use with anaerobic mixed liquor which required higher TMPs. This may have resulted in the unevenness of the filtration resistance graphs.

Although the total resistances of the two different anaerobic mixed liquors did not significantly differ, there were small differences in the resistance due to adsorption. The difference may be due to the higher colloidal solids concentration available in the high concentration mixed liquor, which through mass flux, would migrate and accumulate on the membrane surface to a greater extent. The higher dissolved solids concentration in the form of SMPs, associated usually with higher biomass concentrations, may have also played a significant role. Since the resistance due to adsorption is considered irreversible by physical foulant-reducing measures such as gas sparging and backflushing, chemical cleaning is usually necessary. In cases where longer mixed liquor filtration is required, such as those seen in the AnMBR, resistance due to adsorption could undoubtedly play a more decisive role in membrane fouling.

It must be noted that all results expressed in this section reflect only very short-term membrane fouling phenomena. Further studies would be required to determine the long term effects of membrane fouling caused by anaerobic mixed liquors of different concentrations. Furthermore, the hydrodynamic conditions in the test apparatus were not the same as those which are typical of full scale systems. The

results would therefore require careful interpretation. Nonetheless, anaerobic mixed liquor fouled the membranes to a greater extent than the aerobic mixed liquor.

6.0 CONCLUSIONS AND RECOMMENDATIONS

6.1 Summary

The purpose of this research was to explore the feasibility of treating municipal wastewater with AnMBRs. Both the effectiveness of the anaerobic treatment and the membrane performance aspects of the AnMBR were investigated. Two different experiments were conducted in this study to address, (1) the overall effectiveness of the AnMBR system, as wells as to compare two different commercially available membrane modules, and (2) the type of anaerobic submerged organic membrane fouling present compared to those identified for the same membranes applied in an aerobic MBR System. The first was the treatment performance of the anaerobic membrane bioreactor system while operating under different values of OLR, HRT, and temperature. The second objective was to ascertain the specific membrane fouling mechanism using a reactor designed to explore this issue in detail.

The following summarizes the key observations from this study:

6.1.1 Municipal Wastewater Treatment with AnMBR

1. The highest operational OLR achieved for the submerged AnMBR of this study that could realistically be sustained was about 1.0 kg COD/m^3·d for the treatment of raw fine-screened municipal wastewater without supplemental acetate. This was due largely to the relatively low COD concentration of the influent and the inability for the experimental set-up (i.e., submerged membranes) to sustain an HRT of less than 5 hours.

2. Initial COD removal efficiency of the system was about 84% when the strength of the influent wastewater was augmented with supplemental acetate. However, a significant portion of effluent COD may have been accounted for by dissolved CH_4 in the effluent. If this gas was completely purged during COD analysis, the resulting removal efficiency may have increased to about 95%. The COD removal efficiency of the system without supplemental acetate decreased

to about 68%. This value eventually decreased to about 43%, when the operating temperature was decreased from 30 to 25°C and the OLR was increased from 1.5 to 2.0 kg COD/m^3·d. Further, no significant nutrient removal was observed across the anaerobic process, as expected.

3. Gas production of the system, while the influent was supplemented with acetate, decreased by about 20% following the decrease of temperature by 5 °C. Concurrently, acetate removal within the reactor was decreased to 90% from 100% after the 5 °C temperature drop. The eventual gas production from the treatment of raw wastewater without acetate was about 40 L/m^3·d at 25 °C. Since the amount of CH_4 produced was relatively low due to the low strength of the wastewater, it was hypothesized that a substantial amount (dependent on the saturation concentration and percent saturation of CH_4 in the mixed liquor) of CH_4 produced remained soluble and exited the system with the treated permeate.

4. The effect of the temperature decrease on the membrane flux was immediately evident on the hollow fibre membrane. The rate of flux decline increased by a factor of 2. Insufficient data were available for the flat sheet membrane to produce a similar conclusion.

5. The membrane relaxation cycle was increased to 4 min on, 1 min off, without appreciable impacts on the measured flux of both membrane types. A flux increase of about 40% was observed for the flat sheet membrane when the gas sparge rate was increased by about 36%. However, the flux declined immediately in the days following.

6. Initially, the observed flux decrease of a new flat sheet membrane was nearly 40% greater than that of a new hollow fibre membrane. The hollow fibre membrane was observed to be more resistant to flux loss when the unused membrane was first installed. However, over time, the

permeability of the hollow fibre membrane decreased to less than half of the permeability for the flat sheet membrane.

7. The flux of the flat sheet membrane eventually reached a steady state value that was about 55% of its original flux in just under 50 days after its initiation. The flux of the hollow fibre membrane continued to decline throughout the duration of the study.

8. Flux decline on days immediately following a membrane cleaning event was very severe and exhibited a "memory" like behaviour. This phenomenon was more prevalent for the hollow fibre membrane. The most effective hollow fibre flux recovery method used for this study was cleaning method IV. This required an initial rinse with sodium hypochlorite followed by alternating caustic and acidic solution cleanings. The most effective flat sheet membrane cleaning procedure was through mechanical surface foulant removal via the use of a water jet. This suggests that the method of fouling for the two membranes may have been different in nature.

9. Due to the extremely long SRT made possible by absolute suspended solids retention of the membranes, the calculated solids hydrolization **(Appendix A: Calculations)** was about 90% assuming a net anaerobic biomass growth yield of zero.

6.1.2 Mechanisms of Short Term Membrane Fouling

1. Short term membrane fouling caused by anaerobic mixed liquor was mainly attributed to external fouling, of which cake layer fouling played the most prominent role. Gas sparging and membrane relaxation can effectively reduce external fouling. Back flushing did not contribute significantly to flux recovery because short term pore plugging was not a major issue with the anaerobic mixed liquor.

2. At higher suspended solids concentrations, an increase in resistance due to adsorption was observed. This may have been due to the higher levels of SMPs associated usually with higher mixed liquor concentrations.

3. The observed Rs of the anaerobic mixed liquor was about 100 times greater than the aerobic Rs, when filtering mixed liquors of similar concentrations.

6.2 Conclusions

To address the objective of the research in Section 3, the following conclusions were drawn.

1. As a result of low influent CODs, operating above an OLR of 2.0 kg COD/m^3·d treating municipal wastewater may be unrealistic in terms of sustained operation. Gas sparging is an effective method for reducing the rate of membrane fouling, but extensive fouling maybe inevitable. An on/off cycle of 4 minutes on and 1 minute off is a reasonable compromise between flux and membrane relaxation. The use of longer on cycles may lead to much quicker and irreversible fouling. Decreasing operating temperature will require the acclimatization of biomass over an extended period due to the possibility of decreased biomass activities, as observed in this study. From an operational standpoint, AnMBRs are susceptible to changes in operating temperatures, which could lead to issues with effluent compliance. However, operating at temperatures of about 20 °C appears feasible only if the system HRT (and ultimately the membrane performance) could be decreased and sustained for a suitably high OLR. Finally, due to the lack of nutrient removal in the AnMBR, additional polishing steps maybe required to mitigate effluent nutrient discharge concentrations, should there be a concern.

2. Initially, the hollow fibre submerged membrane out-performed the flat sheet membrane quite considerably in terms of achievable flux and the rate of fouling. However, irreversible fouling

became a much greater issue for the hollow fibre membrane over time. Eventually, the permeability of the hollow fibre membrane was less than half of the permeability for the flat sheet membrane. Furthermore, it appears that the mechanism for membrane fouling between the two different types of membranes maybe different, despite the same material composition. The hollow fibre membrane appeared to be much more susceptible to irreversible fouling while the majority of the recoverable flux for the flat sheet membrane could be obtained through mechanical cleaning. One major drawback of the flat sheet membrane is the packing density. Despite having similar volumes in the reactor, the filtration surface area of the hollow fibre membrane was about 2.5 times that of the flat sheet membrane. This has significant implications on the achievable size of the reactors.

3. Short-term filtration tests with anaerobic and aerobic mixed liquors under similar concentrations and filtration conditions yielded vastly different results. Cake layer fouling was significantly greater in the anaerobic mixed liquor than the aerobic mixed liquor. This is likely the result of significantly different physical and compositional characteristics between the two types of mixed liquor.

6.3 Recommendations

As a result of the findings from this study, the following recommendations will require most attention.

1. Due to the inability for AnMBR nutrient removal, a final aerobic polishing may be required. This could also be a useful step prior to membrane filtration in an anaerobic MBR to reduce the concentration of COD that would normally be degradable only under aerobic conditions.

2. There is the need to better understand the hydraulic operating conditions for the submerged membrane within the bioreactor. Until now, researchers have only really focused on specific

areas of the AnMBR during any given study. A more comprehensive approach to hydraulic and operating parameter optimization is required to prolong membrane flux.

3. An in-depth study into the microbiology of the system in terms of the type, the population and their symbiotic relationship is required to further AnMBR advancement. A great deal of progress gained in this field is without much detailed knowledge of the biological workings of the system at a microscopic level.

4. The change in surface properties of the membrane especially after chemical cleaning should be examined in detail. Some of the currently accepted cleaning techniques may have adverse effects on long-term flux recovery. As well, most of cleaning procedures in the literature apply only to membranes used in aerobic systems. Detailed long term fouling mechanisms of the anaerobic membrane will be required.

5. A comprehensive evaluation between an aerobic MBR versus an anaerobic MBR system operating in parallel. Factors of consideration will not only be the efficiency of treatment, but also the cost and the system energy balance. Are AnMBRs in fact worth the investment due simply to the possibility that they are potentially energy producing systems, or will the capital and operating cost never justify their use?

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APPENDIX A: CALCULATIONS

Solubility of methane in water at 30 °C and 25 °C (Tchobanoglous *et al.*, 2003):

For 30 °C

Percent CH_4 in headspace gas = ~85%

$$x_g = \frac{P_T}{H} P_g \qquad \text{Equation A.1}$$

where P_g = mole fraction of CH_4 in headspace, (mole CH_4/mole of headspace)

H = Henry's law constant, $\frac{atm(mole\ CH_4\ /\ mole\ headspace)}{(mole\ CH_4\ /\ mole\ water)}$ = 37600 at 20 °C

P_T = total pressure (atm)

x_g = mole fraction of CH_4 in water, mole CH_4/mole water

$$= \frac{mole\ CH_4(n_g)}{mole\ CH_4(n_g) + mole\ water(n_w)}$$

Determine x_g:

Convert H to 30 °C;

$$Log_{10}H = \frac{-675.74}{273.15 + 30°C} + 6.880$$

H_{30} = 44765

$$x_g = \frac{1.0}{44765} 0.85 = 1.89 \times 10^{-5}\ mole\ CH_4/mole\ water$$

One litre of water contains 1000 g/ 18 g/mole = 55.6 *mole* H_2O

$$\frac{n_g}{n_g + n_w} = 1.89 \times 10^{-5}$$

$$\frac{n_g}{n_g + 55.6} = 1.89 \times 10^{-5}$$

where,

$n_g + 55.6 \approx 55.6$

therefore,

$n_g = 1.06 \times 10^{-3}$

$$[CH_4] = \frac{(1.06\ x10^{-3}\ mole\ CH_4 / L)(16\ g\ CH_4 / mole)}{(1g / 10^3\ mg)} = \mathbf{16.9\ mg/L}, \text{ at } 30\ °C$$

<u>For 25 °C</u>

$[CH_4]$ at 25 °C is **18.4 *mg/L***

Percent gas production unaccounted for due to dissolved biogas in the effluent (Tchobanoglous *et al.*, 2003):

Take day 139, when effluent flow was 56 *L/day.*

So,

56 *L/day* × 16.9 *mg/L* CH_4 = 952 *mg/day*, or 0.952 *g/day* lost

$$V = \frac{nRT}{P} \qquad \text{Equation A.2}$$

$$V = \frac{(0.952\ g / 16\ g / mole)(0.082057)(273.15\ K + 20\ °C)}{1.0\ atm} = 1.43\ L\ CH_4$$

Total biogas lost is therefore;

1.43 *L* CH_4 / 0.85 (percent CH4 in biogas) = **1.7 *L/day*** of biogas, or about 20% of total gas production. In terms of missing headspace biogas (relative to the calculated theoretical production), this value roughly represents about ½ of the biogas unaccounted for.

Percent N_2 in reactor headspace gas attributed to influent feed:

Take day 111, when effluent flow was 38 *L/day*, and assuming N_2 saturation of 50%.

N_2 at 50% saturation at 10°C = 9.20 mg/L in water where atmosphere has ~78% N_2.

N_2 at 50% saturation at 29°C = 0.78 mg/L in water where reactor headspace has ~10% N_2.

So,

9.20 *mg/L* – 0.78 *mg/L* = 8.42 *mg/L* N_2

8.42 *mg/L* × 38 *L/day* = 320 *mg/day*

$$n = \frac{0.320\,g}{24\,g/mol} = 0.013\,mol$$

$$V = \frac{nRT}{P} = \frac{(0.013)(0.082057)(273.15\,K + 30\,°C)}{1.0\,atm} = 0.323\,L\,N_2$$, or about 8% in reactor head gas.

Solids Accumulation:

Influent TSS ≈ 80 *mg/L*, and flow form day 175 to 204 was about 83 *L/day*.

Reactor MLSS on day 175 was about 8742 *mg/L*.

Solids accumulation over the 29 days was thus:

80 *mg/L* × 83 *L/day* × 29 *days* = 192,560 *mg*

With a reactor volume of 20 *L*, the total reactor MLSS on day 204(assuming no solids destruction or biomass growth or decay) should be:

$$\frac{192{,}560\,mg}{20\,L} + 8742\,mg/L = 18{,}370\,mg/L$$

However, the MLSS on day 204 was only 9,550 *mg/L*.

This reflects a net solids reduction of about **92%**.

Note: The net biomass yield may have contributed significantly to the above value but its effect on the net change in reactor MLSS was not studied in detail.

APPENDIX B: RAW AND CALCULATED DATA

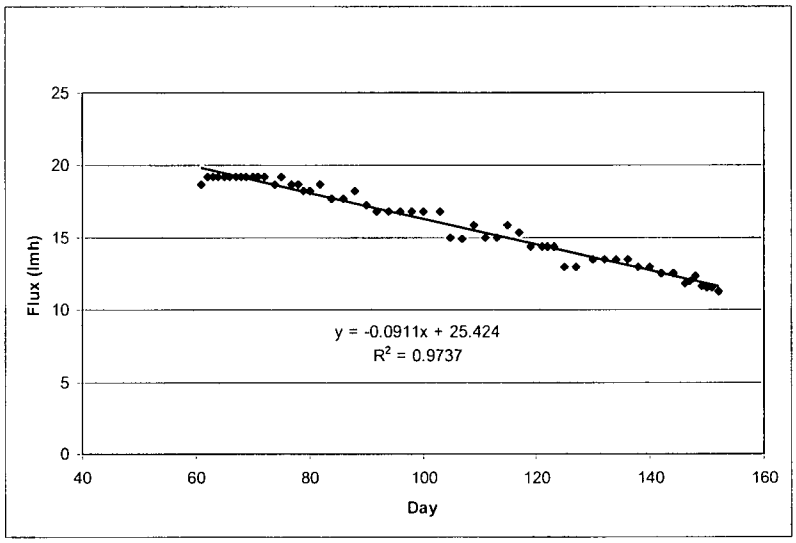

Figure B.1 USF flux at 30 degrees C.

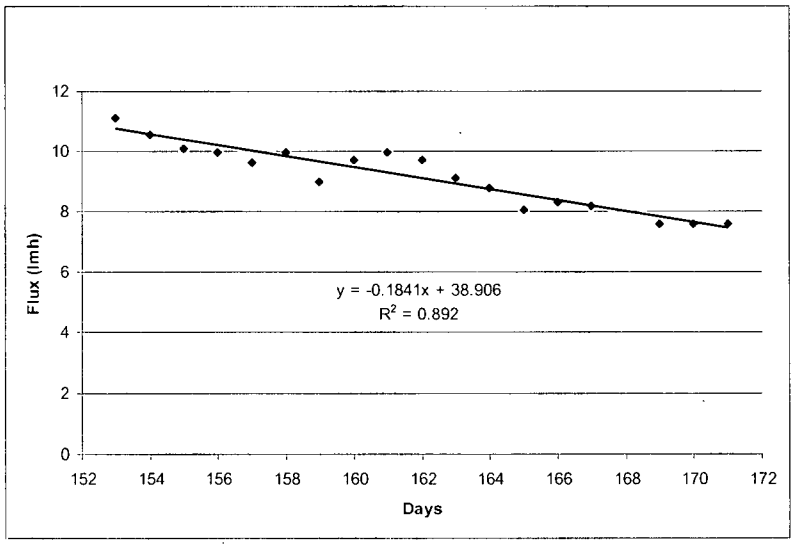

Figure B.2 USF flux at 25 degrees C.

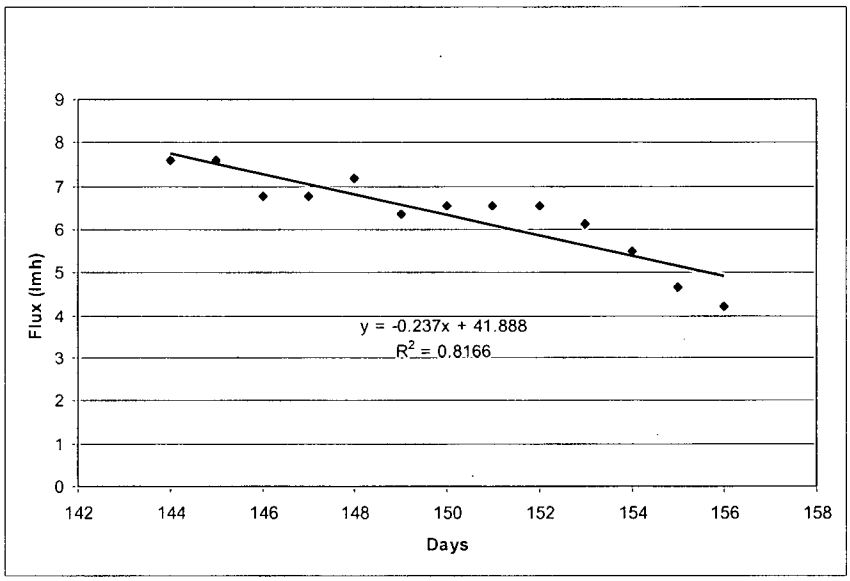

Figure B.3 Initial Toray flux.

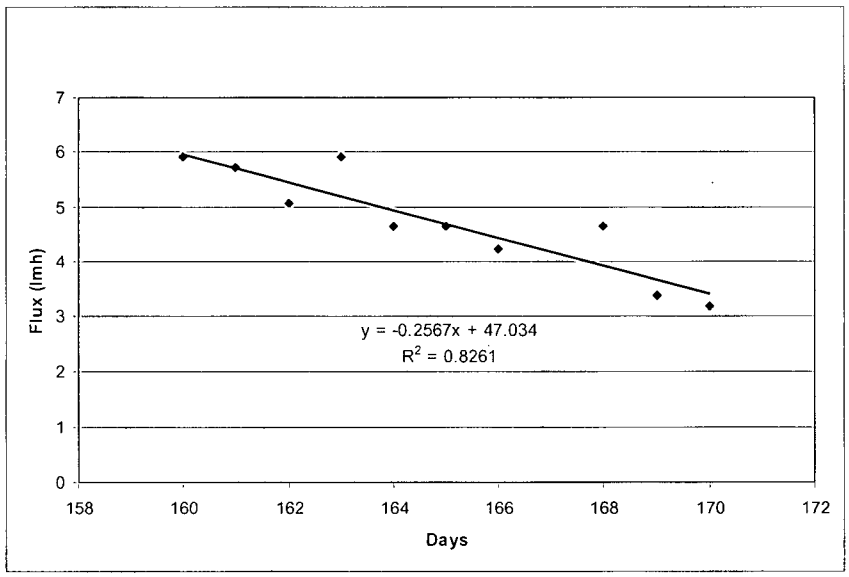

Figure B.4 Toray flux after gas sparge increase.

AnMBR Experimental Data

Date/Events	Day	Flow Rate (L/day)	Reactor Level (cm)	HRT(hrs)	OLR(kgCOD/ m3/d)	Time	Temp (Celsius)	pH Infl	pH Effl	Gas Count (tips)	Gas Prod. Rate (L/d)	Gas Prod. (L/m3/d)	Expected CH4 prod. (L/d))	Expected CH4 (L/m3/d)
19-Oct-04	1													
20-Oct-04	2		100			10:45 AM				63	4.41	144		
21-Oct-04	3		100			10:00 AM	28.0		7.7	66	4.77	156		
22-Oct-04	4		100			11:25 AM				64	4.23	138		
23-Oct-04	5		100			11:05 AM			7.5	44	3.12	102		
24-Oct-04	6		100			10:30 AM			7.5	27	1.94	63		
25-Oct-04	7		100			10:00 AM	27.0		7.7	25	1.79	58		
26-Oct-04	8		100			10:00 AM				22	1.54	50		
27-Oct-04	9		100			10:10 AM	26.0		7.7	0		0		
28-Oct-04	10		100			8:50 AM	29.1		7.8	18	1.33	44		
29-Oct-04	11		100			10:20 AM	26.0		7.8	13	0.86	28		
30-Oct-04	12		100		0.23	11:15 AM	28.8		7.9	16	1.08	35		
31-Oct-04	13		100				29.7		8.0	4				
1-Nov-04	14		100				26.0		7.9	12				
2-Nov-04	15		100				28.0		7.8	15				
3-Nov-04	16		100				27.0		7.8	0				
4-Nov-04	17		100		0.22		26.5		7.8	0				
5-Nov-04	18		100			10:50 AM	29.0		7.8	20	0.96	32		
6-Nov-04	19		100		0.23	11:00 AM	31.0		7.7	12	0.83	27		
7-Nov-04	20		100			12:00 AM	32.0		7.7	11	1.42	46		
8-Nov-04	21		100		0.21	9:40 AM	29.0			7	0.35	11		
9-Nov-04	22		100			9:30 AM	31.5		7.7	0	0.00	0		
10-Nov-04	23		100			11:25 AM	29.0		7.7	15	0.97	32		
11-Nov-04	24		100			10:20 AM	28.0		7.7	11	0.81	26		
12-Nov-04	25		100			10:10 AM	29.0		7.7	13	0.92	30		
13-Nov-04	26		100		0.24	11:30 AM	32.0		7.7	13	0.86	28		
14-Nov-04	27		100			11:10 AM	32.0		7.8	15	1.06	35		
15-Nov-04	28		100		0.24	10:00 AM	32.0		7.7	13	0.96	31		
16-Nov-04	29		100			9:15 AM	29.0		7.7	2	0.14	5		
17-Nov-04	30		100			9:45 AM	29.0		7.7	9	0.62	20		
18-Nov-04	31		100			10:15 AM	30.0		7.7	15	1.03	34		
19-Nov-04	32		100			10:20 AM	30.0		7.8	7	0.49	16		
20-Nov-04	33		100		0.19	11:30 AM	29.0		7.7	8	0.53	17		
21-Nov-04	34		100			11:30 AM	30.0		7.7	5	0.35	11		
22-Nov-04	35		100			9:00 AM	28.0		7.8	1	0.08	3		
24-Nov-04	37		100			10:55 AM	28.5		7.8	6	0.38	13		
25-Nov-04	38		100			10:00 AM	26.0		7.8	0	0.00	0		
26-Nov-04	39		100			12:30 AM	29.5			6	0.70	23		
27-Nov-04	40		100			11:15 AM	29.0		7.7					
28-Nov-04	41		100		0.27	11:15 AM	27.5		7.8	1	0.07	2		
29-Nov-04	42		100			11:00 AM	28.0		7.8	2	0.14	5		
30-Nov-04	43		100			10:22 AM	27.0		7.7	11	0.79	26		
1-Dec-04	44		100			9:30 AM	27.0		7.8	15	1.09	36		
2-Dec-04	45		100			8:55 AM	29.0		7.9	16	1.15	37		
3-Dec-04	46		100			8:00 AM	28.0		7.8	14	1.02	33		
4-Dec-04	47		100			12:15 PM	30.0		7.8	24	1.43	47		
5-Dec-04	48		100		0.30	9:25 AM	27.0		7.9	26	2.06	67		
6-Dec-04	49		100			9:00 AM	25.0		7.8	13	0.93	30		
7-Dec-04	50		100			9:00 AM	30.0		7.9	10	0.70	23		
8-Dec-04	51		100			11:00 AM	28.0		7.9	12	0.78	25		
9-Dec-04	52		100			11:15 AM	28.0		7.9	1	0.07	2		
10-Dec-04	53		100		0.29	11:40 AM	28.0		7.8	7	0.48	16		
11-Dec-04	54		100			11:50 AM	31.0		7.8	2	0.14	5		
12-Dec-04	55		100				30.0		7.8	6				
13-Dec-04	56		100		0.35		28.0		7.8	6				
14-Dec-04	57		100			9:25 AM	28.0		7.9	5				
15-Dec-04	58		100			9:50 AM				13	0.89	29		
16-Dec-04	59		100		0.35	8:50 AM				19	1.39	45		
17-Dec-04	60		100			3:00 PM								
18-Dec-04	61	13	90	50		9:30 AM	30.0		8.1	25	2.27	84		
19-Dec-04	62	30	90	22		10:40 AM	30.0		7.6	19	1.27	47		
20-Dec-04	63	30	90	22		11:00 AM			7.6	30	2.07	77		
21-Dec-04	64	30	90	22		12:00 PM			7.5	24	1.61	60		
22-Dec-04	65	38	90	17		10:20 AM		7.5	7.4	17	1.28	47		
23-Dec-04	66	38	90	17	0.76	9:30 AM			7.4	21	1.52	56		
24-Dec-04	67	38	90	17	0.62	10:20 AM	27.0		7.3	27	1.83	68	3.39	125
25-Dec-04	68	38	90	17		10:00 AM	29.0		7.5	32	2.27	84		
26-Dec-04	69	38	90	17	0.71	9:45 AM			7.4	31	2.19	81		
27-Dec-04	70	38	90	17		9:50 AM			7.3	32	2.23	83		
28-Dec-04	71	38	90	17	0.69	9:55 AM			7.3	36	2.51	93		
29-Dec-04	72	38	90	17	0.61	11:00 AM	29.0		7.3	40	2.68	99	3.05	113
30-Dec-04	73	38	90	17	0.71	11:30 AM			7.2	42	2.88	107		
31-Dec-04	74	38	90	17		11:30 AM			7.2	42	2.94	109	4.52	167
1-Jan-05	75	38	90	17		12:25 PM			7.2	43	2.90	107		
2-Jan-05	76	38	90	17	0.73	12:35 PM	26.0		7.3	47	3.27	121		
3-Jan-05	77	38	90	17	0.60	11:25 AM	25.0		7.3	40	2.94	109	4.19	155
4-Jan-05	78	38	90	17	0.72	1:25 AM			7.3	51	6.12	226		
5-Jan-05	79	38	90	17	0.65	10:25 AM	29.5		7.3	45	2.29	85	4.56	169
6-Jan-05	80	38	90	17	0.89	12:10 PM	29.0		7.3	60	3.91	145		
7-Jan-05	81	38	90	17	0.62	12:00 PM			7.2	54	3.81	141	5.50	203
8-Jan-05	82	38	90	17		12:50 PM			7.2	54	3.65	135		

Date/Events	Day	Flow Rate (L/day)	Reactor Level (cm)	HRT(hrs)	OLR(kgCOD/ m3/d)	Time	Temp (Celsius)	pH Infl	pH Effl	Gas Gas Count (tips)	Gas Prod. Rate (L/d)	Gas Prod. (L/m3/d)	Expected CH4 prod. (L/d))	Expected CH4 (L/m3/d)
9-Jan-05	83	38	90	17	0.75	2:15 PM	29.0		7.2	57	3.77	139		
10-Jan-05	84	38	90	17	0.69	12:10 PM			7.3	49	3.76	139	5.22	193
11-Jan-05	85	38	90	17		12:05 PM			7.4	59	4.14	153		
12-Jan-05	86	38	90	17		11:30 PM	28.5		7.3	54	2.56	95		
13-Jan-05	87	38	90	17	0.79	12:50 PM			7.3	58	7.31	270		
14-Jan-05	88	38	90	17	0.82	10:30 AM			7.1	48	3.72	138	6.31	233
15-Jan-05	89	38	90	17		11:45 AM			7.2	59	3.93	145		
16-Jan-05	90	38	90	17	0.67	11:37 AM			7.1	58	4.08	151		
17-Jan-05	91	38	90	17	0.63	8:40 AM			7.1	52	4.15	154	4.44	164
18-Jan-05	92	38	90	17		8:50 AM			7.2	52	3.61	134		
19-Jan-05	93	38	90	17	0.58	8:40 AM			7.1	52	3.67	136		
20-Jan-05	94	38	90	17	0.82	9:40 AM			7.2	62	4.17	154		
21-Jan-05	95	38	90	17		11:30 AM			7.3	63	4.10	152	6.72	249
22-Jan-05	96	38	90	17		11:15 AM	30.0		7.1	54	3.82	141		
23-Jan-05	97	38	90	17	0.83	11:05 AM			6.9	57	4.02	149		
24-Jan-05	98	38	90	17	0.65	8:50 AM			7.3	50	3.86	143	5.63	208
25-Jan-05	99	38	90	17	0.76	9:45 AM			7.2	55	3.71	137		
26-Jan-05	100	38	90	17	0.58	9:10 AM	28.0	7.3	7.1	61	4.38	162	5.43	201
27-Jan-05	101	38	90	17	0.80	10:10 AM			7.3	63	4.23	157		
28-Jan-05	102	38	90	17	0.90	9:20 AM			7.3	59	4.28	158	7.06	261
29-Jan-05	103	38	90	17		10:00 AM	32.0		7.2	58	3.95	146		
30-Jan-05	104	38	90	17	0.79	10:50 AM	25.0		7.1	30	2.03	75		
31-Jan-05	105	38	90	17	0.75	8:50 AM	25.0		7.5	40	3.05	113	5.49	203
1-Feb-05	106	38	90	17	0.77	8:50 AM	30.0		7.3	62	4.34	161		
2-Feb-05	107	38	90	17	0.57	8:50 AM	31.0		7.3	53	3.71	137	3.09	114
3-Feb-05	108	38	90	17	0.70	10:50 AM	30.0		7.2	59	3.81	141		
4-Feb-05	109	38	90	17	0.64	9:50 AM	29.0		7.1	54	3.94	146	5.61	207
5-Feb-05	110	38	90	17		11:10 AM	30.0			56	3.71	137		
6-Feb-05	111	38	90	17	0.81	11:40 AM	30.0		7.3	58	3.98	147		
7-Feb-05	112	38	90	17	0.84	8:40 AM	30.0			48	3.84	142	6.60	244
8-Feb-05	113	38	90	17	0.66	9:50 AM	30.0		7.1	55	3.67	136		
9-Feb-05	114	38	90	17	0.56	9:30 AM	30.0		7.2	48	3.41	126	4.75	176
10-Feb-05	115	38	90	17	0.70	10:55 AM	30.0		7.2	49	3.24	120		
11-Feb-05	116	38	90	17	0.61	11:35 AM	30.0		7.2	56	3.81	141	5.59	207
12-Feb-05	117	38	90	17		12:50 PM	30.0	7.8	7.3	53	3.53	130		
13-Feb-05	118	38	90	17	0.71	10:45 AM	30.0		7.4	45	3.45	128		
14-Feb-05	119	38	90	17	0.58	8:35 AM	30.0		7.3	27	2.08	77	4.73	175
15-Feb-05	120	38	90	17	0.65	9:25 AM	30.0	7.3	7.2	17	1.15	43		
16-Feb-05	121	38	90	17	0.55	9:00 AM	31.0			51	3.63	134	4.62	171
17-Feb-05	122	38	90	17	0.79	10:55 AM	30.0	7.6	7.4	27	2.21	82		
18-Feb-05	123	38	90	17	0.59	10:25 AM	30.0	7.6	7.3	58	4.15	153	5.57	206
19-Feb-05	124	38	90	17		11:35 AM	31.0			56	3.74	138		
20-Feb-05	125	38	90	17	0.80	10:10 AM	30.0	8.2	7.4	51	3.79	140		
21-Feb-05	126	38	90	17	0.67	8:40 AM	30.0		7.3	46	3.43	127	6.28	233
22-Feb-05	127		90			9:00 AM	29.0		7.37	53	3.66	135		
22-Feb-05	127	56	90	12	1.00	12:00 PM		7.69						
23-Feb-05	128	56	90	12	0.94	9:10 AM	31.0	7.46	7.27	73	5.79	214	8.31	307
24-Feb-05	129	56	90	12		10:30 AM	30.0		7.13	79	5.24	194		
25-Feb-05	130	56	90	12		11:10 AM	30.0	7.95	7.20	83	5.65	209		
26-Feb-05	131	56	90	12		10:40 AM	30.0		7.45	79	5.65	209		
27-Feb-05	132	56	90	12	1.07	11:50 AM	31.0	7.62		88	5.87	217		
28-Feb-05	133	56	90	12	0.95	8:50 AM	30.0	7.45	7.33	79	6.32	234		
1-Mar-05	134	56	90	12	0.91	11:30 AM	30.0		7.33	86	5.42	200		
2-Mar-05	135	56	90	12	1.37	8:45 AM	30.0	7.45	7.22	67	5.30	196	9.51	352
3-Mar-05	136	56	90	12	1.15	11:10 AM	30.0	7.50	7.10	88	5.60	207		
4-Mar-05	137	56	90	12	0.87	10:15 AM	30.0		7.29	84	6.11	226	7.95	294
5-Mar-05	138	56	90	12		9:40 AM	30.0	7.79		86	6.17	228		
6-Mar-05	139	56	90	12	1.08	11:30 AM	29.0	7.66	7.54	90	5.85	217		
7-Mar-05	140	56	90	12	1.35	8:50 AM	30.0			81	6.38	236	9.77	362
8-Mar-05	141	56	90	12	0.88	8:50 AM	31.0	7.59	7.39	82	5.74	212		
9-Mar-05	142	56	90	12	0.77	9:50 AM	31.0			83	5.58	206	5.94	220
10-Mar-05	143	56	90	12	1.02	9:50 AM	31.0	7.84	7.43	87	6.09	225		
11-Mar-05	144	56	90	12	0.84	10:35 AM	31.0		7.28				4.97	184
12-Mar-05	145	56	90	12		9:20 AM	30.0		7.27	83	6.13	227		
13-Mar-05	146	56	90	12		10:15 AM	30.0			87	5.87	217		
14-Mar-05	147	56	90	12	1.15	8:25 AM	30.0	7.57		65	4.93	182		
15-Mar-05	148	56	90	12		8:25 AM	30.0	7.43	7.51	47	3.29	122		
16-Mar-05	149	56	90	12		8:50 AM	30.0			82	5.64	209		

								pH		Gas				
Date/Events	Day	Flow Rate (L/day)	Reactor Level (cm)	HRT(hrs)	OLR(kgCOD/ m3/d)	Time	Temp (Celsius)	Infl	Effl	Gas Count (tips)	Gas Prod. Rate (L/d)	Gas Prod. (L/m3/d)	Expected CH4 prod. (L/d))	Expected CH4 (L/m3/d)
17-Mar-05	150	56	90	12	1.39	10:30 AM	30.0		7.30	78	5.11	189		
18-Mar-05	151	56	90	12	0.89	9:35 AM	31.0		7.22	74	5.39	199	9.41	348
19-Mar-05	152	56	90	12		11:25 AM	30.0	7.66	7.45	98	6.37	236		
20-Mar-05	153	56	90	12	1.13	11:15 AM	28.0		7.28	68	4.79	177		
21-Mar-05	154	56	90	12	0.94	9:25 AM	25.0	7.68	7.20	58	4.40	163	7.49	277
22-Mar-05	155	56	90	12	1.25	9:40 AM	24.0		7.23	69	4.78	177		
23-Mar-05	156	56	90	12	1.20	8:45 AM	24.0		7.23	70	5.09	188	9.96	368
24-Mar-05	157	56	90	12	1.07	10:25 AM	25.0	7.47	7.30	71	4.65	172		
25-Mar-05	158	56	90	12	0.94	8:30 AM	26.0		7.05	62	4.72	175	6.53	241
26-Mar-05	159	56	90	12		10:00 AM	25.0	7.80	7.30	78	5.14	190		
27-Mar-05	160	56	90	12	0.99	10:45 AM	25.0	7.70	7.31	71	4.82	178		
28-Mar-05	161	56	90	12	1.17	10:15 AM	25.0	7.69	7.14	65	4.65	172	5.39	199
29-Mar-05	162	56	90	12	1.03	9:50 AM	24.0	7.85	7.33	70	4.99	184		
30-Mar-05	163	55	70	9	1.21	10:15 AM	25.0	7.50	7.18	62	4.27	215	6.96	351
31-Mar-05	164	55	70	9	1.31	9:50 AM	26.0		7.15	61	4.35	219		
1-Apr-05	165	55	70	9	1.16	9:50 AM	26.0		7.14	59	4.13	208	5.79	292
2-Apr-05	166	55	70	9		10:20 AM	24.0	7.36	7.13	49	3.36	169		
3-Apr-05	167	55	70	9	1.33	7:40 AM	25.0		7.15	46	3.62	182		
4-Apr-05	168	55	70	9	1.16	9:50 AM	25.0	7.45	7.20	42	2.70	136	5.51	277
5-Apr-05	169	55	70	9	1.29	9:05 AM	27.0	7.14	7.20	31	2.24	113		
6-Apr-05	170	55	70	9	1.14	9:30 AM	25.0			40	2.75	139	4.66	235
6-Apr-05	171	55	70	9		4:00PM								
7-Apr-05	172	55	70	9	1.28	10:25 AM	25.0		7.20	32	2.44	123		
8-Apr-05	173	55	70	9	1.22	11:30 AM	26.0			45	3.01	152	5.10	257
9-Apr-05	174	55	70	9		10:50 AM	25.0		7.20	39	2.81	141		
10-Apr-05	175	88	70	5		1:00 PM	26.0	7.00	6.60	21	1.35	68		
11-Apr-05	176	115	70	4	1.67	9:45 AM	25.0	6.7	6.7	14	1.13	57	7.00	353
11-Apr-05	176		70											
12-Apr-05	177	115	70	4		11:35 AM	25.0	7.0	6.7	NA	NA			
13-Apr-05	178					10:25 AM	25.0		6.8	12	0.88			
14-Apr-05	179	78	80	7		12:40 PM	25.0	6.8	6.7	16	1.02	44		
15-Apr-05	180	90	80	6	1.05	10:50 AM	25.0			17	1.29	55	5.74	245
16-Apr-05	181	72	80	8		11:30 AM	25.0	7.0	6.7	17	1.16	49		
17-Apr-05	182	68	80	8	0.74	1:50 PM	26.0	6.9	6.6	22	1.40	60	3.89	166
18-Apr-05	183					9:15 AM	26.0			14	1.21			
19-Apr-05	184	63	80	9		1:45 PM	26.0	6.6	6.7	1	0.06	3		
20-Apr-05	185	66	80	8		11:00 AM	26.0	7.5	7.0	11	0.87	37		
21-Apr-05	186	55	80	10	0.87	9:40 AM	26.0	7.0	6.7	13	0.96	41	5.54	236
22-Apr-05	187	53	80	11		11:00 AM	25.0	7.1	6.6	16	1.06	45		
23-Apr-05	188	54	80	10		12:35 PM	26.0	7.2	6.8	20	1.31	56		
24-Apr-05	189					11:00 AM	25.0			7	0.52			
25-Apr-05	190	58	80	10		8:45 AM	25.0	7.0	6.5	10	0.77	33		
25-Apr-05	190					6:30 PM	25.0							
26-Apr-05	191	110	65	4	1.33	10:40 AM	25.0	7.2	6.8	0	0.00	0	4.31	215
27-Apr-05	192	110	65	4		9:50 AM	27.0	7.1	6.8	8	0.58	29		
28-Apr-05	193	110	65	4		10:10 AM	25.0		6.8	16	1.10	55		
29-Apr-05	194	104	75	5	1.00	10:40 AM	23.0	7.2	6.8	0	0.00	0	3.92	166
30-Apr-05	195	100	75	6		10:17 AM	27.0	7.2	6.8	0	0.00	0		
1-May-05	196	92	75	6		10:31 AM	25.0			22	1.53	64		
2-May-05	197	84	75	7	0.87	8:19 AM	27.0	7.1	6.7	11	0.85	36	2.75	116
3-May-05	198	94	75	6		10:00 AM	27.0		6.8	15	0.98	41		
4-May-05	199	75	75	8	0.88	10:43 AM		7.1	7.0	10	0.68	29	3.58	151
4-May-05	199					4:40PM								
5-May-05	200	94	75	6		10:35 AM	26.0	7.4	6.9					
6-May-05	201	83	75	7	1.02	10:15 AM	25.0	7.4		15	1.06	45	4.46	188
7-May-05	202	75	75	8		9:38 AM	25.0	7.6	6.7	10	0.72	30		
8-May-05	203	75	75	8		12:02 PM	25.0			16	1.02	43		
9-May-05	204	70	75	8	1.05	10:15 AM	26.0	7.2	6.9	10	0.76	32	4.55	192
10-May-05	205	60	75	9		10:30 AM	24.0	7.1	6.7	6	0.42	18		
11-May-05	206	70	75	8	0.73	10:15 AM	26.0		6.8	4	0.28	12	2.29	97
12-May-05	207	66	75	9		10:30 AM	25.0	7.3	6.8	12	0.83	35		
12-May-05	207					5:30 PM								

								pH		Gas				
Date/Events	Day	Flow Rate (L/day)	Reactor Level (cm)	HRT(hrs)	OLR(kgCOD/ m3/d)	Time	Temp (Celsius)	Infl	Effl	Gas Count (tips)	Gas Prod. Rate (L/d)	Gas Prod. (L/m3/d)	Expected CH4 prod. (L/d))	Expected CH4 (L/m3/d)
13-May-05	208	110	65	4	1.96	10:30 AM	25.0	7.4	6.8	5	0.49	25	7.57	377
14-May-05	209	110	65	4		10:50 AM	25.0	7.5	6.7	12	0.83	41		
15-May-05	210	103	55	4		11:50 AM	25.0			14	0.94	57		
16-May-05	211	103	55	4	1.82	8:35 AM	24.0	7.0	6.7	20	1.62	98	3.34	202
17-May-05	212	100	55	4		10:50 AM	25.0	7.5	7.1	1	0.06	4		
18-May-05	213	100	55	4	1.81	10:40 AM	24.0	7.2	6.7	NA			5.09	308
19-May-05	214	96	55	4		11:10 AM	25.0	7.1	6.8	NA				
20-May-05	215	76	55	5	1.33	10:10 AM	24.0	7.4	7.0	NA			3.19	194
21-May-05	216	88	55	5		1:42 PM	25.0			2	0.12	7		
22-May-05	217	93	61	5		11:31 AM	25.0	7.4	6.7	22	1.69	91		
23-May-05	218	90	65	5		11:15 AM	23.0	7.3	6.8	17	1.20	60		
24-May-05	219	85	65	6		10:20 AM	25.0	7.9	6.5	0	0.00			
25-May-05	220	82	65	6	1.20	11:25 AM	25.0	7.1	6.8	10	0.67	33	4.11	204
26-May-05	221	81	65	6		10:50 AM	25.0	7.3	6.9	9	0.65	32		
27-May-05	222	81	65	6	1.12	10:10 AM	26.0			9	0.65	32	2.64	131
28-May-05	223	76	65	6		10:15 AM	24.0	7.5	6.8	12	0.84	42		
29-May-05	224	73	65	7		10:45 AM	24.0			8	0.55	27		
30-May-05	225	69	65	7	1.03	10:40 AM	24.0	7.1	6.5	NA			2.72	135
31-May-05	226	70	65	7		9:50 AM	24.0	7.1	6.5	4	0.29	14		
31-May-05	226					4:00 PM								
1-Jun-05	227	110	55	4	1.92	10:30 AM	24.0	7.1	6.7	NA			5.35	324
2-Jun-05	228	110	55	4		10:07 AM	25.0			7	0.50	30		
3-Jun-05	229	110	55	4	2.55	11:00 AM	24.0	7.1	6.7	9	0.61	37	8.53	517
4-Jun-05	230	104	55	4		10:00 AM	23.0			4	0.29	18		
5-Jun-05	231	99	55	4		11:33 AM	25.0	7.4	6.7	12	0.79	48		
6-Jun-05	232	96	60	5	1.64	11:00 AM	26.0	7.1	6.8	26	1.86	102	5.44	297
7-Jun-05	233	93	65	5		11:20 AM	24.0	7.0	6.7	15	1.04	52		
8-Jun-05	234	91	65	5	1.17	10:30 AM	26.0	7.1	6.7	6	0.44	22	3.39	169
9-Jun-05	235	83	65	6		8:15 AM	24.0			13	1.00	50		
10-Jun-05	236	78	65	6	1.11	8:40 AM	24.0	6.9	6.5	9	0.62	31	1.06	53
11-Jun-05	237													
12-Jun-05	238													
13-Jun-05	239	68	65	7		10:20 AM	22.0	7.1	6.8	8	0.39	19		
14-Jun-05	240	71	65	7		11:30 AM	24.0	7.0	6.7	4	0.27	13		
15-Jun-05	241	66	65	7	0.79	9:45 AM	22.0	7.0	6.7	1	0.08	4	1.53	76
16-Jun-05	242	68	65	7		10:40 AM	24.0	7.2	6.7	6	0.40	20		
17-Jun-05	243	69	65	7	0.75	10:20 AM	25.0	7.0	6.5	7	0.50	25	2.36	117
18-Jun-05	244	66	65	7		12:10 PM	24.0	7.4	6.8	3	0.20	10		
19-Jun-05	245	65	65	7		11:00 AM	23.0	7.3	6.7	4	0.29	15		
20-Jun-05	246	65	65		0.60	10:15 AM	24.0	7.1	6.8	0	0.00	0	1.05	52
21-Jun-05	247	65	65			10:25 AM	25.0	7.1	6.7	7	0.49	24		
22-Jun-05	248	63	65		0.62	10:50 AM	24.0	7.1	6.7	3	0.21	10	1.70	84
23-Jun-05	249	62	65			11:45 AM	24.0	7.0	6.7	0	0.00	0		
24-Jun-05	250	62	65		0.59	8:20 AM	26.0	7.0	6.5	5	0.41	20	0.69	34
25-Jun-05	251	61	65			10:30 AM	25.0	7.3	6.7	6	0.39	19		
26-Jun-05	252	61	65			12:00 PM	24.0	7.4	6.8	8	0.53	26		
27-Jun-05	253	61	65		0.57	8:00 AM	24.0	7.0	6.7	4	0.34	17	0.83	41
28-Jun-05	254	62	65			11:45 AM	25.0	7.1	6.5	8	0.48	24		
29-Jun-05	255	60	65		0.65	9:15 AM	25.0	7.0	6.8	5	0.39	19	1.65	82
30-Jun-05	256	59	65			10:45 AM	24.0	7.3	6.8	6	0.40	20		
1-Jul-05	257	61	65			9:55 AM	27.0	7.6	7.1	11	0.80	40		
2-Jul-05	258	59	65		0.76	11:35 AM	24.0	7.3	6.8	12	0.79	39	2.75	137
3-Jul-05	259	59	65			1:30 PM	24.0	7.2	6.8	17	1.10	55		
4-Jul-05	260	58	65		0.94	1:30 PM	24.0	7.1	6.5	23	1.61	80	3.63	181

		Permeate Flux												COD Measurements			
														Influent			
Date/Events	Day	Jv USF (ml/min)	Jv USF (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Jv Toray (ml/min)	Jv Toray (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Infl. NA	S Infl. NA	Inf. A	S Infl. A
19-Oct-04	1																
20-Oct-04	2																
21-Oct-04	3			-5.5	2.7	18.6				-15.0	7.4	50.8					
22-Oct-04	4			-6.0	2.9	20.3				-15.0	7.4	50.8					
23-Oct-04	5			-6.0	2.9	20.3				-14.0	6.9	47.4					
24-Oct-04	6			-6.0	2.9	20.3				-12.0	5.9	40.6					
25-Oct-04	7			-7.0	3.4	23.7				-14.0	6.9	47.4					
26-Oct-04	8			-7.0	3.4	23.7				-14.0	6.9	47.4					
27-Oct-04	9	130	15.6	-7.0	3.4	23.7	0.66	4.0		-14.0	6.9	47.4					
28-Oct-04	10			-7.0	3.4	23.7				-13.0	6.4	44.0					
29-Oct-04	11	129	15.5	-7.0	3.4	23.7	0.65	9.0		-15.0	7.4	50.8					
30-Oct-04	12			-17.0	8.3	57.6				-25.0	12.3	84.7				6500mg/d	
31-Oct-04	13																
1-Nov-04	14																
2-Nov-04	15																
3-Nov-04	16																
4-Nov-04	17	137	16.4													6400mg/d	
5-Nov-04	18																
6-Nov-04	19															6500mg/d	
7-Nov-04	20																
8-Nov-04	21															6000mg/d	
9-Nov-04	22																
10-Nov-04	23																
11-Nov-04	24																
12-Nov-04	25																
13-Nov-04	26															6800mg/d	
14-Nov-04	27																
15-Nov-04	28															7000mg/d	
16-Nov-04	29																
17-Nov-04	30																
18-Nov-04	31																
19-Nov-04	32																
20-Nov-04	33															5500mg/d	
21-Nov-04	34																
22-Nov-04	35																
24-Nov-04	37																
25-Nov-04	38																
26-Nov-04	39																
27-Nov-04	40			-25.0	12.3	84.7											
28-Nov-04	41			-15.0	7.4	50.8										7800mg/d	
29-Nov-04	42	14	1.7	-17.0	8.3	57.6	0.03										
30-Nov-04	43																
1-Dec-04	44																
2-Dec-04	45																
3-Dec-04	46																
4-Dec-04	47			-22.0	10.8	74.5											
5-Dec-04	48															8700mg/d	
6-Dec-04	49																
7-Dec-04	50																
8-Dec-04	51																
9-Dec-04	52																
10-Dec-04	53															8400mg/L	
11-Dec-04	54																
12-Dec-04	55																
13-Dec-04	56															10000mg/d	
14-Dec-04	57																
15-Dec-04	58																
16-Dec-04	59	152	18.2													1300	
17-Dec-04	60	160	19.2	-2.0	1.0	6.8	2.83										
18-Dec-04	61	156	18.7	-2.0	1.0	6.8	2.76										
19-Dec-04	62	160	19.2														
20-Dec-04	63	160	19.2	-2.0	1.0	6.8	2.83										
21-Dec-04	64	160	19.2	-2.0	1.0	6.8	2.83										
22-Dec-04	65	160	19.2	-2.0	1.0	6.8	2.83										
23-Dec-04	66	160	19.2											310	138	538	343
24-Dec-04	67	160	19.2													438	338
25-Dec-04	68	160	19.2	-3.0	1.5	10.2	1.89										
26-Dec-04	69	160	19.2	-3.0	1.5	10.2	1.89							290	145	503	310
27-Dec-04	70	160	19.2	-2.5	1.2	8.5	2.27										
28-Dec-04	71	160	19.2	-2.5	1.2	8.5	2.27							310	113	493	320
29-Dec-04	72	160	19.2	-3.0	1.5	10.2	1.89									435	303
30-Dec-04	73													268	113	505	330
31-Dec-04	74	156	18.7	-3.0	1.5	10.2	1.84										
1-Jan-05	75	160	19.2	-3.0	1.5	10.2	1.89										
2-Jan-05	76													285	128	520	350
3-Jan-05	77	156	18.7	-3.0	1.5	10.2	1.84									430	323
4-Jan-05	78	156	18.7	-3.0	1.5	10.2	1.84							298	293	513	363
5-Jan-05	79	152	18.2	-3.0	1.5	10.2	1.80									463	365
6-Jan-05	80	152	18.2	-4.0	2.0	13.5	1.35							305	110	630	335
7-Jan-05	81															443	353
8-Jan-05	82	156	18.7	-4.0	2.0	13.5	1.38										

		Permeate Flux												COD Measurements			
														Influent			
Date/Events	Day	Jv USF (ml/min)	Jv USF (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Jv Toray (ml/min)	Jv Toray (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Infl. NA	S Infl. NA	Inf. A	S Infl. A
9-Jan-05	83													363	150	533	355
10-Jan-05	84	148	17.8	-3.6	1.8	12.2	1.46									488	345
11-Jan-05	85																
12-Jan-05	86	148	17.8	-4.0	2.0	13.5	1.31										
13-Jan-05	87													348	143	560	360
14-Jan-05	88	152	18.2	-3.0	1.5	10.2	1.80									585	395
15-Jan-05	89																
16-Jan-05	90	144	17.3	-4.0	2.0	13.5	1.28							443	125	480	338
17-Jan-05	91															448	398
18-Jan-05	92	140	16.8	-5.0	2.5	16.9	0.99										
19-Jan-05	93															415	365
20-Jan-05	94	140	16.8	-4.0	2.0	13.5	1.24							370	115	583	335
21-Jan-05	95																
22-Jan-05	96	140	16.8	-4.0	2.0	13.5	1.24										
23-Jan-05	97													345	145	590	378
24-Jan-05	98	140	16.8	-4.0	2.0	13.5	1.24									463	385
25-Jan-05	99													420	280	540	320
26-Jan-05	100	140	16.8	-5.0	2.5	16.9	0.99									413	328
27-Jan-05	101													333	120	568	290
28-Jan-05	102															643	298
29-Jan-05	103	140	16.8	-5.0	2.5	16.9	0.99										
30-Jan-05	104													333	115	560	333
31-Jan-05	105	125	15.0	-5.5	2.7	18.6	0.81									535	345
1-Feb-05	106													340	110	548	348
2-Feb-05	107	124	14.9	-5.0	2.5	16.9	0.88									403	310
3-Feb-05	108													333	115	495	300
4-Feb-05	109	132	15.8	-5.0	2.5	16.9	0.94									458	338
5-Feb-05	110																
6-Feb-05	111	125	15.0	-5.5	2.7	18.6	0.81							353	175	573	335
7-Feb-05	112															600	370
8-Feb-05	113	125	15.0	-5.5	2.7	18.6	0.81							265	103	470	298
9-Feb-05	114															400	353
10-Feb-05	115	132	15.8	-5.5	2.7	18.6	0.85							308	100	498	310
11-Feb-05	116															433	313
12-Feb-05	117	128	15.4	-5.5	2.7	18.6	0.82										
13-Feb-05	118													340	115	505	210
14-Feb-05	119	120	14.4	-6.0	2.9	20.3	0.71									413	228
15-Feb-05	120													280	98	463	293
16-Feb-05	121	120	14.4													388	315
17-Feb-05	122	120	14.4	-5.5	2.7	18.6	0.77							328	105	565	338
18-Feb-05	123	120	14.4	-6.0	2.9	20.3	0.71									423	415
19-Feb-05	124																
20-Feb-05	125	108	13.0	-6.0	2.9	20.3	0.64							285	105	568	330
21-Feb-05	126															473	365
22-Feb-05	127	108	13.0	-6.0	2.9	20.3	0.64										
22-Feb-05	127													160	58	485	328
23-Feb-05	128															453	325
24-Feb-05	129	112	13.4	-6.0	2.9	20.3	0.66										
25-Feb-05	130																
26-Feb-05	131	112	13.4	-6.5	3.2	22.0	0.61										
27-Feb-05	132													295	105	518	318
28-Feb-05	133	112	13.4	-7.0	3.4	23.7	0.57									460	335
1-Mar-05	134													288	130	440	308
2-Mar-05	135	112	13.4	-7.0	3.4	23.7	0.57									660	293
3-Mar-05	136													260	115	553	385
4-Mar-05	137	108	13.0	-7.0	3.4	23.7	0.55									418	313
5-Mar-05	138																
6-Mar-05	139	108	13.0	-7.2	3.5	24.4	0.53							313	138	523	348
7-Mar-05	140															650	375
8-Mar-05	141	104	12.5	-7.2	3.5	24.4	0.51							195	77	423	280
9-Mar-05	142															373	243
10-Mar-05	143	104	12.5	-7.5	3.7	25.4	0.49							288	160	490	393
11-Mar-05	144	100		-8.0	3.9	27.1		18.0	7.6	-3.0	1.5	10.2	0.75			403	255
12-Mar-05	145	98	11.8	-7.0	3.4	23.7	0.50	18.0	7.6	-3.0	1.5	10.2	0.75				
13-Mar-05	146	100	12.0	-7.1	3.5	24.0	0.50	16.0	6.8	-4.0	2.0	13.5	0.50				
14-Mar-05	147	103	12.4	-6.5	3.2	22.0	0.56	16.0	6.8	-2.0	1.0	6.8	1.00	230	103	553	388
15-Mar-05	148	97	11.6	-7.5	3.7	25.4	0.46	17.0	7.2	-3.0	1.5	10.2	0.71				
16-Mar-05	149	96	11.5	-8.2	4.0	27.8	0.41	15.0	6.3	-5.0	2.5	16.9	0.37				

		Permeate Flux												COD Measurements			
														Influent			
Date/Events	Day	Jv USF (ml/min)	Jv USF (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Jv Toray (ml/min)	Jv Toray (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Infl. NA	S Infl. NA	Inf. A	S Infl. A
17-Mar-05	150	96	11.5	-8.0	3.9	27.1	0.43	15.5	6.6	-3.5	1.7	11.9	0.55	273	123	670	293
18-Mar-05	151	94	11.3	-8.0	3.9	27.1	0.42	15.5	6.6	-4.5	2.2	15.2	0.43			428	
19-Mar-05	152	93	11.1	-8.0	3.9	27.1	0.41	15.5	6.6	-5.0	2.5	16.9	0.39				
20-Mar-05	153	88	10.6	-8.5	4.2	28.8	0.37	14.5	6.1	-6.0	2.9	20.3	0.30	423	130	545	328
21-Mar-05	154	84	10.1	-7.8	3.8	26.4	0.38	13.0	5.5	-7.0	3.4	23.7	0.23			455	375
22-Mar-05	155	83	10.0	-7.5	3.7	25.4	0.39	11.0	4.7	-7.0	3.4	23.7	0.20	430	115	603	323
23-Mar-05	156	80	9.6	-8.0	3.9	27.1	0.35	10.0	4.2	-7.0	3.4	23.7	0.18			578	338
24-Mar-05	157	83	10.0	-8.0	3.9	27.1	0.37	12.0	5.1	-6.5	3.2	22.0	0.23	338	108	518	323
25-Mar-05	158	75	9.0	-8.0	3.9	27.1	0.33	12.0	5.1	-5.0	2.5	16.9	0.30			453	408
26-Mar-05	159	81	9.7	-8.5	4.2	28.8	0.34	13.0	5.5	-6.0	2.9	20.3	0.27				
27-Mar-05	160	83	10.0	-8.9	4.4	30.1	0.33	14.0	5.9	-6.5	3.2	22.0	0.27	348	120	480	358
28-Mar-05	161	81	9.7	-8.5	4.2	28.8	0.34	13.5	5.7	-6.0	2.9	20.3	0.28			563	438
29-Mar-05	162	76	9.1	-8.5	4.2	28.8	0.32	12.0	5.1	-7.0	3.4	23.7	0.21	295	100	495	295
30-Mar-05	163	73	8.8	-8.3	4.1	28.1	0.31	14.0	5.9	-5.5	2.7	18.6	0.32			435	328
31-Mar-05	164	67	8.0	-8.0	3.9	27.1	0.30	11.0	4.7	-7.5	3.7	25.4	0.18	343	118	473	315
1-Apr-05	165	69	8.3	-8.5	4.2	28.8	0.29	11.0	4.7	-8.0	3.9	27.1	0.17			420	328
2-Apr-05	166	68	8.2	-8.3	4.1	28.1	0.29	10.0	4.2	-7.9	3.9	26.8	0.16				
3-Apr-05	167													328	153	480	323
4-Apr-05	168	63	7.6	-9.0	4.4	30.5	0.25	11.0	4.7	-8.5	4.2	28.8	0.16			417	348
5-Apr-05	169	63	7.6	-8.5	4.2	28.8	0.26	8.0	3.4	-9.0	4.4	30.5	0.11	285	85	465	295
6-Apr-05	170	63	7.6	-9.0	4.4	30.5	0.25	7.5	3.2	-9.5	4.7	32.2	0.10			410	340
6-Apr-05	171	120	14.4	-3.0	1.5	10.2	1.42	15.0	6.3	-4.0	2.0	13.5	0.47				
7-Apr-05	172	100	12.0	-7.0	3.4	23.7	0.51	13.0	5.5	-5.0	2.5	16.9	0.32	208	70	463	348
8-Apr-05	173	93	11.2	-7.0	3.4	23.7	0.47	13.0	5.5	-5.5	2.7	18.6	0.30			440	350
9-Apr-05	174	86	10.3	-7.0	3.4	23.7	0.44	14.0	5.9	-5.0	2.5	16.9	0.35				
10-Apr-05	175	85	10.2	-7.5	3.7	25.4	0.40	14.0	5.9	-4.8	2.4	16.3	0.36				
11-Apr-05	176	83	10.0	-8.0	3.9	27.1	0.37	14.0	5.9	-5.5	2.7	18.6	0.32	288			
11-Apr-05	176	88	10.6	-7.0	3.4	23.7	0.45	14.0	5.9								
12-Apr-05	177	74	8.9	-8.5	4.2	28.8	0.31	14.0	5.9	-5.5	2.7	18.6	0.32				
13-Apr-05	178	62	7.4											453	103		
14-Apr-05	179	58	7.0	-9.8	4.8	33.2	0.21	9.5	4.0	-9.3	4.6	31.5	0.13				
15-Apr-05	180	66	7.9					12.0	5.1					275	165		
16-Apr-05	181	53	6.4	-10.0	4.9	33.9	0.19	9.5	4.0	-10.4	5.1	35.2	0.11				
17-Apr-05	182	50	6.0	-10.5	5.2	35.6	0.17	9.0	3.8	-10.0	4.9	33.9	0.11	255	93		
18-Apr-05	183							9.5	4.0								
19-Apr-05	184	46	5.5	-9.5	4.7	32.2	0.17	9.0	3.8	-10.5	5.2	35.6	0.11				
20-Apr-05	185	50	6.0	-10.5	5.2	35.6	0.17	7.5	3.2	-10.5	5.2	35.6	0.09				
21-Apr-05	186	40	4.8	-10.5	5.2	35.6	0.13	8.0	3.4	-10.5	5.2	35.6	0.10	370	105		
22-Apr-05	187	39	4.7	-10.5	5.2	35.6	0.13	7.0	3.0	-11.0	5.4	37.3	0.08				
23-Apr-05	188	40	4.8	-11.5	5.6	38.9	0.12	7.0	3.0	-11.5	5.6	38.9	0.08				
24-Apr-05	189																
25-Apr-05	190	40	4.8	-11.0	5.4	37.3	0.13	10.0	4.2	-10.5	5.2	35.6	0.12				
25-Apr-05	190	115	6.9					10.0	4.2								
26-Apr-05	191	100	6.0	-5.5	2.7	18.6	0.32	10.0	4.2	-7.0	3.4	23.7	0.18	243	128		
27-Apr-05	192	90	5.4	-7.5	3.7	25.4	0.21	10.0	4.2	-9.0	4.4	30.5	0.14				
28-Apr-05	193	90	5.4	-8.5	4.2	28.8	0.19	10.0	4.2	-9.0	4.4	30.5	0.14				
29-Apr-05	194	80	4.8	-9.0	4.4	30.5	0.16	10.0	4.2	-8.0	3.9	27.1	0.16	228	100		
30-Apr-05	195	77	4.6	-9.8	4.8	33.2	0.14	10.0	4.2	-8.5	4.2	28.8	0.15				
1-May-05	196	70	4.2	-10.0	4.9	33.9	0.12	10.0	4.2	-8.5	4.2	28.8	0.15				
2-May-05	197	63	3.8	-10.6	5.2	35.9	0.11	10.0	4.2	-9.0	4.4	30.5	0.14	245	110		
3-May-05	198	70	4.2	-10.0	4.9	33.9	0.12	12.0	5.1	-8.0	3.9	27.1	0.19				
4-May-05	199	55	3.3	-10.5	5.2	35.6	0.09	10.0	4.2	-9.0	4.4	30.5	0.14	278	103		
4-May-05	199	85	5.1	-12.0	5.9	40.6	0.13	13.0	5.5	-7.0	3.4	23.7	0.23				
5-May-05	200	72	4.3	-11.5	5.6	38.9	0.11	10.0	4.2	-9.5	4.7	32.2	0.13				
6-May-05	201	62	3.7	-10.0	4.9	33.9	0.11	10.0	4.2	-9.0	4.4	30.5	0.14	290	133		
7-May-05	202	55	3.3	-10.0	4.9	33.9	0.10	10.0	4.2	-9.0	4.4	30.5	0.14				
8-May-05	203	56	3.4	-10.0	4.9	33.9	0.10	9.3	3.9	-9.0	4.4	30.5	0.13				
9-May-05	204	50	3.0	-10.0	4.9	33.9	0.09	11.0	4.7	-9.0	4.4	30.5	0.15	355	130		
10-May-05	205	43	2.6	-9.0	4.4	30.5	0.08	9.0	3.8	-8.0	3.9	27.1	0.14				
11-May-05	206	51	3.1	-9.8	4.8	33.2	0.09	9.5	4.0	-9.0	4.4	30.5	0.13	248	75		
12-May-05	207	47	2.8	-10.0	4.9	33.9	0.08	10.0	4.2	-7.0	3.4	23.7	0.18				
12-May-05	207	135	8.1	-6.0	2.9	20.3	0.40	15.0	6.3	-5.5	2.7	18.6	0.34				

		Permeate Flux												COD Measurements			
														Influent			
Date/Events	Day	Jv USF (ml/min)	Jv USF (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Jv Toray (ml/min)	Jv Toray (lmh)	TMP (Hg")	Psi	kPa	Jv/kPa	Infl. NA	S Infl. NA	Inf. A	S Infl. A
13-May-05	208	120	7.2	-6.0	2.9	20.3	0.35	9.0	3.8	-9.5	4.7	32.2	0.12	358	133		
14-May-05	209	93	5.6	-8.0	3.9	27.1	0.21	10.0	4.2	-8.0	3.9	27.1	0.16				
15-May-05	210	80	4.8	-8.5	4.2	28.8	0.17	9.0	3.8	-10.0	4.9	33.9	0.11				
16-May-05	211	80	4.8	-8.5	4.2	28.8	0.17	9.0	3.8	-9.0	4.4	30.5	0.12	293	235		
17-May-05	212	77	4.6	-8.5	4.2	28.8	0.16	10.0	4.2	-8.0	3.9	27.1	0.16				
18-May-05	213	76	4.6	-8.5	4.2	28.8	0.16	11.0	4.7	-7.0	3.4	23.7	0.20	298	130		
19-May-05	214	72	4.3	-8.5	4.2	28.8	0.15	11.0	4.7	-6.5	3.2	22.0	0.21				
20-May-05	215	55	3.3	-8.0	3.9	27.1	0.12	11.0	4.7	-6.0	2.9	20.3	0.23	288	113		
21-May-05	216	65	3.9	-8.5	4.2	28.8	0.14	11.0	4.7	-6.5	3.2	22.0	0.21				
22-May-05	217	69	4.1	-8.0	3.9	27.1	0.15	12.0	5.1	-6.2	3.0	21.0	0.24				
23-May-05	218	66	4.0	-8.0	3.9	27.1	0.15	12.0	5.1	-6.5	3.2	22.0	0.23				
24-May-05	219	63	3.8	-8.0	3.9	27.1	0.14	11.0	4.7	-6.0	2.9	20.3	0.23				
25-May-05	220	60	3.6	-8.5	4.2	28.8	0.13	11.0	4.7	-7.0	3.4	23.7	0.20	295	150		
26-May-05	221	59	3.5	-10.0	4.9	33.9	0.10	11.0	4.7	-7.0	3.4	23.7	0.20				
27-May-05	222	59	3.5	-10.5	5.2	35.6	0.10	11.0	4.7	-8.5	4.2	28.8	0.16	280	193		
28-May-05	223	55	3.3	-10.5	5.2	35.6	0.09	11.0	4.7	-8.5	4.2	28.8	0.16				
29-May-05	224	52	3.1	-10.5	5.2	35.6	0.09	11.0	4.7	-8.5	4.2	28.8	0.16				
30-May-05	225	50	3.0	-10.0	4.9	33.9	0.09	10.0	4.2	-8.5	4.2	28.8	0.15	300	200		
31-May-05	226	50	3.0	-10.0	4.9	33.9	0.09	11.0	4.7	-8.5	4.2	28.8	0.16				
31-May-05	226	128	7.7	-5.0	2.5	16.9	0.45	14.5	6.1	-4.0	2.0	13.5	0.45				
1-Jun-05	227	110	6.6	-6.0	2.9	20.3	0.32	11.0	4.7	-8.0	3.9	27.1	0.17	288	138		
2-Jun-05	228	98	5.9	-7.0	3.4	23.7	0.25	10.5	4.4	-8.0	3.9	27.1	0.16				
3-Jun-05	229	88	5.3	-7.5	3.7	25.4	0.21	10.0	4.2	-8.0	3.9	27.1	0.16	383	138		
4-Jun-05	230	80	4.8	-8.0	3.9	27.1	0.18	10.0	4.2	-8.5	4.2	28.8	0.15				
5-Jun-05	231	76	4.6	-8.5	4.2	28.8	0.16	10.0	4.2	-8.0	3.9	27.1	0.16				
6-Jun-05	232	73	4.4	-8.0	3.9	27.1	0.16	10.0	4.2	-7.0	3.4	23.7	0.18	313	155		
7-Jun-05	233	70	4.2	-9.5	4.7	32.2	0.13	11.0	4.7	-7.0	3.4	23.7	0.20				
8-Jun-05	234	69	4.1	-9.5	4.7	32.2	0.13	10.0	4.2	-6.0	2.9	20.3	0.21	258	105		
9-Jun-05	235	62	3.7	-9.5	4.7	32.2	0.12	10.0	4.2	-7.5	3.7	25.4	0.17				
10-Jun-05	236	58	3.5	-10.0	4.9	33.9	0.10	10.0	4.2	-8.0	3.9	27.1	0.16	285	113		
11-Jun-05	237																
12-Jun-05	238																
13-Jun-05	239	49	2.9	-8.0	3.9	27.1	0.11	10.0	4.2	-8.5	4.2	28.8	0.15				
14-Jun-05	240	52	3.1	-10.0	4.9	33.9	0.09	10.0	4.2	-8.0	3.9	27.1	0.16				
15-Jun-05	241	48	2.9	-10.0	4.9	33.9	0.09	9.0	3.8	-7.5	3.7	25.4	0.15	243	133		
16-Jun-05	242	49	2.9	-11.0	5.4	37.3	0.08	10.0	4.2	-8.5	4.2	28.8	0.15				
17-Jun-05	243	49	2.9	-10.5	5.2	35.6	0.08	11.0	4.7	-8.0	3.9	27.1	0.17	218	108		
18-Jun-05	244	47	2.8	-10.5	5.2	35.6	0.08	10.0	4.2	-8.0	3.9	27.1	0.16				
19-Jun-05	245	46	2.8	-11.0	5.4	37.3	0.07	10.0	4.2	-8.0	3.9	27.1	0.16				
20-Jun-05	246	46	2.8	-10.5	5.2	35.6	0.08	10.0	4.2	-7.5	3.7	25.4	0.17	183	140		
21-Jun-05	247	46	2.8	-11.5	5.6	38.9	0.07	10.0	4.2	-8.5	4.2	28.8	0.15				
22-Jun-05	248	45	2.7	-11.0	5.4	37.3	0.07	10.0	4.2	-8.5	4.2	28.8	0.15	193	125		
23-Jun-05	249	44	2.6	-11.0	5.4	37.3	0.07	10.0	4.2	-8.5	4.2	28.8	0.15				
24-Jun-05	250	44	2.6	-11.0	5.4	37.3	0.07	10.0	4.2	-8.0	3.9	27.1	0.16	188	115		
25-Jun-05	251	43	2.6	-11.0	5.4	37.3	0.07	10.0	4.2	-8.5	4.2	28.8	0.15				
26-Jun-05	252	43	2.6	-11.5	5.6	38.9	0.07	10.0	4.2	-8.5	4.2	28.8	0.15				
27-Jun-05	253	43	2.6	-10.5	5.2	35.6	0.07	10.0	4.2	-7.5	3.7	25.4	0.17	183	93		
28-Jun-05	254	43	2.6	-11.0	5.4	37.3	0.07	11.0	4.7	-8.0	3.9	27.1	0.17				
29-Jun-05	255	42	2.5	-11.0	5.4	37.3	0.07	10.0	4.2	-8.5	4.2	28.8	0.15	215	98		
30-Jun-05	256	41	2.5	-11.5	5.6	38.9	0.06	10.0	4.2	-9.0	4.4	30.5	0.14				
1-Jul-05	257	43	2.6	-11.0	5.4	37.3	0.07	10.0	4.2	-8.0	3.9	27.1	0.16				
2-Jul-05	258	41	2.5	-11.0	5.4	37.3	0.07	10.0	4.2	-8.0	3.9	27.1	0.16	255	88		
3-Jul-05	259	41	2.5	-11.5	5.6	38.9	0.06	10.0	4.2	-8.5	4.2	28.8	0.15				
4-Jul-05	260	40	2.4	-11.5	5.6	38.9	0.06	10.0	4.2	-8.5	4.2	28.8	0.15	323	105		

		(mg/L)		TSS/VSS Measurements (mg/L)							VFA (mg/L)				
		Effluent		Influent		Centrifuge method		Filtration method		Expected	Influent				
Date/Events	Day	USF	Toray	TSS	VSS	TSS	VSS	TSS	VSS	VSS	A Acetic	A Propionic	NA Acetic	NA Propionic	USF Acetic
19-Oct-04	1														
20-Oct-04	2														
21-Oct-04	3														
22-Oct-04	4														
23-Oct-04	5														
24-Oct-04	6														
25-Oct-04	7														
26-Oct-04	8														
27-Oct-04	9														
28-Oct-04	10														
29-Oct-04	11														
30-Oct-04	12														
31-Oct-04	13														
1-Nov-04	14														
2-Nov-04	15														
3-Nov-04	16														
4-Nov-04	17														
5-Nov-04	18														
6-Nov-04	19														
7-Nov-04	20														
8-Nov-04	21														
9-Nov-04	22														
10-Nov-04	23														
11-Nov-04	24														
12-Nov-04	25														
13-Nov-04	26														
14-Nov-04	27														
15-Nov-04	28														
16-Nov-04	29	4100													
17-Nov-04	30														
18-Nov-04	31														
19-Nov-04	32														
20-Nov-04	33	3240													
21-Nov-04	34														
22-Nov-04	35														
24-Nov-04	37														
25-Nov-04	38														
26-Nov-04	39														
27-Nov-04	40	3575													
28-Nov-04	41														
29-Nov-04	42							5298	4822						
30-Nov-04	43														
1-Dec-04	44														
2-Dec-04	45														
3-Dec-04	46														
4-Dec-04	47														
5-Dec-04	48														
6-Dec-04	49	2500						5800	4332						
7-Dec-04	50														
8-Dec-04	51														
9-Dec-04	52														
10-Dec-04	53	2925													
11-Dec-04	54														
12-Dec-04	55														
13-Dec-04	56	3475						6050	4575						
14-Dec-04	57														
15-Dec-04	58												8.353	0.000	
16-Dec-04	59							8350	5825		618.944	4.159			1043.080
17-Dec-04	60										280.323	2.930			
18-Dec-04	61	2113						6925	5125		196.112	4.903	48.581	2.067	
19-Dec-04	62										144.945	2.140	2.188	0.000	
20-Dec-04	63	938									93.687	6.313			
21-Dec-04	64														
22-Dec-04	65	260											11.718	1.665	964.288
23-Dec-04	66										190.920	2.245	9.970	1.448	
24-Dec-04	67	233									190.593	4.038			122.460
25-Dec-04	68												10.504	1.700	
26-Dec-04	69	213													
27-Dec-04	70							6700	5325		199.681	3.197	11.380	1.681	82.212
28-Dec-04	71										206.962	2.216			
29-Dec-04	72	235									180.447	3.297			54.994
30-Dec-04	73														
31-Dec-04	74	165											10.464	1.954	55.440
1-Jan-05	75														
2-Jan-05	76										221.556	4.200	13.700	1.800	
3-Jan-05	77	160						6090	4744		205.544	4.000			57.670
4-Jan-05	78														
5-Jan-05	79	145													
6-Jan-05	80										218.233	3.648	13.7	1.902	
7-Jan-05	81	123									170.234	4.818			0.000
8-Jan-05	82														

		(mg/L)		TSS/VSS Measurements (mg/L)							VFA (mg/L)				
		Effluent		Influent		Centrifuge method		Filtration method		Expected	Influent				
Date/Events	Day	USF	Toray	TSS	VSS	TSS	VSS	TSS	VSS	VSS	A Acetic	A Propionic	NA Acetic	NA Propionic	USF Acetic
9-Jan-05	83										226.256	4.120			
10-Jan-05	84	118						6050	4850		183.302	4.724			0.000
11-Jan-05	85														
12-Jan-05	86														
13-Jan-05	87										189.219	3.077	12.958	0.000	
14-Jan-05	88	98													0.000
15-Jan-05	89														
16-Jan-05	90										155.742	2.029	0.000	0.000	
17-Jan-05	91	130									145.041	0.000			0.000
18-Jan-05	92														
19-Jan-05	93	100									156.617	0.000	11.072	2.911	0.000
20-Jan-05	94														
21-Jan-05	95	78													
22-Jan-05	96														
23-Jan-05	97												11.303	2.284	
24-Jan-05	98	103						6831			203.661	4.630			0.000
25-Jan-05	99														
26-Jan-05	100	68													
27-Jan-05	101										188.366	2.506	14.432	2.562	
28-Jan-05	102	75									188.406	0.000			0.000
29-Jan-05	103														
30-Jan-05	104										195.846	3.005	10.179	2.101	35.411
31-Jan-05	105	135				6825	6100	6750	5450		204.766	7.159			
1-Feb-05	106										186.574	3.855	0.000		
2-Feb-05	107	243									185.844	3.684			0.000
3-Feb-05	108										170.249	2.739	14.176	2.865	
4-Feb-05	109	55									196.353	1.683			0.000
5-Feb-05	110														
6-Feb-05	111										198.787	6.759	15.387	3.652	
7-Feb-05	112	90									199.653	6.696			4.573
8-Feb-05	113										189.874	1.692	0.000	0.000	
9-Feb-05	114	78				6475	5250	6900	5350		197.476	2.906			0.000
10-Feb-05	115										191.173	2.465	10.281	1.732	
11-Feb-05	116	45									201.323	3.445			0.000
12-Feb-05	117														
13-Feb-05	118										84.297	2.362	0.000	0.000	
14-Feb-05	119	103				7250	5450	6900	5500		86.725	5.370			0.000
15-Feb-05	120										173.583	4.898	0.000	0.000	
16-Feb-05	121	78									199.193	3.985			0.000
17-Feb-05	122										201.303	1.854	9.753	1.719	
18-Feb-05	123	75									199.004	2.681			0.000
19-Feb-05	124														
20-Feb-05	125										183.707	0.000	2.517	0.000	
21-Feb-05	126	48				8725	7150	7400	5950		203.366	3.081			4.120
22-Feb-05	127														
22-Feb-05	127										178.576	2.129	0.000	0.000	
23-Feb-05	128	45									194.471	4.183			5.629
24-Feb-05	129														
25-Feb-05	130	68													
26-Feb-05	131														
27-Feb-05	132										171.083	3.179	5.662	1.773	
28-Feb-05	133					7450	5950	7800	6300		199.699	6.859			1.721
1-Mar-05	134										179.033	3.263	7.185	0.000	
2-Mar-05	135	65									164.020	5.202			0.000
3-Mar-05	136										187.077	3.090	6.797	0.000	1.589
4-Mar-05	137	80									168.492	5.511			
5-Mar-05	138														
6-Mar-05	139										185.447	3.477	14.777	3.394	
7-Mar-05	140	88				8075	6250	8000	6300		201.302	9.071			0.000
8-Mar-05	141										171.878	2.526	0.000	0.000	37.100
9-Mar-05	142	95									141.300	3.723			
10-Mar-05	143										187.368	3.174	4.261	0.000	
11-Mar-05	144	263	123								140.061	6.264			1.300
12-Mar-05	145														
13-Mar-05	146														
14-Mar-05	147	73	123			7750	5950	7400	5850		237.458	7.267	13.038	2.876	0.000
15-Mar-05	148										202.117	2.860	15.288	2.761	0.000
16-Mar-05	149										167.544	3.744			5.413

		mg/L)		TSS/VSS Measurements (mg/L)							VFA (mg/L)				
		Effluent		Influent		Centrifuge method		Filtration method		Expected	Influent				
Date/Events	Day	USF	Toray	TSS	VSS	TSS	VSS	TSS	VSS	VSS	A Acetic	A Propionic	NA Acetic	NA Propionic	USF Acetic
17-Mar-05	150										181.564	3.760	11.688	3.017	
18-Mar-05	151	60	78								162.926	4.925			
19-Mar-05	152														
20-Mar-05	153										196.459	5.971	24.834	6.133	
21-Mar-05	154	123	113			6850	6175	7900	6450		201.123	8.088			23.643
22-Mar-05	155										201.461	4.094	10.395	4.110	
23-Mar-05	156	75	90								210.147	6.739			13.731
24-Mar-05	157										193.497	4.538	8.645	2.291	
25-Mar-05	158	205	100								211.234	7.884			21.707
26-Mar-05	159														
27-Mar-05	160										191.682	3.143	14.390	2.672	
28-Mar-05	161	390	103			7825	6300	7750	6200		184.718	5.469			24.764
29-Mar-05	162										224.806	2.973	5.660	0.653	
30-Mar-05	163	100	107												21.673
31-Mar-05	164										203.917	2.256	8.837	1.840	
1-Apr-05	165	143	148								228.776	4.584			61.471
2-Apr-05	166														
3-Apr-05	167										218.973	4.481	17.253	3.496	
4-Apr-05	168	160	165								215.735	4.906			75.967
5-Apr-05	169										195.678	3.863	11.733	2.256	
6-Apr-05	170	213	178			7850	6225	8150	6600		207.011	2.726			79.547
6-Apr-05	171														
7-Apr-05	172										204.531	0.000	2.545	0.000	
8-Apr-05	173	180	193								192.433	1.041			94.439
9-Apr-05	174														
10-Apr-05	175														
11-Apr-05	176	105	123			9085	7429	8742	7200				14.322	1.449	1.329
11-Apr-05	176														
12-Apr-05	177														
13-Apr-05	178	80	80										6.671	0.000	3.761
14-Apr-05	179														
15-Apr-05	180	90	95										6.343	0.711	5.797
16-Apr-05	181														
17-Apr-05	182	100	83			8800	6300	8950	6450				6.015	0.633	8.304
18-Apr-05	183														
19-Apr-05	184														
20-Apr-05	185														
21-Apr-05	186	85	83										14.466	0.851	7.108
22-Apr-05	187														
23-Apr-05	188														
24-Apr-05	189														
25-Apr-05	190														
25-Apr-05	190														
26-Apr-05	191	145	117										11.050	2.866	30.505
27-Apr-05	192														
28-Apr-05	193														
29-Apr-05	194	125	115			10525	8663	11900	9600				11.960	0.775	30.126
30-Apr-05	195														
1-May-05	196														
2-May-05	197	150	153										15.730	2.297	37.989
3-May-05	198														
4-May-05	199	145	138			9450	7550	9650	8200				0.000	0.000	36.295
4-May-05	199														
5-May-05	200														
6-May-05	201	140	133										11.452	1.673	34.620
7-May-05	202														
8-May-05	203														
9-May-05	204	180	160			9350	7775	9550	7750				19.539	3.291	44.351
10-May-05	205														
11-May-05	206	148	160										0.000	0.000	35.865
12-May-05	207														
12-May-05	207														

		(mg/L)		TSS/VSS Measurements (mg/L)							VFA (mg/L)				
		Effluent		Influent		Centrifuge method		Filtration method		Expected	Influent				
Date/Events	Day	USF	Toray	TSS	VSS	TSS	VSS	TSS	VSS	VSS	A Acetic	A Propionic	NA Acetic	NA Propionic	USF Acetic
13-May-05	208	145	178										16.612	2.914	46.098
14-May-05	209														
15-May-05	210														
16-May-05	211	210	190			9250	7600	9500	7600				15.858	1.273	49.271
17-May-05	212														
18-May-05	213	133	173										12.210	0.000	35.606
19-May-05	214														
20-May-05	215	213	123					3800					15.594	2.739	35.426
21-May-05	216														
22-May-05	217														
23-May-05	218														
24-May-05	219														
25-May-05	220	123	180										14.590	1.520	38.597
26-May-05	221														
27-May-05	222	158	215										17.380	3.910	56.620
28-May-05	223														
29-May-05	224														
30-May-05	225	140	235			5775	4825	6000	5150				15.410	3.740	60.000
31-May-05	226														
31-May-05	226														
1-Jun-05	227	153	145										23.760	3.070	46.840
2-Jun-05	228														
3-Jun-05	229	153	170										17.750	2.460	41.610
4-Jun-05	230														
5-Jun-05	231														
6-Jun-05	232	148	153			9050	7175	8550	7100				21.520	3.960	47.030
7-Jun-05	233														
8-Jun-05	234	158	145										16.370	2.160	42.190
9-Jun-05	235														
10-Jun-05	236	240	253										25.640	2.800	45.590
11-Jun-05	237														
12-Jun-05	238														
13-Jun-05	239	165	150			8450		8050							57.300
14-Jun-05	240														
15-Jun-05	241	160	193										11.550	0.000	66.520
16-Jun-05	242														
17-Jun-05	243	113	128										15.210	0.000	42.190
18-Jun-05	244														
19-Jun-05	245														
20-Jun-05	246	130	143										14.490	0.000	47.780
21-Jun-05	247														
22-Jun-05	248	115	118												
23-Jun-05	249														
24-Jun-05	250	145	168												
25-Jun-05	251														
26-Jun-05	252														
27-Jun-05	253	138	150			8500	7200	8290	7000						
28-Jun-05	254														
29-Jun-05	255	128	145												
30-Jun-05	256														
1-Jul-05	257														
2-Jul-05	258	123	120												
3-Jul-05	259														
4-Jul-05	260	128	158												

					Gas Analysis						Ammonia Measurements (mg N/L)			
		Effluent			Percent Composition				CO2/CH4 % Ratio		influent		Effluent	
Date/Events	Day	USF Propionic	TORAY Acetic	TORAY Propionic	CO2	O2	N2	CH4	CO2	CH4	Infl. NA	Inf. A	USF	Toray
19-Oct-04	1													
20-Oct-04	2													
21-Oct-04	3													
22-Oct-04	4													
23-Oct-04	5													
24-Oct-04	6													
25-Oct-04	7													
26-Oct-04	8													
27-Oct-04	9													
28-Oct-04	10													
29-Oct-04	11													
30-Oct-04	12													
31-Oct-04	13													
1-Nov-04	14													
2-Nov-04	15				9.03%	0.36%	44.68%	45.93%	16.43%	83.57%				
3-Nov-04	16													
4-Nov-04	17													
5-Nov-04	18													
6-Nov-04	19													
7-Nov-04	20													
8-Nov-04	21				10.13%	0.17%	27.66%	62.03%	14.04%	85.96%				
9-Nov-04	22													
10-Nov-04	23													
11-Nov-04	24				9.49%	0.00%	34.75%	55.76%	14.55%	85.45%				
12-Nov-04	25													
13-Nov-04	26				9.77%	0.00%	30.86%	59.37%	14.13%	85.87%				
14-Nov-04	27													
15-Nov-04	28				9.36%	0.32%	28.58%	61.75%	13.16%	86.84%				
16-Nov-04	29													
17-Nov-04	30													
18-Nov-04	31				8.69%	0.81%	35.46%	55.04%	13.63%	86.37%				
19-Nov-04	32													
20-Nov-04	33				8.56%	0.18%	33.19%	58.07%	12.85%	87.15%				
21-Nov-04	34													
22-Nov-04	35				7.76%	0.19%	31.16%	60.89%	11.30%	88.70%				
24-Nov-04	37													
25-Nov-04	38				6.94%	0.48%	32.18%	60.40%	10.30%	89.70%				
26-Nov-04	39													
27-Nov-04	40													
28-Nov-04	41													
29-Nov-04	42				6.97%	1.53%	32.22%	59.28%	10.52%	89.48%				
30-Nov-04	43													
1-Dec-04	44													
2-Dec-04	45				6.18%	0.32%	22.56%	70.94%	8.02%	91.98%				
3-Dec-04	46													
4-Dec-04	47													
5-Dec-04	48													
6-Dec-04	49				6.06%	0.20%	15.44%	78.29%	7.18%	92.82%				
7-Dec-04	50													
8-Dec-04	51													
9-Dec-04	52													
10-Dec-04	53													
11-Dec-04	54													
12-Dec-04	55													
13-Dec-04	56				5.55%	0.22%	13.14%	81.10%	6.40%	93.60%				
14-Dec-04	57													
15-Dec-04	58													
16-Dec-04	59	79.660			5.76%	0.00%	16.00%	78.24%	6.86%	93.14%				
17-Dec-04	60													
18-Dec-04	61													
19-Dec-04	62													
20-Dec-04	63													
21-Dec-04	64													
22-Dec-04	65	26.894			1.89%	0.20%	58.52%	39.40%	4.57%	95.43%				
23-Dec-04	66													
24-Dec-04	67	6.392			1.88%	0.38%	47.65%	50.08%	3.62%	96.38%				
25-Dec-04	68													
26-Dec-04	69										30.2	27.8		
27-Dec-04	70	10.080			2.43%	0.35%	32.71%	64.51%	3.63%	96.37%			38.5	
28-Dec-04	71													
29-Dec-04	72	1.739			2.44%	0.38%	27.33%	69.85%	3.38%	96.62%				
30-Dec-04	73													
31-Dec-04	74	10.304			4.45%	0.00%	21.33%	74.22%	5.65%	94.35%				
1-Jan-05	75													
2-Jan-05	76				2.91%	0.33%	16.12%	80.64%	3.48%	96.52%	31.5	30.3		
3-Jan-05	77	11.823										30.4	39.2	
4-Jan-05	78													
5-Jan-05	79				3.79%	0.00%	10.12%	86.09%	4.22%	95.78%				
6-Jan-05	80													
7-Jan-05	81	0.000			4.49%	0.00%	12.86%	82.65%	5.15%	94.85%				
8-Jan-05	82													

					Gas Analysis						Ammonia Measurements (mg N/L)			
		Effluent			Percent Composition				CO2/CH4 % Ratio		influent		Effluent	
Date/Events	Day	USF Propionic	TORAY Acetic	TORAY Propionic	CO2	O2	N2	CH4	CO2	CH4	Infl. NA	Inf. A	USF	Toray
9-Jan-05	83										36.1	35.2		
10-Jan-05	84	0.000			2.94%	0.37%	11.61%	85.08%	3.34%	96.66%		34.9	41.7	
11-Jan-05	85													
12-Jan-05	86				3.10%	0.31%	10.30%	86.92%	3.46%	96.54%				
13-Jan-05	87													
14-Jan-05	88	0.000			3.16%	0.21%	9.53%	87.10%	3.50%	96.50%				
15-Jan-05	89													
16-Jan-05	90													
17-Jan-05	91	0.000			3.16%	0.93%	10.22%	85.69%	3.55%	96.45%			42.6	
18-Jan-05	92													
19-Jan-05	93	0.000												
20-Jan-05	94													
21-Jan-05	95				3.37%	0.20%	9.64%	86.79%	3.74%	96.26%				
22-Jan-05	96													
23-Jan-05	97										26.2	26		
24-Jan-05	98	0.000										26.5	29.2	
25-Jan-05	99													
26-Jan-05	100				3.03%	0.26%	10.86%	85.85%	3.41%	96.59%				
27-Jan-05	101													
28-Jan-05	102	0.000			3.28%	0.20%	10.54%	85.98%	3.67%	96.33%				
29-Jan-05	103													
30-Jan-05	104	4.943									30.1	30.8		
31-Jan-05	105				1.91%	0.27%	11.74%	86.07%	2.18%	97.82%		32.6	33.4	
1-Feb-05	106													
2-Feb-05	107	0.000												
3-Feb-05	108													
4-Feb-05	109	0.000			2.69%	0.30%	11.61%	85.39%	3.06%	96.94%				
5-Feb-05	110													
6-Feb-05	111										34.1	31		
7-Feb-05	112	0.000			2.94%	0.31%	11.18%	85.57%	3.33%	96.67%		34.3	41.5	
8-Feb-05	113													
9-Feb-05	114	0.000			2.58%	0.36%	11.69%	85.37%	2.94%	97.06%				
10-Feb-05	115													
11-Feb-05	116	0.000			2.43%	0.33%	11.66%	85.57%	2.77%	97.23%				
12-Feb-05	117													
13-Feb-05	118										32.6	32.8		
14-Feb-05	119	0.000			2.54%	0.00%	12.16%	85.30%	2.89%	97.11%		33.2	37	
15-Feb-05	120													
16-Feb-05	121	0.000												
17-Feb-05	122													
18-Feb-05	123	0.000			3.04%	0.11%	63.38%	33.48%	8.32%	91.68%				
19-Feb-05	124													
20-Feb-05	125										29.2	28.9		
21-Feb-05	126	0.000			2.70%	0.75%	31.92%	64.63%	4.01%	95.99%		29.1	32.4	
22-Feb-05	127													
22-Feb-05	127													
23-Feb-05	128	0.000			2.57%	0.36%	22.46%	74.61%	3.33%	96.67%				
24-Feb-05	129													
25-Feb-05	130				2.81%	0.23%	15.53%	81.43%	3.34%	96.66%				
26-Feb-05	131													
27-Feb-05	132										31	29.5		
28-Feb-05	133	0.000			2.59%	0.25%	12.39%	84.78%	2.96%	97.04%		31.7	34.1	
1-Mar-05	134													
2-Mar-05	135	0.000			2.63%	0.32%	12.67%	84.38%	3.03%	96.97%				
3-Mar-05	136	0.000												
4-Mar-05	137				3.06%	0.26%	11.89%	84.79%	3.48%	96.52%				
5-Mar-05	138													
6-Mar-05	139										29.6	28.6		
7-Mar-05	140	0.000										29.9	35.7	
8-Mar-05	141	0.000	0.000	0.000										
9-Mar-05	142													
10-Mar-05	143													
11-Mar-05	144	0.000	0.000	0.000	2.44%	0.36%	57.48%	39.72%	5.79%	94.21%				
12-Mar-05	145													
13-Mar-05	146													
14-Mar-05	147	0.000			3.23%	0.01%	29.82%	66.85%	4.61%	95.39%	32.2	27.4	40.7	41.7
15-Mar-05	148	0.000												
16-Mar-05	149	0.000			2.64%	0.91%	19.61%	76.83%	3.33%	96.67%				

					Gas Analysis						Ammonia Measurements (mg N/L)			
		Effluent			Percent Composition				CO2/CH4 % Ratio		influent		Effluent	
Date/Events	Day	USF Propionic	TORAY Acetic	TORAY Propionic	CO2	O2	N2	CH4	CO2	CH4	Infl. NA	Inf. A	USF	Toray
17-Mar-05	150													
18-Mar-05	151		0.000	0.000	2.62%	0.28%	14.39%	82.71%	3.08%	96.92%				
19-Mar-05	152													
20-Mar-05	153										33.2	32.6		
21-Mar-05	154	7.471	17.031	6.363	2.44%	0.32%	16.28%	80.96%	2.92%	97.08%		33.8	37.4	37.4
22-Mar-05	155													
23-Mar-05	156	4.842	17.980	4.800	2.34%	0.34%	14.40%	82.92%	2.74%	97.26%				
24-Mar-05	157													
25-Mar-05	158	6.713	24.944	7.048	2.60%	0.20%	13.52%	83.68%	3.01%	96.99%				
26-Mar-05	159													
27-Mar-05	160													
28-Mar-05	161	4.282	24.784	4.455	2.15%	0.37%	13.46%	84.01%	2.50%	97.50%				
29-Mar-05	162													
30-Mar-05	163	1.873	21.099	1.304	2.40%	0.33%	13.58%	83.69%	2.79%	97.21%				
31-Mar-05	164													
1-Apr-05	165	5.000	60.167	3.847	2.21%	0.29%	30.68%	66.82%	3.21%	96.79%				
2-Apr-05	166													
3-Apr-05	167										25.3	24.8		
4-Apr-05	168	4.934	76.316	4.498	2.29%	0.72%	25.87%	71.12%	3.12%	96.88%		25.1	30.7	30.6
5-Apr-05	169													
6-Apr-05	170	1.803	79.239	1.440	2.13%	0.33%	23.53%	74.01%	2.79%	97.21%				
6-Apr-05	171													
7-Apr-05	172													
8-Apr-05	173	1.131	85.278	1.358	1.72%	0.21%	66.30%	31.77%	5.13%	94.87%				
9-Apr-05	174													
10-Apr-05	175													
11-Apr-05	176	2.648	1.821	1.782										
11-Apr-05	176													
12-Apr-05	177													
13-Apr-05	178	3.150	3.160	4.617	1.61%	0.26%	81.00%	17.13%	8.60%	91.40%	24		29.4	28.6
14-Apr-05	179													
15-Apr-05	180	5.862	5.674	4.448	1.83%	0.38%	78.07%	19.72%	8.48%	91.52%				
16-Apr-05	181													
17-Apr-05	182	6.949	7.197	4.139	1.59%	4.79%	72.18%	21.45%	6.90%	93.10%				
18-Apr-05	183													
19-Apr-05	184													
20-Apr-05	185													
21-Apr-05	186	5.882	6.084	3.076	2.64%	1.12%	59.88%	36.37%	6.76%	93.24%				
22-Apr-05	187													
23-Apr-05	188													
24-Apr-05	189													
25-Apr-05	190													
25-Apr-05	190													
26-Apr-05	191	5.730	25.908	3.851	1.65%	0.00%	88.96%	9.36%	14.90%	85.10%	25.9		31.5	30.9
27-Apr-05	192													
28-Apr-05	193													
29-Apr-05	194	7.832	29.701	5.783	1.51%	2.87%	83.29%	12.33%	10.88%	89.12%				
30-Apr-05	195													
1-May-05	196													
2-May-05	197	10.866	34.894	9.681	2.26%	0.54%	73.08%	24.12%	8.55%	91.45%	29.3		32.9	33.9
3-May-05	198													
4-May-05	199	10.287	32.910	7.126	2.42%	0.00%	70.86%	26.72%	8.29%	91.71%				
4-May-05	199													
5-May-05	200													
6-May-05	201	7.870	36.084	8.103										
7-May-05	202													
8-May-05	203													
9-May-05	204	10.476	48.948	12.842	2.65%	0.45%	78.01%	18.89%	12.30%	87.70%	30.2		36.9	37.2
10-May-05	205													
11-May-05	206	7.207	44.653	11.184	2.56%	1.05%	74.22%	22.17%	10.35%	89.65%				
12-May-05	207													
12-May-05	207													

					Gas Analysis						Ammonia Measurements (mg N/L)			
		Effluent			Percent Composition				CO2/CH4 % Ratio		influent		Effluent	
Date/Events	Day	USF Propionic	TORAY Acetic	TORAY Propionic	CO2	O2	N2	CH4	CO2	CH4	Infl. NA	Inf. A	USF	Toray
13-May-05	208	8.212	48.848	8.054										
14-May-05	209													
15-May-05	210													
16-May-05	211	13.359	51.336	14.258							29.5		37.9	35.5
17-May-05	212													
18-May-05	213	8.196	36.881	8.098										
19-May-05	214													
20-May-05	215	8.510	32.942	8.879	1.33%	0.00%	83.87%	14.79%	8.27%	91.73%				
21-May-05	216													
22-May-05	217													
23-May-05	218													
24-May-05	219													
25-May-05	220	11.266	43.120	12.378	2.31%	1.72%	81.44%	14.53%	13.69%	86.31%	25.3		30.2	31.3
26-May-05	221													
27-May-05	222	13.660	55.540	13.390										
28-May-05	223													
29-May-05	224													
30-May-05	225	15.780	62.470	15.790							30.4		35.7	36.8
31-May-05	226				2.81%	0.00%	73.59%	23.60%	10.63%	89.37%				
31-May-05	226													
1-Jun-05	227	9.290	51.570	11.870										
2-Jun-05	228													
3-Jun-05	229	10.410	39.670	10.650										
4-Jun-05	230													
5-Jun-05	231													
6-Jun-05	232	12.080	47.830	11.890							27.3		33.3	34
7-Jun-05	233													
8-Jun-05	234	11.710	47.900	13.080										
9-Jun-05	235													
10-Jun-05	236	11.710	0.000	0.000	2.66%	0.00%	77.92%	19.43%	12.03%	87.97%				
11-Jun-05	237													
12-Jun-05	238													
13-Jun-05	239	12.370	57.460	11.390	2.32%	0.00%	75.37%	22.31%	9.42%	90.58%			32.7	33.5
14-Jun-05	240													
15-Jun-05	241	10.040	55.060	11.120										
16-Jun-05	242													
17-Jun-05	243	9.270	43.930	9.460	2.60%	0.00%	71.91%	25.49%	9.24%	90.76%				
18-Jun-05	244													
19-Jun-05	245													
20-Jun-05	246	9.280	44.560	8.560	2.83%	0.00%	70.47%	26.70%	9.58%	90.42%	26.8		32.4	30.6
21-Jun-05	247													
22-Jun-05	248				1.73%	6,78%	80.54%	10.96%	13.60%	86.40%				
23-Jun-05	249													
24-Jun-05	250				3.05%	0.00%	67.75%	29.20%	9.47%	90.53%				
25-Jun-05	251													
26-Jun-05	252													
27-Jun-05	253				2.99%	0.48%	62.72%	33.81%	8.12%	91.88%	24.7		29.5	30.4
28-Jun-05	254													
29-Jun-05	255				3.01%	0.54%	62.13%	34.32%	8.06%	91.94%				
30-Jun-05	256													
1-Jul-05	257				3.10%	0.00%	60.77%	36.13%	7.90%	92.10%				
2-Jul-05	258													
3-Jul-05	259													
4-Jul-05	260				3.11%	0.00%	60.11%	36.77%	7.81%	92.19%				

		Nitrate &Nitrite mg N/L)				Phosphate (mg P/L)			
		Influent		Effluent		Influent		Effluent	
Date/Events	Day	Inf.NA.	Inf. A	USF	Toray	Inf.NA.	Inf. A	USF	Toray
19-Oct-04	1								
20-Oct-04	2								
21-Oct-04	3								
22-Oct-04	4								
23-Oct-04	5								
24-Oct-04	6								
25-Oct-04	7								
26-Oct-04	8								
27-Oct-04	9								
28-Oct-04	10								
29-Oct-04	11								
30-Oct-04	12								
31-Oct-04	13								
1-Nov-04	14								
2-Nov-04	15								
3-Nov-04	16								
4-Nov-04	17								
5-Nov-04	18								
6-Nov-04	19								
7-Nov-04	20								
8-Nov-04	21								
9-Nov-04	22								
10-Nov-04	23								
11-Nov-04	24								
12-Nov-04	25								
13-Nov-04	26								
14-Nov-04	27								
15-Nov-04	28								
16-Nov-04	29								
17-Nov-04	30								
18-Nov-04	31								
19-Nov-04	32								
20-Nov-04	33								
21-Nov-04	34								
22-Nov-04	35								
24-Nov-04	37								
25-Nov-04	38								
26-Nov-04	39								
27-Nov-04	40								
28-Nov-04	41								
29-Nov-04	42								
30-Nov-04	43								
1-Dec-04	44								
2-Dec-04	45								
3-Dec-04	46								
4-Dec-04	47								
5-Dec-04	48								
6-Dec-04	49								
7-Dec-04	50								
8-Dec-04	51								
9-Dec-04	52								
10-Dec-04	53								
11-Dec-04	54								
12-Dec-04	55								
13-Dec-04	56								
14-Dec-04	57								
15-Dec-04	58								
16-Dec-04	59								
17-Dec-04	60								
18-Dec-04	61								
19-Dec-04	62								
20-Dec-04	63								
21-Dec-04	64								
22-Dec-04	65								
23-Dec-04	66								
24-Dec-04	67								
25-Dec-04	68								
26-Dec-04	69								
27-Dec-04	70								
28-Dec-04	71								
29-Dec-04	72								
30-Dec-04	73								
31-Dec-04	74								
1-Jan-05	75								
2-Jan-05	76	0.0172	0.0235			3.29	3.21		
3-Jan-05	77		0.045	0.0198			3.07	3.57	
4-Jan-05	78								
5-Jan-05	79								
6-Jan-05	80								
7-Jan-05	81								
8-Jan-05	82								

		Nitrate &Nitrite mg N/L)				Phosphate (mg P/L)			
		Influent		Effluent		Influent		Effluent	
Date/Events	Day	Inf.NA.	Inf. A	USF	Toray	Inf.NA.	Inf. A	USF	Toray
9-Jan-05	83	0.0518	0.0596			3.71	3.63		
10-Jan-05	84		0.00604	-0.000376			3.6	4.46	
11-Jan-05	85								
12-Jan-05	86								
13-Jan-05	87								
14-Jan-05	88								
15-Jan-05	89								
16-Jan-05	90								
17-Jan-05	91			0.00779				3.93	
18-Jan-05	92								
19-Jan-05	93								
20-Jan-05	94								
21-Jan-05	95								
22-Jan-05	96								
23-Jan-05	97	0.0351	0.0884			2.46	2.4		
24-Jan-05	98		0.047	0.014			2.44	3.03	
25-Jan-05	99								
26-Jan-05	100								
27-Jan-05	101								
28-Jan-05	102								
29-Jan-05	103								
30-Jan-05	104	0.0389	0.0931			3.29	3.05		
31-Jan-05	105		0.065	0.0111			3.56	3.81	
1-Feb-05	106								
2-Feb-05	107								
3-Feb-05	108								
4-Feb-05	109								
5-Feb-05	110								
6-Feb-05	111	0.0379	0.0945			3.26	3.34		
7-Feb-05	112		0.131	-0.00557			3.46	4.18	
8-Feb-05	113								
9-Feb-05	114								
10-Feb-05	115								
11-Feb-05	116								
12-Feb-05	117								
13-Feb-05	118	0.105	0.0794			2.88	3.03		
14-Feb-05	119		0.102	0.0535			3.3	3.94	
15-Feb-05	120								
16-Feb-05	121								
17-Feb-05	122								
18-Feb-05	123								
19-Feb-05	124								
20-Feb-05	125	0.0445	0.0396			2.81	2.71		
21-Feb-05	126		0.0562	0.00807			3.01	3.69	
22-Feb-05	127								
22-Feb-05	127								
23-Feb-05	128								
24-Feb-05	129								
25-Feb-05	130								
26-Feb-05	131								
27-Feb-05	132	0.0678	0.0801			3.02	2.84		
28-Feb-05	133		0.0756	0.0226			3.3	3.88	
1-Mar-05	134								
2-Mar-05	135								
3-Mar-05	136								
4-Mar-05	137								
5-Mar-05	138								
6-Mar-05	139	0.139	0.0271				2.73		
7-Mar-05	140		0.0409	0.00601			3.18	3.94	
8-Mar-05	141								
9-Mar-05	142								
10-Mar-05	143								
11-Mar-05	144								
12-Mar-05	145								
13-Mar-05	146								
14-Mar-05	147	0.0485	0.0377	-0.0134	0.0229	3.02	3.54	4.64	5.3
15-Mar-05	148								
16-Mar-05	149								

		Nitrate &Nitrite mg N/L)				Phosphate (mg P/L)			
		Influent		Effluent		Influent		Effluent	
Date/Events	Day	Inf.NA.	Inf. A	USF	Toray	Inf.NA.	Inf. A	USF	Toray
17-Mar-05	150								
18-Mar-05	151								
19-Mar-05	152								
20-Mar-05	153	0.0975	0.128			3.83	3.76		
21-Mar-05	154		0.115	0.112	0.0973		3.74	3.81	4.05
22-Mar-05	155								
23-Mar-05	156								
24-Mar-05	157								
25-Mar-05	158								
26-Mar-05	159								
27-Mar-05	160								
28-Mar-05	161								
29-Mar-05	162								
30-Mar-05	163								
31-Mar-05	164								
1-Apr-05	165								
2-Apr-05	166								
3-Apr-05	167	0.102	0.0595			2.56	2.68		
4-Apr-05	168		0.0608	0.0187	0.0434		2.74	2.87	3.67
5-Apr-05	169								
6-Apr-05	170								
6-Apr-05	171								
7-Apr-05	172								
8-Apr-05	173								
9-Apr-05	174								
10-Apr-05	175								
11-Apr-05	176								
11-Apr-05	176								
12-Apr-05	177								
13-Apr-05	178	0.0723		0.0308	0.0311	2.38		3.1	3.2
14-Apr-05	179								
15-Apr-05	180								
16-Apr-05	181								
17-Apr-05	182								
18-Apr-05	183								
19-Apr-05	184								
20-Apr-05	185								
21-Apr-05	186								
22-Apr-05	187								
23-Apr-05	188								
24-Apr-05	189								
25-Apr-05	190								
25-Apr-05	190								
26-Apr-05	191	0.105		0.0655	0.0703	2.58		3.34	3.28
27-Apr-05	192								
28-Apr-05	193								
29-Apr-05	194								
30-Apr-05	195								
1-May-05	196								
2-May-05	197	0.237		0.0768	0.0868	2.66		3.35	3.57
3-May-05	198								
4-May-05	199								
4-May-05	199								
5-May-05	200								
6-May-05	201								
7-May-05	202								
8-May-05	203								
9-May-05	204	0.0603		0.0456	0.0429	3.2		3.59	3.78
10-May-05	205								
11-May-05	206								
12-May-05	207								
12-May-05	207								

		Nitrate &Nitrite mg N/L)				Phosphate (mg P/L)			
		Influent		Effluent		Influent		Effluent	
Date/Events	Day	Inf.NA.	Inf. A	USF	Toray	Inf.NA.	Inf. A	USF	Toray
13-May-05	208								
14-May-05	209								
15-May-05	210								
16-May-05	211	0.0790		0.0189	0.00811	2.52		3.44	3.5
17-May-05	212								
18-May-05	213								
19-May-05	214								
20-May-05	215								
21-May-05	216								
22-May-05	217								
23-May-05	218								
24-May-05	219								
25-May-05	220	0.0449		0.0168	0.0177	2.72		3	3.41
26-May-05	221								
27-May-05	222								
28-May-05	223								
29-May-05	224								
30-May-05	225	0.3090		0.00796	0.00911	2.88		3.27	3.58
31-May-05	226								
31-May-05	226								
1-Jun-05	227								
2-Jun-05	228								
3-Jun-05	229								
4-Jun-05	230								
5-Jun-05	231								
6-Jun-05	232	0.132		0.0461	0.0887	3		2.89	3.13
7-Jun-05	233								
8-Jun-05	234								
9-Jun-05	235								
10-Jun-05	236								
11-Jun-05	237								
12-Jun-05	238								
13-Jun-05	239			0.05	0.0578			3.48	3.73
14-Jun-05	240								
15-Jun-05	241								
16-Jun-05	242								
17-Jun-05	243								
18-Jun-05	244								
19-Jun-05	245								
20-Jun-05	246	0.596		0.198	0.162	2.69		3.31	2.23
21-Jun-05	247								
22-Jun-05	248								
23-Jun-05	249								
24-Jun-05	250								
25-Jun-05	251								
26-Jun-05	252								
27-Jun-05	253	0.492		0.201	0.243	2.48		3.29	3.48
28-Jun-05	254								
29-Jun-05	255								
30-Jun-05	256								
1-Jul-05	257								
2-Jul-05	258								
3-Jul-05	259								
4-Jul-05	260								