Industrial Wastewater Treatment With Anaerobic Moving Bed Biofilm Reactor

by

Alessandro di Biase

A Thesis submitted to the Faculty of Graduate Studies of

The University of Manitoba

in partial fulfilment of the requirements of the degree of

MASTER OF SCIENCE

Department of Civil Engineering

University of Manitoba

Winnipeg

Abstract

The overall goal of the thesis was to develop and optimize the moving bed biofilm reactor technology under anaerobic conditions. The thesis work was divided into two different series of experiments. Hence, at first, the reactor start-up on synthetic substrate was evaluated and it was proven that the anaerobic moving bed biofilm reactor technology could successfully treat concentrated wastewater. Subsequently, a study on Fort Garry Brewery wastewater was conducted to optimize the process for a typical North American industrial wastewater. The aim was successfully achieved and a potential design to treat Fort Garry Brewery wastewater was developed. The anaerobic moving bed biofilm reactor was found to be capable in treating brewery wastewater with potential savings to the industry paying surcharges for discharging wastewater over the city sewer bylaw limits.

Acknowledgements

I would like to acknowledge my advisor, Professor Jan A. Oleszkiewicz for his support in my studies. His broad experience and knowledge in the environmental engineering field impressed me during an international exchange program period held at the University of Manitoba. His unique character and attitude motivated me to join his research group as an M.Sc. student. This continuous expression of trust allowed me to go through several research topics and complete my studies by acquiring knowledge in a different field of studies than my environmental biology background. The work completed and important life experiences gained in Canada would not have been possible without his sincere support and advice. I appreciate his understanding and empathy in these tough years of my life spent far away my homeland country. I had the luck to work with and learn from one of the most important professional engineers. This is the most precious thing I have ever had in my entire work carrier.

Fort Garry Brewery supported the research, with the excellent collaboration by the brew master Mr. Matt Wolff which is gratefully acknowledged. The media was provided by Headworks BIO, Victoria BC, Canada.

My decision to join the master program at the University of Manitoba was driven mostly by my family support, the fortune to have Professor Jan A. Oleszkiewicz as an advisor and the important friendship developed with Tanner Devlin. If it was not for his help and positive attitude, my experience in Canada would have been more difficult. All my life issues and work challenges were possible to overcome thanks to the immense aid that was brought to me from him and his family. This unexpected behaviour is rare and it has a precious value to me. My family back in Florence played a major role in my staying here in Canada. Particularly, my mom, Francesca, has always believe in me and fully supported all my decisions. These three years

abroad were not easy at all for my family and me. I had experienced a very tough period and several life challenges during my stay. Despite difficulties encountered during these years, my mom and my brother, Riccardo, never ever asked me to return when important family issues occurred that required my presence. A few times, I was discouraged and felt bad not being able to go back to sustain and be with them in delicate moments but the will in continuing my work and completing what I started never waivered. Although all the problems and peculiarity of situation, my family always supported me and gave me the power to keep going. Without my brother and mom positive motivation, the time spent overseas would have been unmanageable as these years were the toughest years in my whole life. In addition, a special thought goes to Alessio Gatteschi that brought stability and tranquillity to my family in the past few years. My relatives, Daniela, zio Enzo e zia Teresa have also contribute to my journey and there aid was essential in the achievement of my work and studies abroad. I would also like to thank a family friend, Laura Scarpellini, who helped make the international experience happen. Without her support, it would have been very difficult or probably impossible for me to make it through the first year of the international exchange program. In addition, the hospitality of Virginia Draghi and Jeremy Ranson was fundamental and appreciated for the first weeks after my arrival in Canada.

I would like to acknowledge Dr. Qiuyan Yuan for letting me be a teaching assistant for the environmental engineering design course for two semesters and giving me the experience of being “on the other side of the desk”. I would like to acknowledge Dr. Richard Sparling for letting me run GC samples in his laboratory with the skilled aid of Monika Niemczyk. Also the help of GC measurement carried out by Warren Blunt during the first part of my research thesis, as well as the precious HPLC analysis on VFA made by Elsie Jordan. The lab manager Victor

Wei, for his precise and meticulous supplies management and help in reactors setup. Finally, I would like to thank Charlene Urfano for her thesis review and English editing.

Table of Contents

1. CHAPTER 1 - INTRODUCTION 1

1.1 Anaerobic Digestion process 3

1.1.1 Hydrolysis 4

1.1.2 Acidogenesis 5

1.1.3 Acetogenesis 6

1.1.4 Methanogenesis 7

1.1.5 Biogas production 9

1.2 High rate digesters for industrial wastewater treatment 10

1.2.1 Fixed film reactor 11

1.2.2 Up-flow anaerobic sludge blanket reactor 13

1.2.3 Anaerobic fluidized bed reactor 16

1.2.4 Anaerobic baffled reactor 18

1.2.5 Anaerobic hybrid reactor 20

1.2.6 Anaerobic moving bed biofilm reactor 22

1.3 Brewery Industry 30

1.4 Objectives 32

2. CHAPTER 2 – MATERIAL AND METHODS 33

2.1 Reactors setup 33

2.1.1 Start-up on synthetic and transition to brewery wastewater 33

2.1.2 Organic Loading Rates and HRT impact 35

2.2 Analytical methods 36

2.2.1 Performance measurements and equations 36

2.3 Kinetics tests 38

2.3.1 Start-up on synthetic and transition to brewery wastewater 38

2.3.2 Organic Loading Rates and HRT impact 39
2.4 Brewery characterization 39
3. CHAPTER 3 – RESULTS AND DISCUSSION 41
3.1 Start-up and transition to brewery wastewater 41
3.1.1 Physico-chemical Parameters 41
3.1.2 Removal Performances 41
3.1.3 Volatile Fatty Acid 44
3.1.4 Biogas production and characterization 48
3.2 Organic loading rate increasing via decreased HRT 50
3.2.1 Influent and effluent characteristics 50
3.2.2 Biomass activity 54
3.2.3 Biogas production and characterization 58
4. CHAPTER 4 – ENGINEERING SIGNIFICANCE 60
4.1 Design considerations 60
4.2 Case Study – Fort Garry Brewery 67
SUMMARY AND CONCLUSIONS 69
REFERENCE 71
APPENDIX 84
6.1 Start-up and transition to brewery wastewater 84
6.1.1 pH, alkalinity, influent and effluent COD, and OLR 84
6.1.2 VFA raw data in mg/L 88
6.1.3 Total gas production and composition 89
6.2 Appendix B – HRT study 92
6.2.1 Influent and effluent solids, total and soluble COD, Alkalinity and pH 92
6.2.2 Total and soluble COD removal, OLR and SSAA 101

6.2.3 Kinetic tests 106
6.2.4 Total biogas and methane production, and biogas composition 109
6.2.5 COD conversion into methane 113
6.3 Engineering significance 115
6.3.1 OLR and SALR and sCOD removal at different HRT 115
6.3.2 sCOD loadings, reactor and media volume requirement at different fillings and percentage removal 116
6.3.3 Fort Garry design parameters using AC920 118

List of Figures

Figure 1-1 Anaerobic digestion process (Developed from Metcalf & Eddy, 2013) 3

Figure 1-2 – Fixed film up-flow reactor 11

Figure 1-3 – Up-flow anaerobic sludge blanket reactor 14

Figure 1-4 – Anaerobic fluidized bed reactor 17

Figure 1-5 – Simplified baffle reactors cross section (single gas headspace) 19

Figure 1-6 Anaerobic hybrid reactor 21

Figure 1-7 – Anaerobic moving bed biofilm reactor 23

Figure 2-1 – Start-up reactors setup 34

Figure 2-2 - Reactor setup of the HRT and OLR study 35

Figure 3-1 - Type of substrate and OLR supplied shown with percent removal in the 25% and 35% media reactors 43

Figure 3-2 – Total influent VFA and effluent VFA speciation of the 25 % reactor 46

Figure 3-3 – Influent total VFA and effluent VFA speciation of the 35 % reactor 46

Figure 3-4 – OLR and VFA:Alk ratio correlations of both 25 % and 35 % reactors with (on top) and without (on the bottom) alkalinity addition 47

Figure 3-5 – Biogas composition on synthetic and brewery substrate in 25 % and 35 % reactors 48

Figure 3-6 - Total biogas and methane production related to the OLR 49

Figure 3-7 - Averaged soluble COD removal shown with averaged total OLR during the study. The different intervals of HRT are presented as blocks at the top of each graph. 51

Figure 3-8 - Specific surface area activity (SSAA) measured on soluble COD at different HRTs 55

Figure 3-9 - Averaged total COD removal compared with removal by detached suspended biomass alone, at different HRT 56

Figure 3-10 - Total organic loading rate (OLR) and specific surface area activity (SSAA) obtained at different HRT. Percentages represent the load of COD removed relative to the oncoming load 57

Figure 3-11 - Biogas composition as percent methane and carbon dioxide (on top). Methane and carbon dioxide ratio (on the bottom). Error bars represent one standard deviation between triplicates 59

Figure 4-1 – Surface area loading rate and percentage removal of sCOD correlation from experimental data 62

Figure 4-2 – Comparison of reactor volume requirement to achieve 80% removal by implementation of three media (i.e., AC450, AC515 and AC920) 65

Figure 4-3 – Comparison of reactor volume requirement to achieve 90% removal by implementation of three media (i.e., AC450, AC515 and AC920) 66

List of Tables

Table 1-1 Fermentation reactions examples 5

Table 1-2 Acetogenic reactions examples 7

Table 1-3 – Different types of methane-forming microorganism 8

Table 1-4 – Fixed film reactors treating industrial wastewaters 12

Table 1-5 – UASB reactors treating different industrial wastewaters 13

Table 1-6 – AFR treating different industrial wastewaters 18

Table 1-7 – Anaerobic baffled reactor treating different industrial wastewaters 19

Table 1-8 – Anaerobic hybrid reactor treating different wastewaters 21

Table 1-9 – Different type of media and specific surface area 26

Table 1-10 – AMBBR treating industrial wastewater 29

Table 1-11 – Typical brewery wastewater characteristics (Chastain et al., 2011) 31

Table 1-12 – Requirement from City of Winnipeg sewer by-law No.92/2010 31

Table 2-1 – Micronutrient concentrations 34

Table 2-2 - Brewery wastewater characterization 40

Table 3-1 – AMMBR influent and effluent characteristics for the entire study 50

Table 3-2 – Review of AMBBR processes treating industrial wastewater 53

Table 4-1 – Beer production and wastewater generation rates 60

Table 4-2 – Mass of COD produced in different breweries size 61

Table 4-3 – Surface area loading rates at particular sCOD percentage removal 62

Table 4-4 – Total surface area required to achieve COD 80 and 90 % removals 63

Table 4-5 – Media volume (m^3) required for the three types of media selected 64

Table 4-6 – Design parameters for AC920 media at 70% fillings 68

Abbreviations

AD	Anaerobic Digestion
CHP	Combined Heat and Power
UASB	Up-flow Anaerobic Sludge Blanket
AMBBR	Anaerobic Moving Bed Biofilm Reactor
VFA	Volatile Fatty Acid
BOD	Biological Oxygen Demand
COD	Chemical Oxygen Demand
EGSB	Expanded Granular Sludge Bed
OLR	Organic Loading Rate
HRT	Hydraulic Retention Time
SSA	Specific Surface Area
SRT	Sludge Retention Time
AFR	Anaerobic Fluidized-bed Reactor
IFBR	Inverse Fluidized Bed Reactor
ABR	Anaerobic Baffled Reactor
AHR	Anaerobic Hybrid Reactor
AF-UASB	Anaerobic Filter – Up-flow Anaerobic Sludge Blanket
UASFF	Up-flow Anaerobic Sludge Bed Fixed Film
MBBR	Moving Bed Biofilm Reactor
SSAA	Specific Surface Area Activity
TSS	Total Suspended Solids
VSS	Volatile Suspended Solids
SALR	Surface Area Loading Rate
TSA	Total Surface Area

CHAPTER 1 - INTRODUCTION

The anaerobic digestion (AD) process converts organic matter into biogas with the aid of several microorganisms belonging to bacteria and archaea kingdoms. The anaerobic digestion involves the biological organic matter breakdown into mainly methane and carbon dioxide gas in an environment without oxygen. Anaerobic digestion occurs naturally in the environment and it is widely used to process any carbon-containing material, including kitchen and paper wastes, sewage sludge wastes, yard trimmings, and solid waste (Hidaka, Wang, & Tsumori, 2015; Meyer & Edwards, 2014; Xu, Wang, Lin, & Li, 2016). The discovery of anaerobic microorganisms occurred in 1861 when Pasteur, while studying fermentative reactions by organic matter decay, produced flammable gases used to heat water. The appearance of this process in wastewater treatment arose in the nineteenth century. The first full scale application was implemented in the 1890s by the city of Exeter in United Kingdom which adopted AD in wastewater treatment. Since then, it continues to be widely used as a mean for sludge sewage stabilization (Mahony et al., 2002; Oleszkiewicz & Mavinic, 2001).

The microbiological pathways and biochemistry behind the anaerobic reactions are not fully understood, but within the last 40 years, several researchers have shown a broad outline of the processes involved. In the 1930s, the development of the microbiology field allowed the study and identification of the mechanism of biogas production and its link with anaerobic microorganisms metabolism (Gerardi, 2003; Lusk, 1999). From there on, the scientific community looked at the optimization of the environmental conditions to enhance biogas production. Anaerobic digestion produces biogas suitable to heating reuse for digesters themselves or into combined heat and power (CHP) systems for energy production. Hence, AD processes have been improved over the past few decades and the biogas production and its

potential reuse gained attention for worldwide application. The interest in understanding metabolic pathways and type of microorganisms involved in the complexity of the AD process had been shown by recent studies on metagenomics sequencing. Guo et al. (2015) had determine the profile at Domain level in a full-scale anaerobic reactor digesting wasted activated sludge. The authors found that the Bacteria was the dominant Domain (93 %) while Archaea accounted for 5.6 % of the total DNA sequences analysed and 1.1 % were Eukaryota. Other authors, studying two different digester samples, have investigated the abundancy in terms of phyla of the Bacteria community (Yang et al., 2014). *Proteobacteria*, *Bacteroidetes* and *Firmicutes* were the prevalent phyla quantified in AD sludge.

In the middle of the twentieth century, the benefits associated to the biogas production and its use intensified the studies on the improvement of the conventional digesters into more suitable designs. At first, the AD studies were mainly focused on farm application where manure was digested to produce heat and later electricity (Bacenetti, Negri, Fiala, & González-García, 2013; Monteiro, Mantha, & Rouboa, 2011; Vlyssides, Mai, & Barampouti, 2015). As knowledge increased, AD was implemented in treating municipal and industrial wastewater, as well as municipal solids waste. Recently, the interest in AD is mainly focused on treating industrial wastes, municipal wastewater and solids wastes. In Europe, Australia, North America and Asia, AD has become a typical option for treating the organic fractions of industrial waste with applications ranging from breweries to paper mills and food processing wastewater (Rajeshwari, Balakrishnan, Kansal, Lata, & Kishore, 2000). The pressure on companies to achieve more stringent wastewater effluent limits increased the attention of researchers to develop design configurations that will optimize both carbon removal efficiency and biogas production. Therefore, the development of the so called “second generation digester” or “high-rate anaerobic

reactors" began (Rajeshwari et al., 2000). In particular, the implementation of these technologies to the industrial sector depends on associated capital costs and a quick return of investments. The development of second-generation digester designs such as the up-flow anaerobic sludge blanket reactor (UASB), anaerobic fixed film reactor, and anaerobic moving bed biofilm reactor (AMBBR) reflected the need to shorten the hydraulic retention times (HRT) and therefore increase the loading rates in order to have the smaller footprint as possible. Therefore, achieving compact reactors that have maximized biogas production properties is the key factor to make one technology more appealing over another for full-scale industrial applications.

1.1 Anaerobic Digestion process

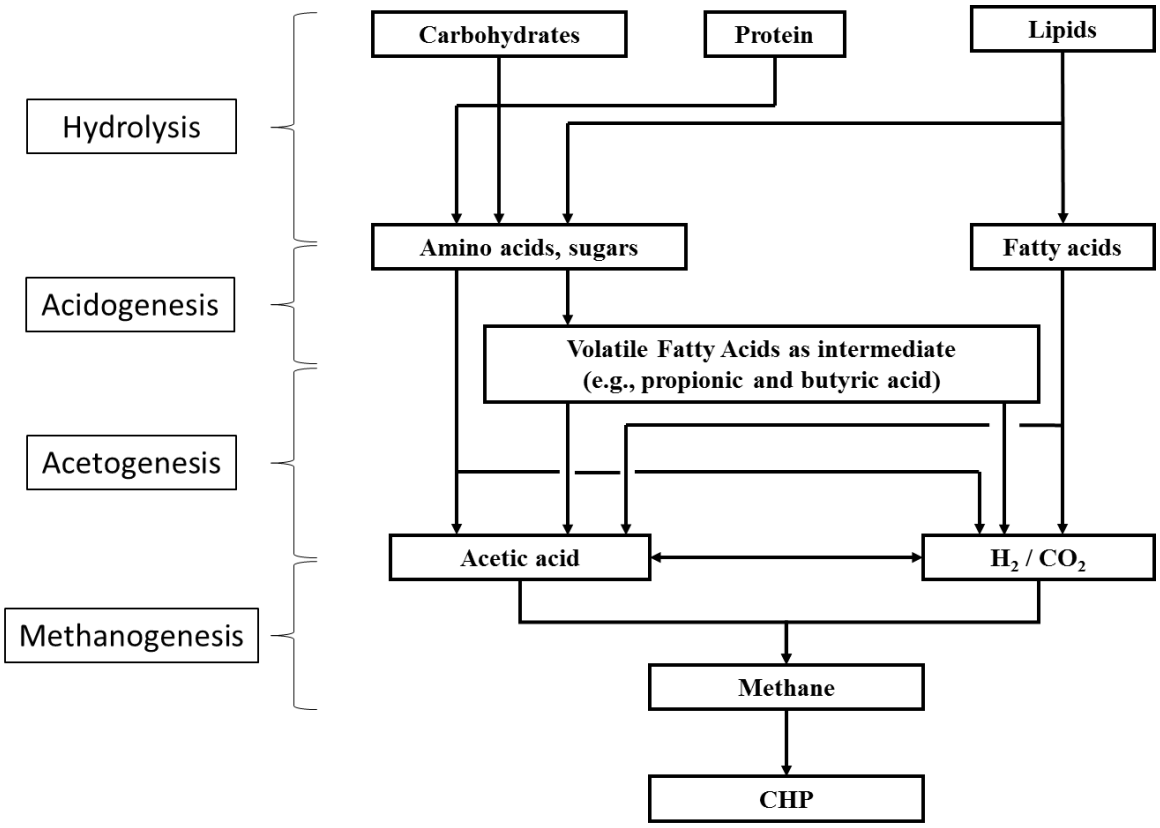


Figure 1-1 Anaerobic digestion process (Developed from Metcalf & Eddy, 2013)

Although the AD process is described and can be considered to take place in four stages (Figure 1-1), the processes occur simultaneously and synergistically.

Dairies, wineries, breweries and food processing industries generate wastewaters that must be treated prior discharge to sewer systems. In the absence of oxygen, microorganisms such as fermentative bacteria, hydrolyze organic matter into simpler organics that subsequently can be available as substrate. The broken down end products are, therefore, available for other organisms either for direct use by microorganisms belonging to the Domain Archaea (e.g., methanogens) or indirectly through conversion of the fermentation end products to substrates the methanogens can use (these organisms are called syntrophs). The methanogens convert the end product (i.e., acetate, hydrogen and carbon dioxide gases) into biogas mostly composed of carbon dioxide and methane. Anaerobic digestion is the process that involves these reaction chains driven by a syntrophic interaction between bacteria and archaea. The anaerobic digestion process can be described in four phases – Figure 1-1.

1.1.1 Hydrolysis

Hydrolysis is the process in which complex organic molecules such as protein, carbohydrates and lipids are broken down into soluble monomers (amino acids, fatty acids, glucose). A particular type of bacteria (i.e., hydrolytic fermentative) is responsible for the formations of the monomers which metabolize the organic material into a product available to and suitable for other microorganisms. The hydrolysis and initial fermentation is led by various bacteria genera; most of them are obligate anaerobes. *Clostridium* is the major group within the obligate anaerobes bacteria along with *Bacteroides* (Cai et al., 2016). Enzymes such as amylase, cellulase, protease, nuclease, and lipase are excreted from the bacteria and catalyze the hydrolysis (Maunoir, Sabil, Rambaud, Philip, & Coletti-Previero, 1991). Complex feedstock

such as cellulosic waste containing lignin will slow down the hydrolytic phase and therefore a longer contact time for hydrolysis to occur will be required. Other carbohydrates such as starch, hemicellulose and chitin, however, are known to be more rapidly converted to simple sugars via hydrolysis. These compounds subsequently are fermented to volatile fatty acids (VFA; Mata-Alvarez 2000).

Presented in Table 1-1 are examples of fermentation reactions of glucose into various VFA products (i.e., lactate, butyrate, and propionate and acetate).

Table 1-1 Fermentation reactions examples

Product	Reaction	$\Delta G^{0'}$(kJ/mole)
Lactate	$C_6H_{12}O_6 \rightleftharpoons 2\ CH_3CH(OH)COO^- + 2\ H^+$	-198.1
Butyrate	$C_6H_{12}O_6 + 2H_2O \rightleftharpoons CH_32(CH_2)COO^- + 2\ HCO_3^- + 2\ H_2 + 3H^+$	-254.4
Propionate + Acetate	$1.5\ C_6H_{12}O_6 \rightleftharpoons 2\ CH_3CH_2COO^- + CH_3COO^- + HCO_3^- + 4H^+$	-109.9

Lipids are hydrolyzed to glycerol and long-chain fatty acids. While glycerol can move across the cell wall, the fatty acid molecules are too large to be assimilated by the bacterial cell. The acids dissolve in the lipopolysaccharide on the bacteria surface are than metabolized by β-oxidation. The end products of hydrolysis are generally used as carbon and energy sources by acidogenic bacteria, which carry out the first stage fermentation.

1.1.2 Acidogenesis

Following the hydrolysis is the acid-forming phase: the Acidogenesis. Acidogenic bacteria transform the soluble monomer into simple organic compounds, mainly short chain VFA, ketones and alcohols (e.g., propionic, lactic, butyric acids, ethanol, glycerol, and acetone). This step is completed by a variety of bacteria operating in an anaerobic environment (Bertin et al., 2010). If the digestion were to stop at this stage, the accumulation of acids would lower the pH

and would inhibit the methanogenesis process (Li, Park, & Zhu, 2011). The concentrations of organics compound produced in this phase depend on the type of bacteria as well as with culture conditions, such as temperature and pH. Studies conducted by Wang et al. (2014) reviled that pH 6.0 is the optimum for the formation of VFA in AD of food waste.

1.1.3 Acetogenesis

The acetogenesis phase is often considered, with the acidogenesis phase, as acetate forming stage. During acetate production from more complex VFA, portion of carbon it is lost as carbon dioxide along with hydrogen gas (Fig. 1-1). Therefore, biological oxygen demand (BOD) and chemical oxygen demand (COD) are reduced throughout this step. Acetogenesis relies on carbohydrate fermentation and long chain fatty acids formed from the hydrolysis, which by this process are oxidized to acetate or propionate. The end products are a combination of acetate, carbon dioxide and hydrogen gases. The hydrogen as an intermediary is critical in AD reactions. The hydrogen partial pressure need to be low enough to thermodynamically allow the reaction to continue. The hydrogen-scavenging methanogens lower the concentration of hydrogen partial pressure by consuming hydrogen present in the digester and therefore ensure the conversion of acids occurs. As a result, the concentration of hydrogen measured by partial pressure is an indicator of the health of a digester (Mata-Alvarez, 2000).

Table 1-2 reports reaction examples driven by the metabolism of the acetogenic bacteria in converting the intermediate VFA (e.g., propionic and butyric acids) to acetate plus hydrogen.

In the reactions presented above the free energy value of the reaction that converts propionate to acetate is + 76.1 kJ, making it thermodynamically impractical. However, when methanogens and hydrogenotrophic microorganism utilize hydrogen gas, the free energy becomes negative (Table 1-2) and the low partial pressure of hydrogen allows the reaction to proceed. The transformation

of the organic material causes a decrease in the pH of the system by hydrogen gas production from fermentation reactions (Table 1-1). However, the maintenance of a proper methanogenic and hydrogenotrophic microorganisms community ensure avoiding drop down in pH which might compromising the AD process.

Table 1-2 Acetogenic reactions examples

Substrate	Reaction	$\Delta G^{0'}$ (kJ/mole)
Lactate	$CH_3CH(OH)COO^- + 2\ H_2O \rightleftharpoons CH_3COO^- + HCO_3^- + 2H_2 + H^+$	-3.96
Butyrate	$(CH_3)_2(CH_2)COO^- + 2\ H_2O \rightleftharpoons 2\ CH_3COO^- + 2\ H_2 + H^+$	+48.1
Propionate	$CH_3CH_2COO^- + 3\ H_2O \rightleftharpoons CH_3COO^- + HCO_3^- + 3\ H_2 + H^+$	+76.1

1.1.4 Methanogenesis

The methanogenic anaerobic organisms consist exclusively of strict anaerobes archaea belonging to the Euryarchaeota phylum (Rosenzweig & Ragsdale, 2011). These microorganisms lead the fourth stage of AD, known as methanogenesis or methane fermentation, in which the soluble organic matter will be converted into methane. These microorganisms are naturally present within the sediments or in the rumen of herbivores (Morgavi, Martin, Jouany, & Ranilla, 2012). Methane-forming microorganisms are a broad, oxygen-sensitive cluster of Archaea with different structures, enzymes, substrate utilizations rates, and temperature range of growth. Therefore, methane-forming microorganism can grow in an aerobic environment with doubling time ranging from 3 days at 35°C to 50 days at 10°C. Methane-forming bacteria are divided into different clusters based on the substrate utilization – Table 1-3.

The hydrogenotrophic methanogens (i.e., *Methanobacteriales, Methanococcales Methanomicrobiales, Methanopyrales*) produce methane and water molecules while metabolizing the carbon dioxide and hydrogen (Madigan, Martinko, Stahl, & Clark, 2012;

Ozuolmez et al., 2015). They contribute to reducing the hydrogen concentration pressure and therefore maintain the pH at neutral values. This lead to acetic acid formation avoiding accumulation of other VFA. Acetic acid can be used by the acetoclastic methanogens (i.e., *Methanosarcina, Methanosaeta*) to produce methane and carbon dioxide. Other compounds such as formate and methylamines can be metabolized and convert into methane by different methanogens not able to use acetate as substrate or capable in oxidize methanol (Madigan et al., 2012). About two thirds of methane is derived from acetate conversion or fermentation of an alcohol, such as methyl alcohol, and one third is the results of carbon dioxide reduction by hydrogen.

Table 1-3 – Different types of methane-forming microorganism

Substrate	Substrate utilization reaction	Archaea
Hydrogen	$CO_2 + 4\ H_2 + H^+ \rightarrow CH_4 + 3\ H_2O$	*Methanobacteriales, Methanococcales, Methanomicrobiales, Methanopyrales*
Acetic acid	$CH_3COOH \rightarrow CH_4 + CO_2$	*Methanosarcinales*
Formate	$4\ HCOO^- + 4\ H^+ \rightarrow CH_4 + 3\ CO_2 + 2\ H_2O$	*Methanobacteriales, Methanomicrobiales, Methanococcales*
Methanol	$4\ CH_3OH \rightarrow 3\ CH_4 + CO_2 + 2\ H_2O$	*Methanosarcinales*
Methylamines	$4\ (CH_3)_3N + H_2O \rightarrow 9\ CH_4 + 3\ CO_2 + 6\ H_2O + 4\ NH_3$	*Methanosarcinales*

The consistency of the digestion depends on proper balance of biological processes involved. Sudden loading of organic matter or a quick change in temperature will result on a buildup of organic acids and the pH will drop, facilitating the acidogens and eventually inhibiting the methanogenic process, which has an optimal pH between 6.5 and 7.5 (Rodrigues et al., 2014).

1.1.5 Biogas production

Biogas is made up mostly of methane and carbon dioxide, with trace amount of other gases (e.g., N_2, H_2S). The percentage of methane is typically between 50 and 70 %, which is the most valuable product of the process. To reuse the gas as an electricity source, it is fundamental to remove water vapor, and hydrogen sulfide, and nitrogen gases. The resulting gas, a mixture of methane and carbon dioxide, can be used directly as fuel in CHP systems, which are designed to operate with low heating value gases.

Biogas produced from AD typically contains 65% methane on average with a heating value of approximately 22,400 kJ/m^3. Methane gas at standard temperature and pressure (20°C, 1 atm) has a heating value of 35,800 kJ/m^3, while a natural gas mixture of methane, propane, and butane, has a heating value of 37,300 kJ/m^3 (Metcalf & Eddy, 2013). Ertem (2011) reports an energy content of 6.0-6.5 kWh/m^3, which is equivalent to 0.60-0.65 L oil/m^3 of biogas.

It is possible to estimate the specific methane production per unit of COD removed. Under anaerobic conditions at 35°C the COD conversion into methane gas is equal to 0.395 L CH_4/g COD since 25.29 L of methane equals 1 mole of gas at 35°C at 1 atm and the COD of one mole of methane is equal to 64 g COD/mole CH_4 (Metcalf & Eddy, 2013; Moody, Burns, Wu-Haan, & Spajić, 2009).

In smaller facilities, the biogas produced may be used to generate enough heat to maintain the temperature (i.e., 35°C) within the digester itself or likely save costs for digester heating if feasibility studies reveal it as practicable. On the other hand, larger facilities typically use the CHP systems to heat the reactor, as well as generate electricity to be used within the facility (Ferreira, Marques, & Malico, 2012; Shen, Linville, Urgun-Demirtas, Mintz, & Snyder, 2015). Besides lowering the electricity expenses, there are several advantages to generating electricity.

For instance, producing biogas on-site might lead to tax credits for renewable energy and carbon credits where available.

1.2 High rate digesters for industrial wastewater treatment

Considerable attention has been paid towards the development of high rate anaerobic digester for industrial wastewater treatment and their organic matter conversion into biogas optimization. The main benefit are that these reactors design can handle high loading rate and HRT in the order of hours making it possible to consistently reduce the reactor footprint. However, the need for long sludge retention time (SRT) to maintain slow growing microorganisms such as the methanogens biomass to be grown on carriers in a biofilm structure. Hence, all modern high rate anaerobic digesters designs are based on the concept of retaining high viable biomass by several means of bacterial sludge immobilization (e.g., carriers, packing material).

One of the following methods could be effective to achieve this need (Pol & Lettinga, 1986):

- Highly settleable sludge aggregates such as granules and/or flocs (e.g. up-flow anaerobic sludge blanket reactor and anaerobic baffled reactor),
- Biofilm bacteria development on high-density particulate carrier materials (e.g. fluidized bed reactors and anaerobic expanded bed reactors), or
- Biofilm development on packing materials (e.g. down-flow/up-flow anaerobic filter and anaerobic moving bed biofilm reactor).

Typical granular sludge systems are the up-flow anaerobic sludge bed (UASB) and the expanded granular sludge bed (EGSB) reactors (Latif et al. 2011; Ratanatamskul and Siritiewsri 2014; Seghezzo et al. 1998). Biofilm systems include the anaerobic fixed film, anaerobic fluidized-bed, and expanded-bed as well as AMBBR.

1.2.1 Fixed film reactor

Figure 1-2 reports a type of fixed film up-flow reactor. The support for biofilm development is usually activated carbon, polyvinyl chloride, hard rock particles or ceramic rings. The influent wastewater can be pumped from either above or below depending if it is an up-flow or down-flow reactor. The headspace at the top of the reactor allows the biogas to leave the solution and be collected into a biogas outlet.

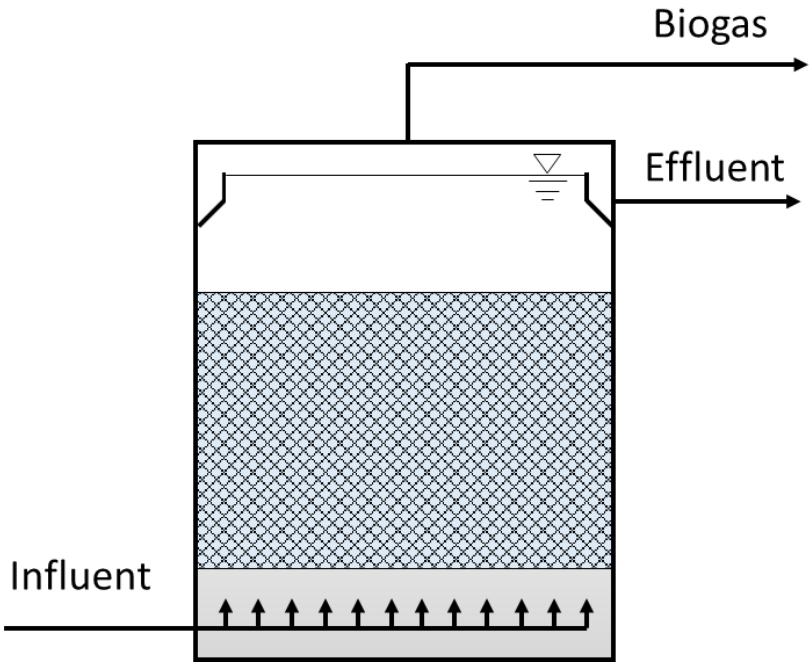


Figure 1-2 – Fixed film up-flow reactor

Fixed film reactors have the advantage of being both constructed and operated simply. In addition, it does not require mechanical mixing, which could save costs but has the constraint of being clogged by increasing in biofilm thickness and/or high influent total suspended solids (TSS) and soluble suspended solids (VSS) concentrations. However, fixed film processes have been proven to have stability at higher loading rates, and the capability to withstand a wide spectrum of toxic compounds and organic shock loads (Gangagni Rao et al., 2005; van den Berg,

Kennedy, & Samson, 1985). In addition, Van den Berg et al. (1985) has shown that this process can recover very quickly after starvation periods. However, the main limitation of this design is the high reactor volume requirement compared to others second-generation designs, due to the lower surface area of the packing material. Table 1-4 shows several studies of the capability of the fixed film reactor treating different wastewater streams.

Table 1-4 – Fixed film reactors treating industrial wastewaters

Wastewater treated	Packing material	OLR (kg COD/m^3 d)	COD_{rem} (%)	HRT (d)	Methane Yield (m^3CH_4/kg COD_{rem})	References
Distillery	Charcoal, coconut coir and nylon	23.3	64	8	0.31	Acharya et al. (2008)
Bulk drug industry	PVC rings	10	60 - 70	1.7	0.25	Gangagni Rao et al. (2005)
Distillery	Charcoal	8 - 19	80 - 51	8	-	Acharya et al. (2011)
Sugary	PVC pipe pieces	3.0	82	0.5	-	Pradeep et al. (2014)

Studies conducted on distillery wastewater by Acharya et al. (2008), show that fixed film reactors tested with different packing materials (i.e., charcoal, coconut coir and nylon fibers) were capable to achieve 64 % COD removal with an 8 d HRT and a maximum organic loading rate (OLR) of 23 kg COD/m^3 d. In addition, different OLRs were tested (i.e., 8 and 19 kg COD/m^3 d) at HRT ranging between 8 and 20 days with an overall 50 – 80 % COD removal and maximum averaged production of 27 m^3/d (Acharya et al., 2011). Recycled polyvinyl chloride (PVC) pipe pieces, however, were shown to be efficient for biofilm development and treatment of industrial sugar cane processing wastewater with 82 % COD removal at 0.5 d HRT and OLR ranging from 2.5 to 3.0 kg COD/m^3 d (Pradeep et al., 2014). Similar PVC pipe support material

with specific surface area equal to 224 m^2/m^3 was also shown to have high removal rates (i.e., 60 – 70 % COD) at an optimum OLR of 10 kg COD/m^3 d (Gangagni Rao et al., 2005).

1.2.2 Up-flow anaerobic sludge blanket reactor

Up-flow anaerobic sludge blanket reactor technology is being widely used to treat wastewater streams from different industrial processes such as distilleries, food-processing units and tanneries – Table 1-5.

Table 1-5 – UASB reactors treating different industrial wastewaters

Wastewater treated	OLR (kg COD/m^3d)	COD_{rem} (%)	HRT (h)	Methane Yield (m^3CH_4/kg COD_{rem})	References
Dairy	28	> 90	48	-	Kalyuzhnyi et al. (1997)
Sugary	4	> 90	43	0.34	Mijaylova Nacheva et al. (2009)
	24	80	7	0.35	
Textile	6	> 90	24	-	Somasiri et al. (2008)
Brewery	3	85	12	0.34	Cronin and Lo (1998)

The biomass in the UASB is in the form of granular sludge and high SRTs are achieved by direct settling. Therefore maintaining slow growing microorganisms such as methanogens. The main advantage of this technology compared to anaerobic filter or fluidized bed systems, is the lower capital costs requirements due to its small footprint. However, longer start-up periods are required to achieve sufficient stability and amount of sludge into granular form compared to attached growth systems. Granulation is sensitive to several factors such as wastewater composition, food to microorganisms ratio, HRT, and type of selection pressure. A study conducted by Oleszkiewicz and Romanek (1989) showed that a deficiency in ferric ions played a significant role in the granule's stability. Recent studies conducted by Lu et al. (2015) elucidate the factors affecting long term instability of granular sludge in a lab-scale UASB reactor.

Degranulation was found to be due to insufficient cation supply (e.g., calcium) which led to insufficient capacity of granular core formation and disequilibrium of extracellular polymeric substances ratios.

The principle behind a UASB reactor is essentially a solid/gas separation retaining the anaerobic sludge within the reactor. Figure 1-3 reports a typical design of an UASB reactor.

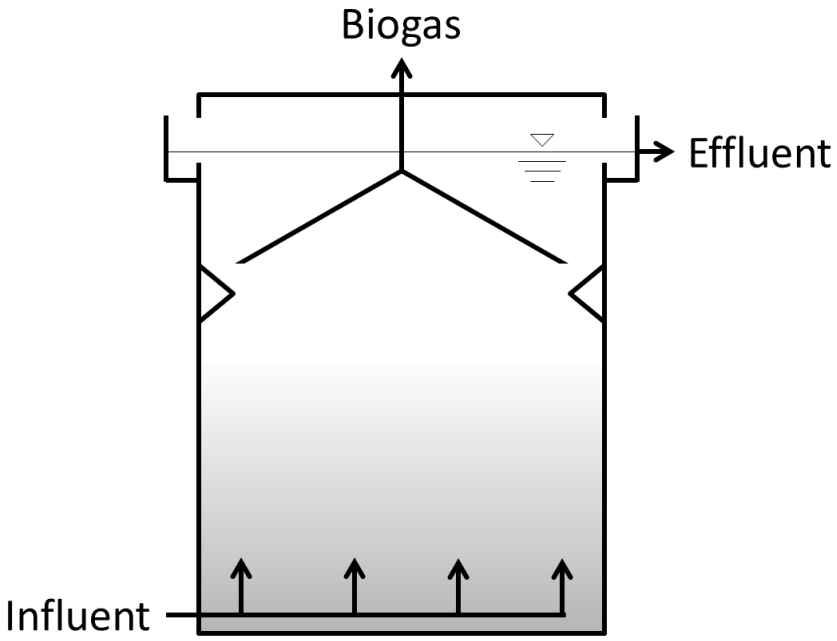


Figure 1-3 – Up-flow anaerobic sludge blanket reactor

The influent enters the reactor through the bottom and the effluent flows out of the top. The effluent can be recycled to fluidize the sludge bed whether the contact time between wastewater and sludge is not sufficient to achieve desired performance by the adoption of a distributed influent on the reactor bottom.

Van Lier et al. (1994) developed a different UASB configuration to overcome mixing problems related to influent short-circuiting that brought to accumulation of intermediate products such as hydrogen and VFA in conventional up flow systems. Hence, the authors introduced baffles along

the reactor length to increase turbulence inside the reactor which resulted in the development of upflow staged sludge bed (USSB) reactor. The baffled system gave to this reactor type an approximation of a plug flow system in which different stages of AD are separated from each other. Therefore, optimal condition of hydrolysis, acidogenesis and methanogenesis can be achieved by appropriate positioning of the baffles preventing accumulation of intermediate compounds and sludge washout risks (Sevilla-Espinosa, Solórzano-Campo, & Bello-Mendoza, 2010).

The up-flow anaerobic sludge blanket reactor can achieve significant high loading rates and low HRT from hours to days. The up-flow anaerobic sludge blanket reactor has been shown to be reliable in treating industrial wastewater such as dairy, sugary, textile, and brewery – Table 1-5. High COD removal greater than 90 % were obtain treating dairy, sugary, and textile processes water with HRTs greater than 12 h and OLRs from 1 to 28 kg COD/m^3 d (Kalyuzhnyi et al., 1997; Mijaylova Nacheva et al., 2009; Somasiri et al., 2008). Furthermore, a study conducted by Cronin and Lo (1998) demonstrate that UASB technology is able to remove around 85 % of the COD treating brewery wastewater with an OLR equal to 3 kg COD/m^3 d and HRT of 12 h. However, the UASB reactors efficiencies depend on different parameters like wastewater composition, especially the concentration of various ions (Zhang & Maekawa, 1996) and presence of toxic compounds such as phenol (Fang & Chan, 1997). The temperature and pH are also known to have an impact on performances by affecting the degree of acidification and the product formation.

A modification of UASB systems is the EGSB, in which higher superficial liquid velocity is applied from 5 to 10 m/h compared to 1 to 3 m/h in a UASB configuration (Britz & Robinson,

2008). Therefore, the sludge blanket is expanded and granular sludge can be also found in the upper level of the reactor itself, which will provide better contact time between the wastewater with the microorganisms and enhance the capability of the system. Hence, the EGSB can achieve higher loading rates compared to the UASB. The EGSB technology can achieve OLR as high as 40 kg COD/m^3 d (Seghezzo et al., 1998). At the beginning of this century, more than 1200 full-scale high-rate anaerobic reactors were implemented to treating industrial wastewater around the world and most of them consist of UASB and EGSB (Lim & Kim, 2014).

1.2.3 Anaerobic fluidized bed reactor

The anaerobic fluidized bed reactor (AFR) technology relies on the implementation of media for bacterial attachment and growth kept in a fluidized state by drag forces exerted by the up flowing wastewater – Figure 1-4.

Under a fluidized state, each media composed of small particle size material such as sand and activated carbon provides a large surface area for biofilm formation and development. The high biofilm buildup surface promotes system efficiency. Thus, the AFR ensures the possibility to achieve higher OLRs and a more reliable stability than the UASB. In fact as attached growth system, the AFR is more effective than granular or flocculent biomass agglomerates as it favors the transport of microbial cells from the bulk to the media surface and thus enhances the contact between microorganisms and substrate Fernandez et al. (2008). Therefore, a greater surface area per unit of reactor volume can be achieved by the elimination of bed clogging or high solids effluent concentrations and the increase of SRT. Reduced reactor volumes would be beneficial in reducing capital costs. If necessary, a higher up-flow velocity and effluent recycling can be implemented to achieve bed expansion properties and increase the design features. In the

expanded bed designs a support material similar to the AFR is used (e.g., sand, gravel or plastics) but the diameter of the particles is slightly bigger than in fluidized beds configurations.

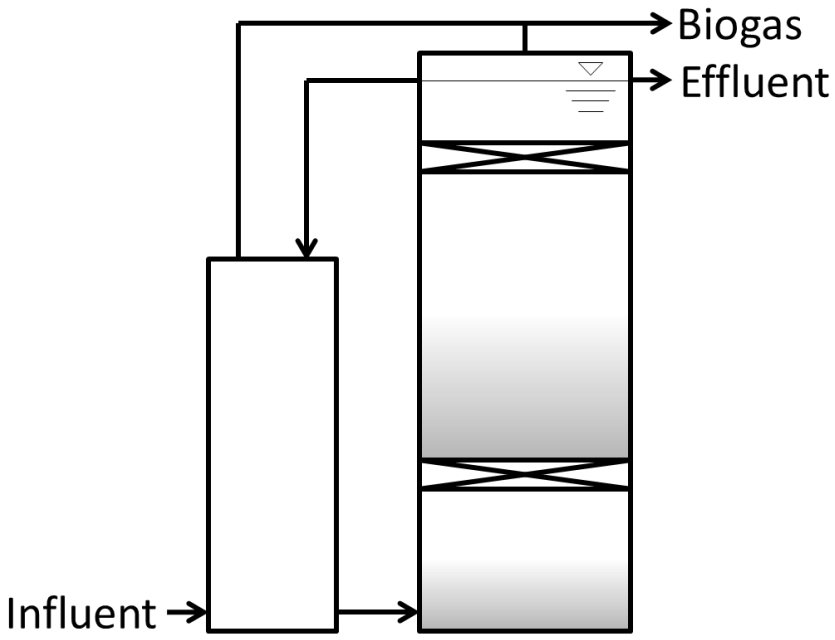


Figure 1-4 – Anaerobic fluidized bed reactor

Several studies on AFR process demonstrated its feasibility to treat different industrial wastewater streams – Table 1-6. Studies on winery wastewater conducted by Garcia-Calderon et al. (1998) demonstrated that an AFR with an inverse flow was capable of 85 % removal of total organic carbon (TOC) at 1.5 d HRT and an OLR equal to 4.5 kg COD/m^3 d utilizing perlite as a carrier. Furthermore, inverse fluidized bed reactor (IFBR) was tested with different supports (e.g., triturated polyethylene and silica particles) on brewery wastewater (Alvarado-Lassman, Rustrián, García-Alvarado, Rodríguez-Jiménez, & Houbron, 2008). The authors used two different ORLs (i.e., 10 and 30 kg COD/m^3 d) depending on the surface area of the carriers utilized. Triturated polyethylene had available surface area for biofilm development equal to 16.6 m^2/m^3, while 35.5 m^2/m^3 was estimated for the silica particles (extendosphere). Removal rates

slightly below 90 % as COD were achieved with the IFBR with an HRT and OLR of 4 d and 30 kg COD/m^3 d, respectively.

The anaerobic fluidized bed reactor technology has shown to be more reliable also with different wastewaters than the one produced by the beverage industry. For example, a study carried out by (Patel & Madamwar, 2001) showed 95 % COD removal at 4 d HRT and OLRs above 20 kg COD/m^3 d for the treatment of petrochemical processed wastewater.

Table 1-6 – AFR treating different industrial wastewaters

Wastewater treated	Carriers type	OLR (kg COD/ m^3 d)	COD_{rem} (%)	HRT (d)	Methane Yield (m^3CH_4/kg COD_{rem})	Reference
winery distillery (IFBR)	perlite	4.5*	85*	1.5	-	Garcia-Calderon et al. (1998)
Brewery (IFBR)	triturated polyethylene	10	< 90	4	0.35	Alvarado-Lassman et al. (2008)
	Silica	30	< 90	4		
petrochemical	bone charcoal	21.7	95	2.5	0.45	Patel and Madamwar (2001)
		20.4	95	4	0.37	

IFBR – Inverse Fluidized Bed Reactor; * as Total Organic Carbon (TOC).

1.2.4 Anaerobic baffled reactor

McCarty and coworkers at Stanford University were the pioneers of the anaerobic baffled reactor (ABR). An Anaerobic baffled reactor can be describe as a series of down and upflow anaerobic sludge blanket reactors spaced by a series of vertical baffles forces which drive the wastewater to flow under and over them as it passes from inlet to outlet (Figure 1-5). Microorganisms within the reactor rise and settle due to flow characteristics and gas is produced in each compartment. The biomass moves horizontally down the reactor at a relatively slow rate giving an SRT of 100 days at 20 h HRT (Grobicki & Stuckey, 1991). Therefore, the wastewater can be exposed to a large amount of active biomass as it passes through the ABR with as short an HRT as 6 h.

The most significant advantage of the ABR is its ability to separate acidogenesis and methanogenesis longitudinally down the reactor, allowing the different microorganisms involved to develop under favorable conditions. Disadvantages of the baffled reactor design at pilot/full scale include wide, shallow reactor to maintain acceptable liquid and gas up-flow velocities, and problems associated with even influent distribution (Tilche & Vieira, 1991).

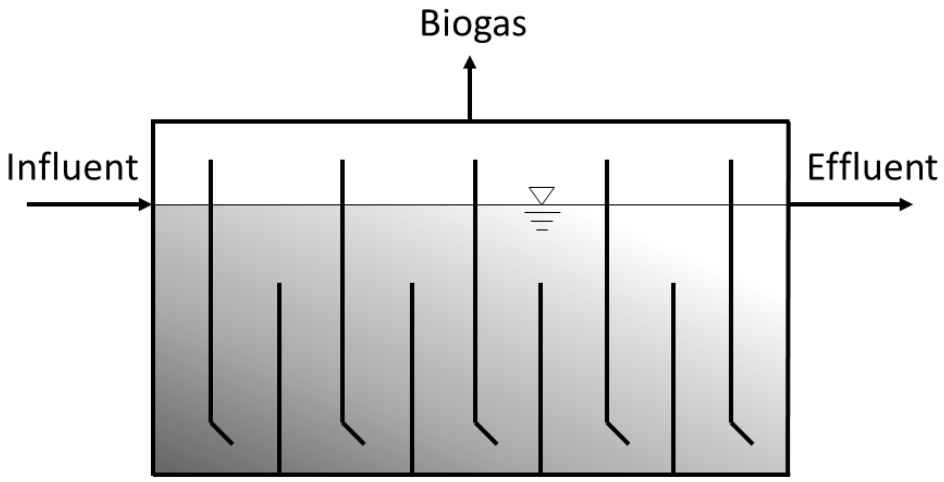


Figure 1-5 – Simplified baffle reactors cross section (single gas headspace)

Table 1-7 – Anaerobic baffled reactor treating different industrial wastewaters

Wastewater treated	OLR (kg COD/m^3 d)	COD_{rem} (%)	HRT (d)	Methane Yield (m^3CH_4/kg COD_{rem})	References
Brewery	5.6	90	15	-	Hassan and Dahlan (2013)
Pulp and Paper	5	60	48	0.18	Grover et al. (1999)
Palm Oil Mill	1.6	77	3	0.33	Faisal and Unno (2001)

Other issues with the ABR system is the increasing of solid loss over time, increasing of hydraulic dead space, disruption of bacteria communities and bioflocs (Barber & Stuckey, 1999). The anaerobic baffled reactor has been used to treat a wide range of industrial wastewater. Examples of the performance of ABR are reported in the following table – Table 1-7.

1.2.5 Anaerobic hybrid reactor

Anaerobic hybrid reactor (AHR) were developed to overcome problems in sludge retention which has been a major issue encountered in conventional UASB reactor. In fact, these reactors present difficulties in treating complex wastewaters containing suspended and colloidal particles (e.g. fats, proteins and lipids). The presence of these materials hinder the methanogenic activity by reducing the mass transfer properties as well as the biogas escape because of adsorbed on the sludge surface. Therefore, the granular sludge will become fluffier or even degranulate into flocs and ultimately wash out from the reactor. An overview of an anaerobic filter UASB hybrid reactor is presented in Figure 1-6.

The lower part of the reactor consist of suspended growth biomass while in the upper part a material is used to facilitate the attachment of biomass. This type of design is called anaerobic filter UASB hybrid reactor and it has been mainly used for industrial wastewater with high content of toxic and inhibitory substances such as oil, grease, phenolic compounds and tannin (Jafarzadeh, Mehrdadi, & Hashemian, 2012; Kumar, Yadav, Sreekrishnan, Satya, & Kaushik, 2008). The suspended sludge bed act as buffer zone for these harmful compounds while the filter on top ensure digestion of less harmful compounds, by-products of uncomplete digestion and residual VFA. For these reasons several type of AHR were developed and a comparison between different reactor configurations and treated wastewater is presented in Table 1-8.

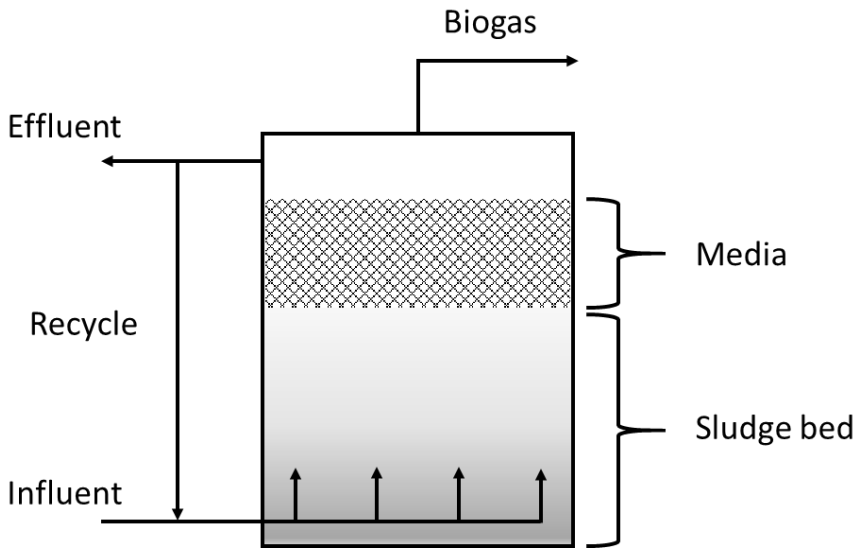


Figure 1-6 Anaerobic hybrid reactor

Table 1-8 – Anaerobic hybrid reactor treating different wastewaters

Reactor Type	Wastewater treated	OLR (kg COD/m^3d)	COD_{rem} (%)	HRT (d)	Methane Yield (m^3CH_4/kg COD_{rem})	References
AF-UASB	distillery	8.7	80	5	0.34	Sunil Kumar et al. (2007)
UASFF	yeast	9	75	3	0.32	Büyükkamaci & Filibeli (2002)
UASB-MBBR	fruit-juice	6	65	0.8	0.33	Tawfik & El-Kamah, (2012)
	winery	25	> 80	0.75	-	Wahab et al. (2014)
UASB-AFR	synthetic*	6	75	1.7	0.21	Kundu, Sharma, & Sreekrishnan (2013)

*glucose and yeast; AF-UASB: anaerobic filter – upflow anaerobic sludge blanket reactor; UASF: upflow anaerobic sludge bed fixed film reactor; UASB-MBBR: upflow anaerobic sludge blanket – moving bed biofilm reactor; UASB-AFR: upflow anaerobic sludge blanket – anaerobic fluidized bed.

Anaerobic filter upflow - anaerobic sludge blanket (AF-UASB) reactor is a combination of sludge blanket in the lower part and filter in the upper. It has been demonstrate to achieve 80 % COD removal at 5 d HRT and 8.7 kg COD/m^3 d (Sunil Kumar et al., 2007). Upflow anaerobic sludge bed fixed film (UASFF) reactor however, instead of a filter it utilizes a fixed media on the top of the reactor on which the biofilm is developed. These has shown to have an advantage in

retaining higher biomass concentrations inside the reactor by preventing biomass washout (Tauseef, Abbasi, & Abbasi, 2013). The UASFF, it also reduces short circuiting problems and improves gas-solid/gas-liquid separation (Najafpour, Zinatizadeh, Mohamed, Hasnain Isa, & Nasrollahzadeh, 2006). Büyükkamaci and Filibeli (2002) reported that this type of AHR could achieve removal of 75 % with an OLR of 9 kg COD/m^3 d treating yeast industry effluent. Kumar et al. (2008) proposed a reactor configuration UASB-AFR that has features of upflow anaerobic sludge blanket reactor (UASB) as well as anaerobic fluidized-bed reactor (AFR) by maintain a superficial liquid velocity that was higher than in a typical UASB but lower than in a AFR. This type of reactor achieved up to 75 % COD removal treating glucose and yeast synthetic wastewater at OLR of 6 kg COD/m^3 d at 1.7 d HRT (Kundu et al., 2013). Another modification to prevent biomass washout in UASB configuration is the implementation of free-floating biocarriers (e.g., lignocellulosic supports) supports on upper part of the reactor. This hybrid UASB-MBBR design was capable to remove more than 80 % of the COD with less than one day HRT at OLR up to 25 kg COD/m^3 d (Kundu et al., 2013).

1.2.6 Anaerobic moving bed biofilm reactor

Anaerobic moving bed biofilm reactor (AMBBR) systems use media that facilitate the attachment of bacteria on their surfaces. This particular feature allows an optimal contact time between biomass and substrate while maintaining an appropriate sludge age on the retained media – Figure 1-7.

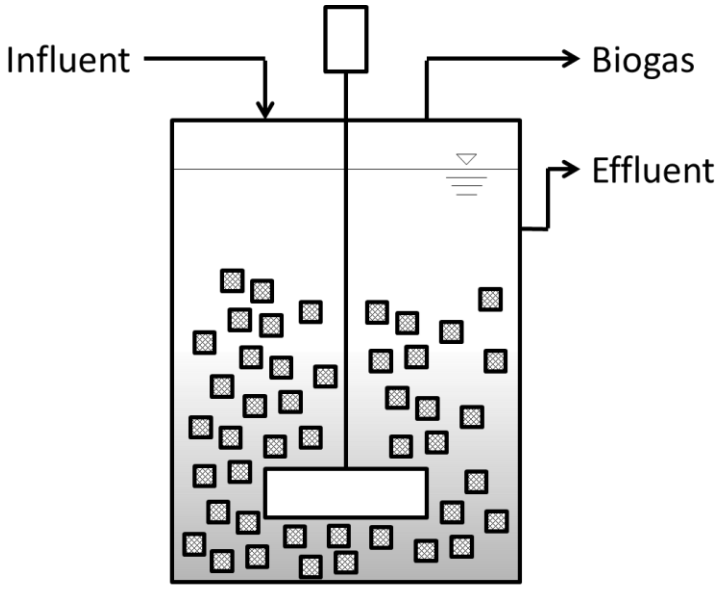


Figure 1-7 – Anaerobic moving bed biofilm reactor

The capability to separate SRT from HRT is the key component of the high-strength industrial wastewater reactors. The anaerobic moving bed biofilm reactor retains appropriate microbiological community in the system by biofilm development on suspended media. The key benefits of this type of high rate anaerobic digester are the cost effectiveness that lower capital and maintenance expenditures due to its small footprint. Moving bed biofilm reactor (MBBR) has been proven to be very reliable in aerobic treatment because of its high volumetric loading rates and low sludge accumulation in the reactor (Jahren, Rintala, & Odegaard, 2002). Although the aerobic MBBR is widely used in full-scale applications, the AMBBR technology is not yet implemented. Ødegaard et al. (1994, 1999) carried out consistent studies on MBBR systems within the last two decades in Northern Europe on municipal wastewater. McQuarrie and Boltz (2011) recently developed a North American design guide. With his research, Ødegaard proved the effectiveness and potential behind both aerobic and anaerobic design (e.g., BOD/COD-removal, nitrification, denitrification) in attached-growth systems. The feasibility of the MBBR reactors was demonstrated for a wide range of treatment purposes: municipal and industrial

wastewater, aquaculture, secondary and tertiary treatment, and side stream applications (McQuarrie & Boltz, 2011).

Mixing is done by aeration agitation in aerobic reactors, or mechanical mixing and/or recirculation systems in anaerobic and anoxic configurations. The carriers are usually made of polyethylene with a density close to 1 g/cm^3 (Ødegaard, 1999), which allows them to move freely even with 70 % of the volume occupied by carriers in a reactor (Metcalf & Eddy, 2013). Because the reactor is completely mixed, the whole reactor volume is active, compared to most of the other modern biofilm reactors on the market (e.g., fixed film reactor).

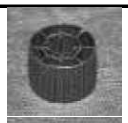
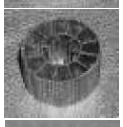
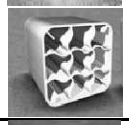
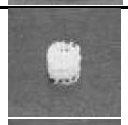
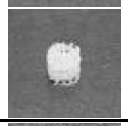
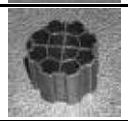
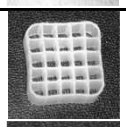
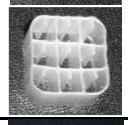
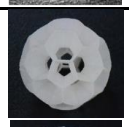
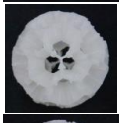
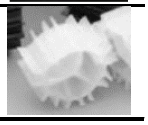
Most of the interest in MBBR technology is on development of different types of media that improve the surface area colonisable by the biofilm (protected surface area) and therefore will lead to better performance. Moreover, the media shape and dimensions are crucial in maintaining the protected surface area and limiting mass transfer limitation issues. The carrier features are important for limiting the volume utilized to have desirable performance and optimal mixing properties optimization. Most of the carriers manufactured by several companies and their main characteristics are presented in Table 1-9 Different shapes have been manufactured, such as cylindrical (AC 450, K3, K1, CM-10D^{TM}), cubical (AC920, BWTX^{TM}), spherical (3D-1, 2 and 3), and sponge-like (APG). These media are used for different attached growth processes mainly for anaerobic-aerobic MBBR for municipal wastewater, sidestream processes but very few are used in industrial wastewater applications. The release and manufacture of different carriers is under continuous development. For instance, the AnoxKaldnes biofilm chip (P) were used by Veolia for anaerobic ammonia oxidation biofilm development in partial-nitritation systems (ANITA Mox). However, Veolia had recently developed a unique media called Z-MBBR that overcame mass transfer limitations by a defined height grid that by sour mechanisms do not

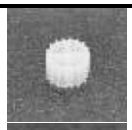
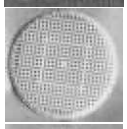
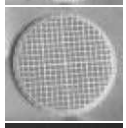
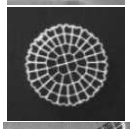
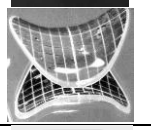
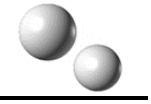
allow the biofilm to grow more than a certain thickness. The thin biofilm formation of slow growing bacteria such as anaerobic ammonia oxidation ensure mass transfer limitation and inactive microorganisms or biomass aging on the support (Kermani, Bina, Movahedian, Amin, & Nikaein, 2008).

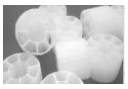
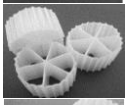
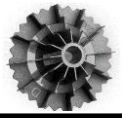
In 2014, more than 1,200 wastewater treatment plants in at least 50 countries utilize the MBBR technology (Biswas, Taylor, & Turner, 2014). According to Rusten et al. (2006) the first MBBR installed was implemented in Norway (European Patent No. 0.575,314; U.S. Patent No. 5,458,779) and was inspected for more than 15 years and no performance-influencing plastic biofilm carrier were observed ensuring long life expectancy. Most of the MBBRs processes are implemented for municipal wastewater streams since the high-strength wastewaters are mainly treated with UASB at full scale. In the past decades, the interest in testing the performance of media in anaerobic/aerobic condition for industrial wastewater treatment arose.

Table 1-10 outlines studies on AMBBR for industrial wastewater treatment and their main findings. In particular, Sheli et al. (2007, 2010, 2014) research was focused on testing and design optimization using Bioflow 30 and 9 as carriers treating winery wastewater. The authors shown that the Bioflow 9, with a wider surface area (530 m^2/m^3) than the Bioflow 30 (320 m^2/m^3) were capable to achieve comparable removal rates of 80 % COD at 36 – 37 h HRT and 66 % of media as reactor filling. Wang et al. (2009) tested the performances and operational design parameters using Bioflow 9 at 65 % media volume treating dairy wastewater. In this study, 85 % of the COD were removed at an OLR equal to 20.0 kg COD/m^3 d within 6 h HRT.

Table 1-9 – Different type of media and specific surface area

Manufacturer	Name	Specific surface area (m^2/m^3)	Shapes
Headworks BIO	ActiveCell™ 450	402	
	ActiveCell™ 515	485	
	ActiveCell™ 920	680	
AqWise	ABC4™	600	
	ABC5™	650	
Entex Technologies Inc.	Bioportz™	589	
EVOQUA	CM-10D™	750	
Biowater Technology	BWT15™	828	
	BWTX™	624	
Dong et al. (2015)	3D-1	437	
	3D-2	560	
	3D-3	600	
Nisshinbo Chemical Inc.	APG	-	
RVT	Bioflow 9	800	

Manufacturer	Name	Specific surface area (m^2/m^3)	Shapes
Veolia Inc.	AnoxKaldnes™ K1	500	
	AnoxKaldnes™ K3	500	
	AnoxKaldnes™ Biofilm Chip (M)	1,200	
	AnoxKaldnes™ Biofilm Chip (P)	900	
	AnoxKaldnes™ K5	800	
	Z-MBBR	-	
GEA 2H Water Technologies GmbH	BCN 011	520	
	BCP 750	635	
	BCN 020	400	
	BCN 040	272	
	BCN 030	259	
	BCN 060	189	
Refill-Tech S.r.l	BFM	160	
	SAGM 500	513	
	HOLLOW BALLS	82-141	

Manufacturer	Name	Specific surface area (m^2/m^3)	Shapes
World Water Works	Ideal MBBR	650	
Process Engineered Water Equipment, LLC	ASO™ Bio-Carrier	626	
Bioprocess H2O	BioFAS™ B-460	402	
	BioFAS™ B-585	515	
Createch Aqua ApS	Curler Advance X-1	800	
	Cylinder Plus	350-550	
	Cylinder X-0	600-900	
Energy Equipments Private Ltd.	MBBR Media	400	
Mutag	Mutag BioChip™	3000	
Hebei Sinta FRP	PE05	> 500	
	PE04	> 800	
	PE10	> 1200	
Warden Biomedia	Biofil	135*	
	BioMarble	310*	

* Total surface area

Table 1-10 – AMBBR treating industrial wastewater

Wastewater treated	Carrier type	V_M %	SSA** (m^2/m^3)	OLR ($kgCOD/m^3d$)	COD %	Max SSAA*** ($gCOD/m^2d$)	HRT (h)	Methane Yield ($m^3CH_4/KgCOD_{rem}$)	References
Winery	30*	66	320	18.5	80	28	37	0.33	Sheli et al. (2014)
	9*	66	530	29.5	80	112		0.25	
Winery	9*	66	530	29.5	81	76	36	0.36	Sheli et al. (2007)
Winery	30*	66	320	18.5	86	75	60	0.30	Sheli et al. (2010)
Dairy	9*	65	530	20.0	73	50	6	0.34	Wang et al. (2009)
Synthetic	K3	50	600	4.5	64	-	3	-	Sabzali et al. (2012)
	Cigarette filters	8.3	18,520 – 23,560	4.5	61		21		

* Bioflow type; ** As total COD; SSA – Specific Surface Area; *** SSAA – Specific Surface Area Activity

1.3 Brewery Industry

Based on the Canadian Beer Association, breweries across the Canadian provinces are steeply growing over the last few years. In 2014, there were 520 licensed breweries, an increase of 70 % compared to 1999 (BeerCanada, 2016). Typically, the beer production process involves three main steps: brew house, fermentation and beer processing. The brew house is where beer develops its taste and character through different production steps such as milling, mashing, and wort boiling/cooling. The step after is the fermentation process in which the sugars are transformed into alcohol by the yeast metabolism. Typically, the maturation of the beer occurs in two steps within the first and second stage fermentation tanks. Afterwards, the final step of beer processing is filtration, bottle washing and filling before delivery. Based on the "Canadian Breweries Industrial Report" (Petrillo, 2015), the Canadian breweries are most predominant in Ontario (34.3 %), followed by British Columbia (26.5 %) and Quebec (22.4 %) while Manitoban breweries account for 1.4 % of the total.

The brewing industry is one of the largest industrial users of water despite the significant technological improvements made over the last few decades. Water and energy consumption, wastewater production, solid waste, by-productions management, and air emission are the major environmental issues associate with the brewery industry (Olajire, 2012). Typically, the water usage for beer production ratio is between 4 to 10 L/L while the wastewater generation per beer ratio ranges from 1.3 to 3.5 L/L – Table 1-10. Brewery wastewater is classified as high-strength because of its high organic carbon content concentrations. The COD concentrations are typically between 1.8 to 5.5 kg COD/m^3 and the BOD_5 ranges from 0.6 to 5 kg BOD_5/m^3 with an average BOD_5/COD ratio of 0.77. Moreover, 0.03 – 0.1 kg/m^3 is the concentration of both total nitrogen and total phosphorus.

Table 1-11 – Typical brewery wastewater characteristics (Chastain et al., 2011)

Parameter	Typical Range	Unit
Water to beer ratio	4 – 10	L/L
Wastewater to beer ratio	1.3 – 3.5	L/L
BOD_5	0.6 – 5	kg/m^3
COD	1.8 – 5.5	kg/m^3
Total Nitrogen	0.03 – 0.1	kg/m^3
Total Phosphorus	0.03 – 0.1	kg/m^3
pH	3 – 12	-
Total suspended solids	0.2 – 1.5	kg/m^3

Table 1-12 – Requirement from City of Winnipeg sewer by-law No.92/2010

Parameter	Threshold	Unit
TSS	0.35	kg/m^3
BOD_5	0.3	kg/m^3
TN	0.06	kg/m^3
TP	0.01	kg/m^3
Animal or vegetable oil and grease	0.1	kg/m^3
Mineral or synthetic oil and grease	0.15	kg/m^3
pH	5.5 – 11	-

The Canadian by-law industrial discharges in sewer systems to 300 g BOD_5 /m^3, 350 g TSS /m^3, pH 6.5 – 10.5 (Brewers Association of Canada, 2011). Table 1-12 shows the thresholds for high-strength wastewater discharged into the sewer system for the city of Winnipeg. The BOD_5 and TSS thresholds are the same, while the pH is slightly different (5.5 – 11). Total nitrogen is regulated to not exceed 0.06 kg/m^3, total phosphorus is 10 g/m^3.

1.4 Objectives

The goal of this project was to develop an AMBBR to treat industrial wastewaters using AC920 plastic media as support for biomass growth. The thesis work was divided into three parts: reactors start-up on synthetic substrate and transition to brewery wastewater, analysis of AMBBR treating brewery wastewater when subjected to increasing organic load via decreased HRT, design consideration and case study for the Fort Garry brewery.

To meet the goal specific objectives were defined as follows:

I. Reactor start-up and development of attached biomass at bench scale;
II. Determination of feasibility to treat industrial wastewater at a high OLR;
III. Comparison of the organic load to gas production;
IV. Determination of process stability and resilience in conditions of variable OLR;
V. Testing the reactors with real wastewater from a local brewery;
VI. Determination of process performance at different HRTs (24h, 18h, 12h, 8h, and 6h);
VII. Design suggestions for full-scale application.

CHAPTER 2 – MATERIAL AND METHODS

Four-liter working volume reactors were operated at 35°C (SR-Chamber, CONVIRON, Winnipeg, Manitoba, Canada) and inoculated with biomass from an anaerobic digester (North End Water Pollution Control Centre, Winnipeg, CA) to an initial solids content of 15 kg TSS/m^3 (0.6 VSS/TSS). Gas-proof tubes (Nalgene 180 PVC 0.635 cm ID, 1.5 OD) and glass gas bottles with septum lids were utilized for gas collection. Digital mass flow meters (FMA 4000, OMEGA, Laval, Quebec, Canada) were applied inline to determine the rate of total biogas production. A syphon was used to maintain a constant working volume under a flow of approximately 4 L per day and mechanical stirrers were implemented for mixing purpose. Cubic AC920 plastic media (Headworks-Bio; Victoria, BC) were used providing specific protected surface areas of 680 m^2/m^3.

2.1 Reactors setup

2.1.1 Start-up on synthetic and transition to brewery wastewater

Two completely sealed reactors were implemented in the first part of the research. Plastic media AC920 was added as 25 % working volume in one reactor and 35 % in the other, and 1 d HRT was maintained in both the systems. Since the protected surface area of the media is 680 m^2/m^3 approximately 0.68 and 0.95 m^2 was available for biomass cultivation in the 25 % and 35 % media reactors, respectively. An overview of the final reactor configuration is presented in Figure 2-1. Synthetic feed was initially provided at the start-up period with 0.5 kg COD/m^3, comprising acetate (CH_3COONa), glucose ($C_6H_{12}O_6$) and yeast extract (Sigma-Aldrich, Oakville, Ontario, Canada) at a ratio of 1.2 : 1.0 : 1.8. To ensure buffer capacity and maintain

neutral pH between 6.9 and 7.7 to support methanogenesis, 1.5 kg $CaCO_3/m^3$ and trace elements were added - Table 2-1.

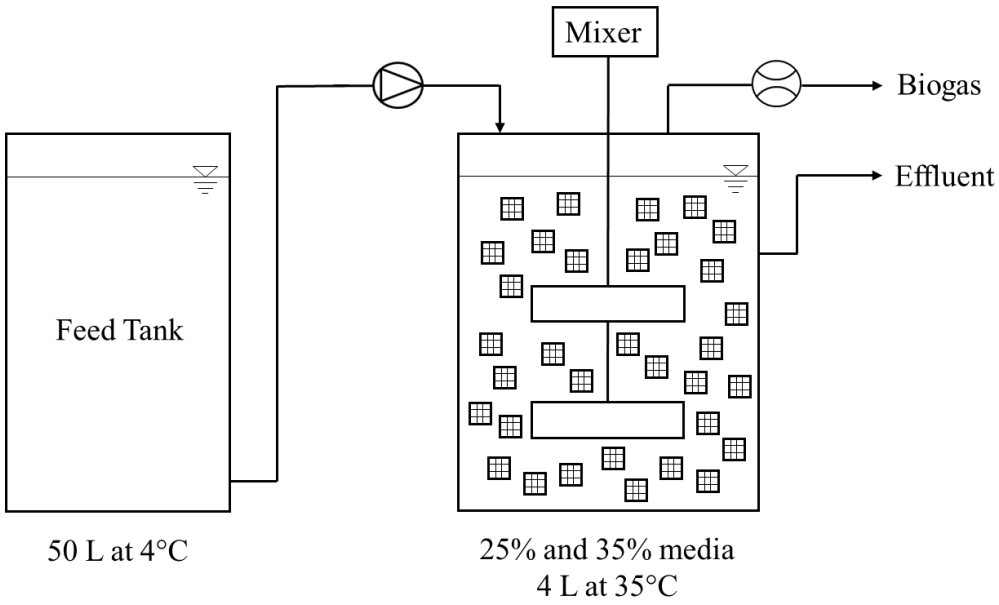


Figure 2-1 – Start-up reactors setup

Table 2-1 – Micronutrient concentrations

Micronutrient	Value	Units
$FeCl_2•4H_2O$	1.5	mg/L
$MnSO_4•H_2O$	0.085	mg/L
$CaCl_2$	0.18	mg/L
$MgSO_4•H_2O$	0.69	mg/L
$CoCl_2•6H_2O$	0.48	mg/L
H_3BO_3	0.006	mg/L
$ZnSO_4•7H_2O$	0.14	mg/L
$CuSO_4•5H_2O$	0.0025	mg/L
$Na_2MoO_4•2H_2O$	0.556	mg/L
$NiCl_2•6H_2O$	2.38	mg/L
$Na_2SeO_3•5H_2O$	0.263	mg/L

The trace elements were only incorporated during the start-up period and subsequently removed once the reactors started consistently removing COD. At the same time, acetate and glucose were removed and the total COD was provided by a mixture of dried milk (Similac, Winnipeg,

Canada), beef and yeast extracts (Sigma-Aldrich, Oakville, Ontario, Canada) at a ratio of 1.75 : 1.00: 2.25.

Once acclimated to the new substrate, as determined by COD removal performance, the OLR was brought up by increasing the concentration of COD components in the feed. After achieving the load of 20 kg COD/m^3 d the feed was switched to brewery wastewater supplied at around 3 kg COD/m^3 d. The proper concentration in the brewery wastewater reactors influent was obtained by diluting stock brewery wastewater withdrawn from the fermenters.

2.1.2 Organic Loading Rates and HRT impact

Three reactors with identical operational conditions were used within the second part of the thesis work. After the start-up on synthetic wastewater and transition to brewery wastewater study, the colonized AC920 media were distributed between three reactors in order to achieve 40 % media volume (500 ± 50 units) in each of the three reactors with addition of new uncolonized media – Figure 2-2.

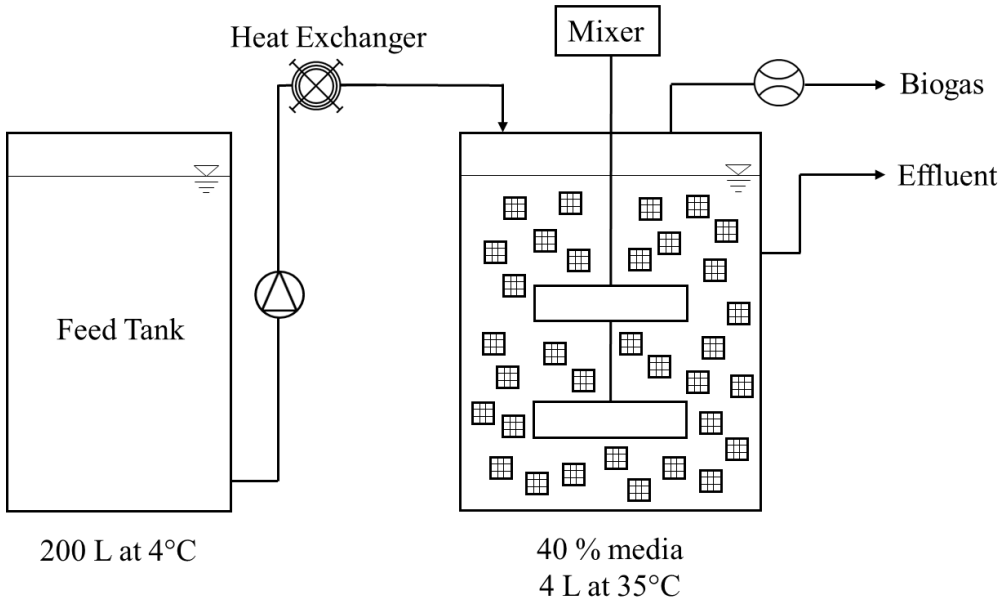


Figure 2-2 - Reactor setup of the HRT and OLR study

A 200 L tank was used to store the brewery wastewater in a chamber at 4°C in order to prevent decomposition and fermentation of the feedstock before being pumped into the reactors. The cold feed was passed through a heat exchanger to bring the temperature up to 35°C before being continuously pumped into the three reactor with a digital pump. Protected surface area of 680 m^2/m^3 resulted approximately in 1.1 m^2 available surface for biofilm development in each reactor. Variable HRT were applied (24, 18, 12, 10, 8 and 6 h). The time between load/HRT changes was approximately three weeks.

2.2 Analytical methods

2.2.1 Performance measurements and equations

Samples were taken from the reactors' influent and effluent three times per week. The pH was recorded immediately upon sampling and alkalinity was measured according with Standard Methods (APHA, 2012). Total COD and soluble COD were analyzed spectrophotometrically as per Standard Methods (APHA, 2012), as well as total and volatile suspended solids (TSS and VSS). Furthermore, BOD_5 was determined using Oxitop bottles (model IS 12, Xylem, CA). The Volatile Fatty Acid were also determined by waters Breeze 2 HPLC system (Waters, Milford, MA, USA) equipped with an Aminex HPX-87H column using 5 mM H_2SO_4 as the mobile phase, calibrated using a VFA standard 10 mM mixture (Supelco, Bellefonte, PA, USA).
Specific Surface Area Activity (SSAA) was determined assuming the protected surface area as available surface for microbial growth and calculated as follows:

$$\text{SSAA}\ (\text{g COD m}^2/\text{d}) = \frac{Q\cdot(S_0-S)}{SSA\cdot(V_M/100)\cdot V} \qquad \text{(Equation 1)}$$

In which Q is the continuous flow applied (m^3/d), S_0 is the substrate concentration in the influent and S in the effluent as kg sCOD/m^3 or kg COD/m^3, SSA is the Specific Surface Area equal to

680 m^2/m^3, V is the reactor volume (i.e., 0.004 m^3) while V_M is the total volume of media represented as percentage of working volume (40 %). Since mixing intensity and collisions with other media likely did not allow for significant biofilm growth on the outer edges of media, SSA was taken as the interior surface area of the AC920 media (680 m^2/m^3) instead of the total surface area (920 m^2/m^3).

The total gas was recorded every second in mL/min using FMA 4000 OMEGA mass flow meter. The composition of gas (CH_4, CO_2, O_2 and H_2) collected in the septum gas bottles was determined by gas chromatography (GC, 490 Micro GC, Agilent Technologies, Santa Clara, CA) once a week. Since the flow meter registered the gas produced as nitrogen, it was necessary to convert the gas produced into the actual biogas after GC analysis was conducted to determine the composition as methane and carbon dioxide. Therefore, the total gas recorded was first converted to m^3/d and by applying gas correction factor of K = 0.7210 and with temperature correction factor T = 1.1 at 35° C provided by OMEGA. Hence, it was possible to determine the total gas as actual total biogas by multiplying the gas produced per day as nitrogen with both gas and temperature correction factors. Ultimately, to quantify the m^3/d produced as single gas (i.e., methane and carbon dioxide) the following equation was used:

$$CH_4(m^3/d) = \left(\frac{\text{averaged methane (\%)}}{100}\right) * \text{Total biogas } (m^3/d) \qquad \text{(Equation 2)}$$

Where the averaged methane (%) was quantified from three replicates of each sample taken from the three reactors once per week. Specific methane generation rate per unit of COD removed was determined as:

$$\frac{m^3CH_4}{\text{kg COD}_{rem}} = \frac{m^3CH_4/d}{(\text{kg COD}_0/m^3 - \text{kg COD}/m^3) * Q\ (m^3/d)} \qquad \text{(Equation 3)}$$

2.3 Kinetics tests

Periodic kinetic tests were performed on the mixed liquor to evaluate if there was any contribution to the organic removal capability of the system by suspended microorganisms. Different approaches were adopted within the two stages of the project. In both stages, pure glucose at 98 % (Sigma-Aldrich Canada Co.) was chosen as substrate because of monomeric structure of more complex sugars. This approach overestimate the suspended biomass contribution in case any removal was observed and therefore gives conservative and consistent results. From graphs obtained by plotting the average removal in kg sCOD/m^3, a correlation curve was developed and a removal rate coefficient was extrapolated. The contribution of suspended biomass to total removal of sCOD as kg sCOD m^3/d as well as percentage of the total sCOD removed was thus determined.

2.3.1 Start-up on synthetic and transition to brewery wastewater

In the first study, performance of suspended biomass was quantified with kinetic tests during the period of the highest loading at the OLR greater than 20 kg COD/m^3 d. During this time, effluent COD was greater than 3 kg/m^3 in both reactors and the acetate portion was between 0.25 to 0.50 kg/m^3, which was greater than the half-saturation coefficient for acetoclastic methanogenesis (Environsim, 2015). Therefore, maximum removal rates observed during kinetic tests were representative of rates occurring in the reactors, especially since the same composition of feed supplied to the reactors was used. Solution was withdrawn from the reactors and transferred to 0.5 L flasks (mixed at 65 rpm at 35°C), the COD of which was supplemented to 2 kg/m^3 as 98 % glucose (Sigma-Aldrich Canada Co.). Samples were taken every 30 minutes from time zero and up to three hours. Samples were immediately filtered through 0.45 μm filters (Whatman, Sigma-

Aldrich, Oakville, Ontario, Canada) and were analyzed for soluble COD only. Additional samples were also collected after 24 h from the time zero. Maximum removal rates observed during kinetic tests were assumed representative of rates occurring in the reactors at that condition, even though the type of substrate dosed was different.

2.3.2 Organic Loading Rates and HRT impact

Performance of suspended biomass was also quantified during the second stage of the research. The kinetic tests were carried out at the end of each period in which one particular HRT was maintained and before reducing it to desirable values. Therefore, mixed liquor was withdrawn from each reactors and transferred to 0.5 L flasks in which as much as 5 kg sCOD/m^3 was supplied as 98 % glucose (Sigma-Aldrich Canada Co.). The kinetic tests were 4 hours long and every hour (i.e., 0, 1, 2, 3, 4) 5 mL of sample was collected, filtered through 0.45 μm filters for sCOD measurement as well as 24 hour after from the time zero.

2.4 Brewery characterization

The brewery wastewater used in this study came from Fort Garry Brewery in Winnipeg, CA. This is a small sized brewery equipped with 15 continuously operating fermenters and 8 maturation tanks. The total daily beer production is 46 m^3/d. The brewery produces 138 m^3/d of wastewater, which is within typical range for North American breweries of 2:1 to 8:1 (v/v) reported by (Chastain et al., 2011). During the higher production period in summer, two months (May and June, 2015) were analyzed to identify the wastewater characterization profile reported in Table 2-2.

Table 2-2 - Brewery wastewater characterization

Parameter	Value	Unit
sCOD	5.8 ± 2.2	kg sCOD/m^3
COD	6.9 ± 2.7	kg TCOD/m^3
BOD_5	5.3 ± 0.2	kg BOD_5/m^3
BOD_5/COD	0.77	-
TSS	0.35 ± 0.12	kg TSS/m^3
VSS	0.3 ± 0.1	kg VSS/m^3

BOD_5 is the 5 days Biological Oxygen Demand; sCOD is the soluble Chemical Oxygen Demand; COD is the total Chemical Oxygen Demand; TSS is the Total Suspended Solids and VSS are the Volatile Suspended Solids.

Over this period, 5.8 and 6.9 were the average soluble and total COD in kg sCOD/m^3 and kg COD/m^3, respectively while 5.3 was the kg BOD_5/m^3 resulting in a BOD/COD ratio of 0.77 higher than the typical values of 0.6 - 0.7 (Brito et al., 2004).

In the thesis work, the COD concentrations were kept as close as possible close to the averaged brewery wastewater concentration of 3.6 kg COD/m^3 according to North American Brewers Association (Chastain et al., 2011) by diluting the yeast sampled from the bottom of the fermenters to desirable values. This resulted, in influent concentrations of 4.0 ± 1.0 as kg sCOD/m^3 and 5.2 ± 2.1 kg COD/m^3. The BOD was measured to be 3.9 ± 1.6 kg BOD_5/m^3 with a BOD_5/COD ratio of 0.75.

CHAPTER 3 – RESULTS AND DISCUSSION

3.1 Start-up and transition to brewery wastewater

3.1.1 Physico-chemical Parameters

Two months were required for a mature biofilm to establish itself on the media and begin performing to desirable levels of treatment (80 % COD removal). This is in line with results from Fernández et al. (2008), studying a UASB treating brewery wastewater. The authors demonstrated, using microbial ecology methods (i.e., FISH, DGGE), that the three steps of biofilm development (attachment, consolidation, and maturation) occur within two weeks to two months from inoculation.

In this study pH ranged between 6.9 and 7.7, while alkalinity was 2 and 5 g $CaCO_3$/L (Appendix 6.1.1). The pH values demonstrated that the AMBBRs were able to promote methanogenesis and the colonization of media. Dosing 1.5 kg $CaCO_3/m^3$ ensured the pH was buffered to neutral values, thereby providing a suitable environment for methanogenesis without the risk of acidic conditions (pH < 6), which will consistently inhibit methanogenic activity (Kleerebezem, Joosse, Rozendal, & Van Loosdrecht, 2015). The increase of pH above neutrality (7.3 to 7.7) and alkalinity values greater than 1.5 g $CaCO_3$/L demonstrated that the reactors had reached a balance between acidogenesis, acetogenesis and methanogenesis.

3.1.2 Removal Performances

Performance was quantified in terms of percent soluble COD removal throughout the study (Figure 3-1) and raw results are reported in Appendix 6.1.1. Both the 25 % and 35 % media

reactors reached stable removal (80 % COD) after approximately two months from inoculation, suggesting that the biofilm achieved a mature state.

Afterwards, the OLR was increased every week by increasing the concentration of COD in the synthetic feed until a significant decrease in COD removal was registered. Within the following two and a half months the loading was increased to 23.5 kg sCOD/m^3 d. Up until 20 kg sCOD/m^3 d, both reactors were able to remove more than 80 to 90 % of the sCOD. Once a drop in performance was registered at 23 kg sCOD/m^3 d, the feed was changed to a lower loading rate in order to recover at least 80% removal in both reactors. Specifically, the reactors were supplied 11 to 14 kg sCOD/m^3 d. An error in feed preparation occurred 10 days later and resulted in a load of 24 kg sCOD/m^3 d supplied to the reactors for four days. This corresponded with drops in performance as low as 60 % removal in the 25 % media reactor and 70 % removal in the 35 % media reactor. The load was then corrected down to 9 kg sCOD/m^3 d for four days and consequently dropped to 4 kg sCOD/m^3 d for two more days before performances above 80 % removal were observed. The reactors were then transitioned from synthetic feed to brewery wastewater, which was initially supplied at 3 kg sCOD/m^3 d. The brewery wastewater was gradually increased to 9 kg sCOD/m^3d within the next two weeks and both reactors were able to remove more than 80 % of the sCOD steadily.

The removal efficiency did not change significantly when the OLR was increased from 5 to 20 kg sCOD/m^3 d on a weekly basis. In fact, removal rates between 80 % to 90 % in both the 25 % and 35 % media reactors were achieved. The removal rates obtained in this study were comparable with other studies, reported in Table 1-9, in which different types of real wastewater were treated.

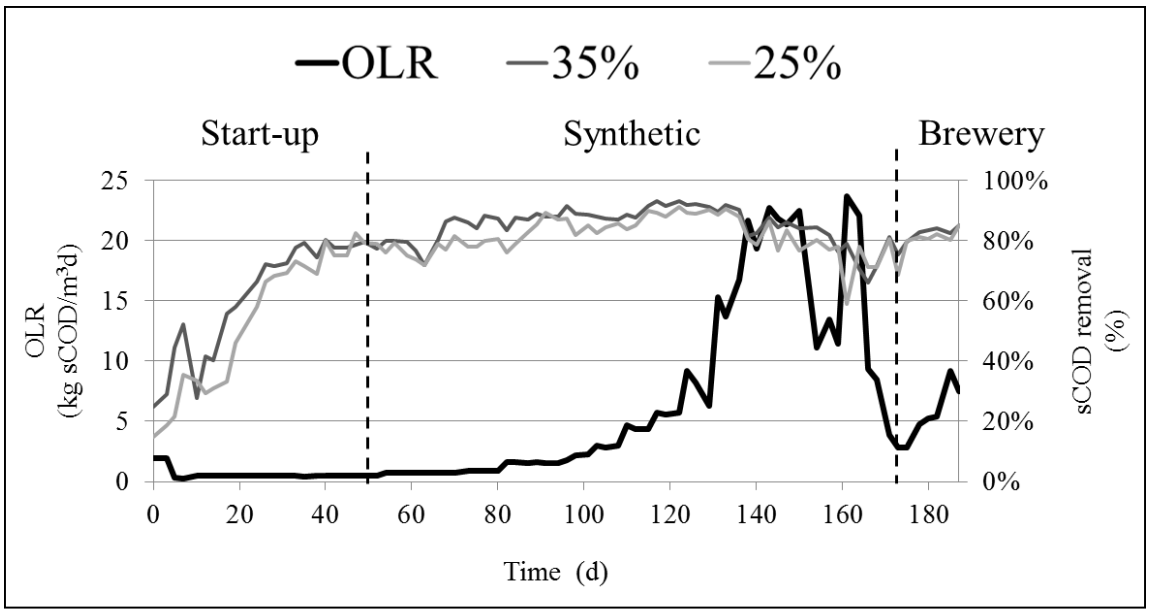


Figure 3-1 - Type of substrate and OLR supplied shown with percent removal in the 25% and 35% media reactors

In particular, the AMBBR technology examined has been used mainly to treat wastewater from wine distilleries and dairies. For instance, Sheli and Moletta (2007, 2010) and Sheli et al. (2014) used winery distillery wastewater to test AMBBR performance at 66 % media content by volume and different types of manufactured media (Bioflow 9 and Bioflow 30). Moreover, a media content of 65 % was utilized by Wang et al. (2009) to treat milk permeate wastewater. The AMBBR in the work reported here achieved comparable degrees of COD removal, from 0.4 to 20 kg sCOD/m^3 d, even though the relative quantity of media was consistently less (25 and 35 %). It should be noted, however, that the SSA (m^2/m^3) and shape (cylindrical vs cubical) were different between the compared media. Therefore, in order to exclude differences in manufacturing, the SSAA was calculated (Equation 1) and compared to results from the other studies. When supplied with synthetic wastewater the maximum specific activity per unit surface area was 98 g sCOD/m^2 d in the 25 % media reactor and 75 g sCOD/m^2 d in the 35 % media reactor. Both were obtained at an OLR above 20 kg sCOD/m^3 d. Lower specific activities were

observed when brewery wastewater was loaded since the OLR was never brought above 10 kg sCOD/m^3 d. In particular, 38 and 26 g sCOD/m^2 d were obtained in the 25 % and 35 % media reactors, respectively. The contribution of suspended growth to the overall removal was quantified through kinetic tests and excluded when calculating specific activity per unit surface area. Results had shown that the suspended biomass contributed 2.5 % of the removal at the maximum OLR in both reactors. When comparing the attached growth efficiency in the 25 % media content reactor it became apparent that the AC920 media performed significantly better in terms of SSAA than all other media, with the exception of Bioflow 9 – Table 1-9. The cubical features and larger surface area of the AC920 in contrast with the cylindrical and smaller shape of the other media may have accounted for the improved support of the anaerobic biofilm. Moreover, the lower media content in both reactors demonstrated that reliable performance can be achieved with lower amounts of media than typical 60 % to 70 % recommended for MBBR systems (Barwal & Chaudhary, 2014).

3.1.3 Volatile Fatty Acid

Volatile fatty acid measurements were performed over the increasing in synthetic influent concentrations (i.e. day 122, 131, 138 and 145) and raw results are shown in Appendix 6.1.2. Figure 3-2 and 3-3 report the total VFA and effluent composition in 25 % and 35 % media volume reactors. Results show a linear increase in total influent VFA as the OLRs were increased from 6 to 15 and finally 22 kg sCOD/m^3 d while at 23 kg sCOD/m^3 d the total influent VFA concentrations decreased slightly. As expected, a similar trend was observed in the effluent VFA composition measured.

The efficiency of AD was determined by quantifying and plotting the VFA/Alkalinity (Figure 3-2 and 3-3). The VFA/Alk is an indicator of the stability of the methanogenesis process. If it is too

high it may upset the process by decreasing the pH to levels in which the methane production will suffer (Ertem, 2011). A study conducted by Mosquera-Corral et al. (2001) reports 0.3 VFA/Alk as the maximum ratio beyond which methanogenesis is unstable. For this reason, the reactors were supplemented with extra alkalinity on top of the 1.5 g $CaCO_3$/L during the start-up period. The total alkalinity added was 2.7, 3.6, 4.7 g $CaCO_3$/L at OLRs of 6, 15, 22 kg sCOD/m^3 d respectively, resulting in an increase of around 1 g $CaCO_3$/L per OLR changing. Hence, the reactors did not reach a critical VFA/Alk ratio value while increasing the influent concentrations to the maximum applied load (i.e., 23 kg sCOD/m^3 d). The maintenance of the buffer capacity and stability of the systems were also demonstrated by the 7.4 - 7.5 average pH values measured over the particular OLRs applied at the time of the VFA measurements. A comparable trend of acetate and propionate was observed in both reactors effluents but the reactor filled with 35 % of AC920 did not show an accumulation of butyrate and isovalerate in the effluent as per the 25 % reactor. The 35 % reactor had a better stability with lower VFA/Alk ratio values compared to the 25 % because of the larger surface area available for biomass formation and therefore a more efficient VFA removal. Figure 3-4 reports a scatter plot in which the VFA/Alk ratios and the OLRs were correlated. The graph compares the VFA/Alk ratio resulting from alkalinity addition (around 1g $CaCO_3$/L per each OLR) to the VFA/Alk ratio that results without any alkalinity addition (2.7 g $CaCO_3$/L as per OLR of 6 kg COD/m^3 d).

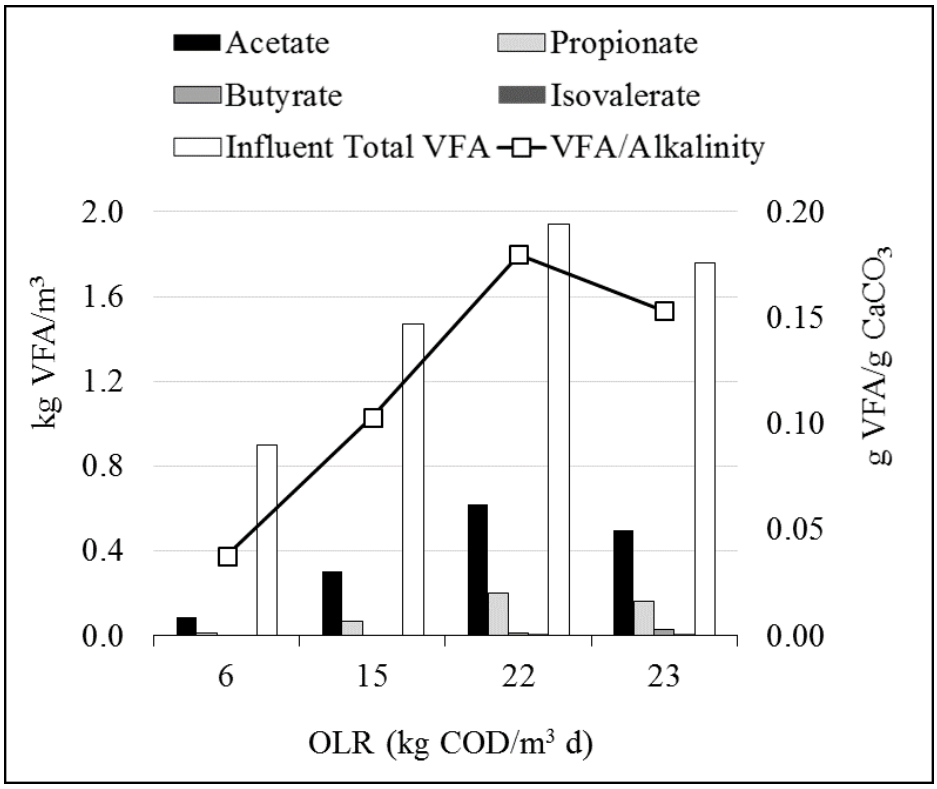


Figure 3-2 – Total influent VFA and effluent VFA speciation of the 25 % reactor

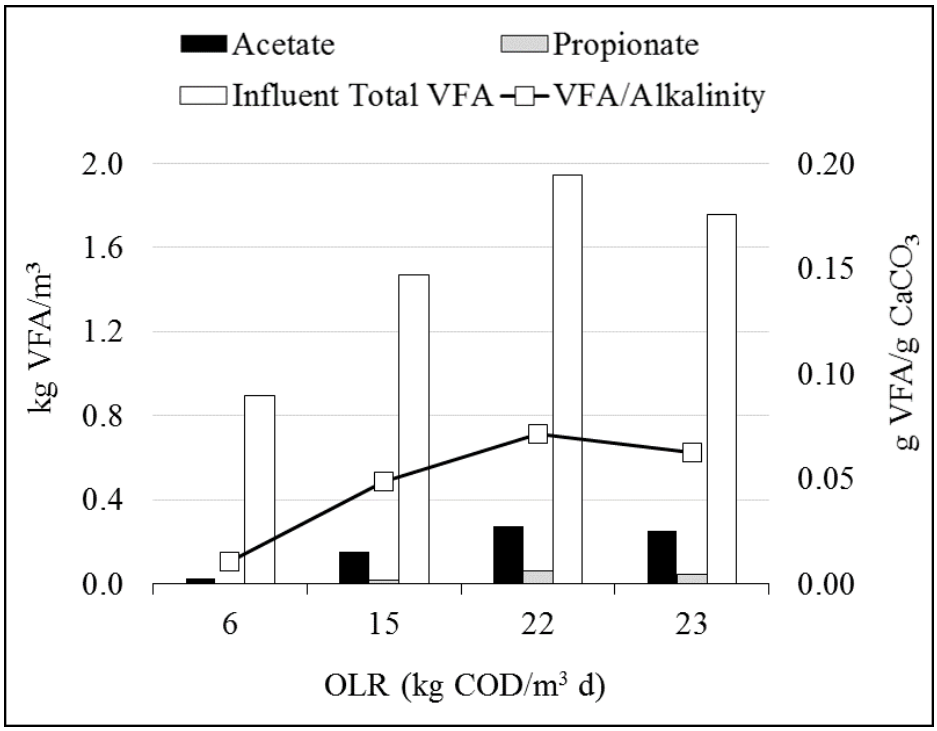


Figure 3-3 – Influent total VFA and effluent VFA speciation of the 35 % reactor

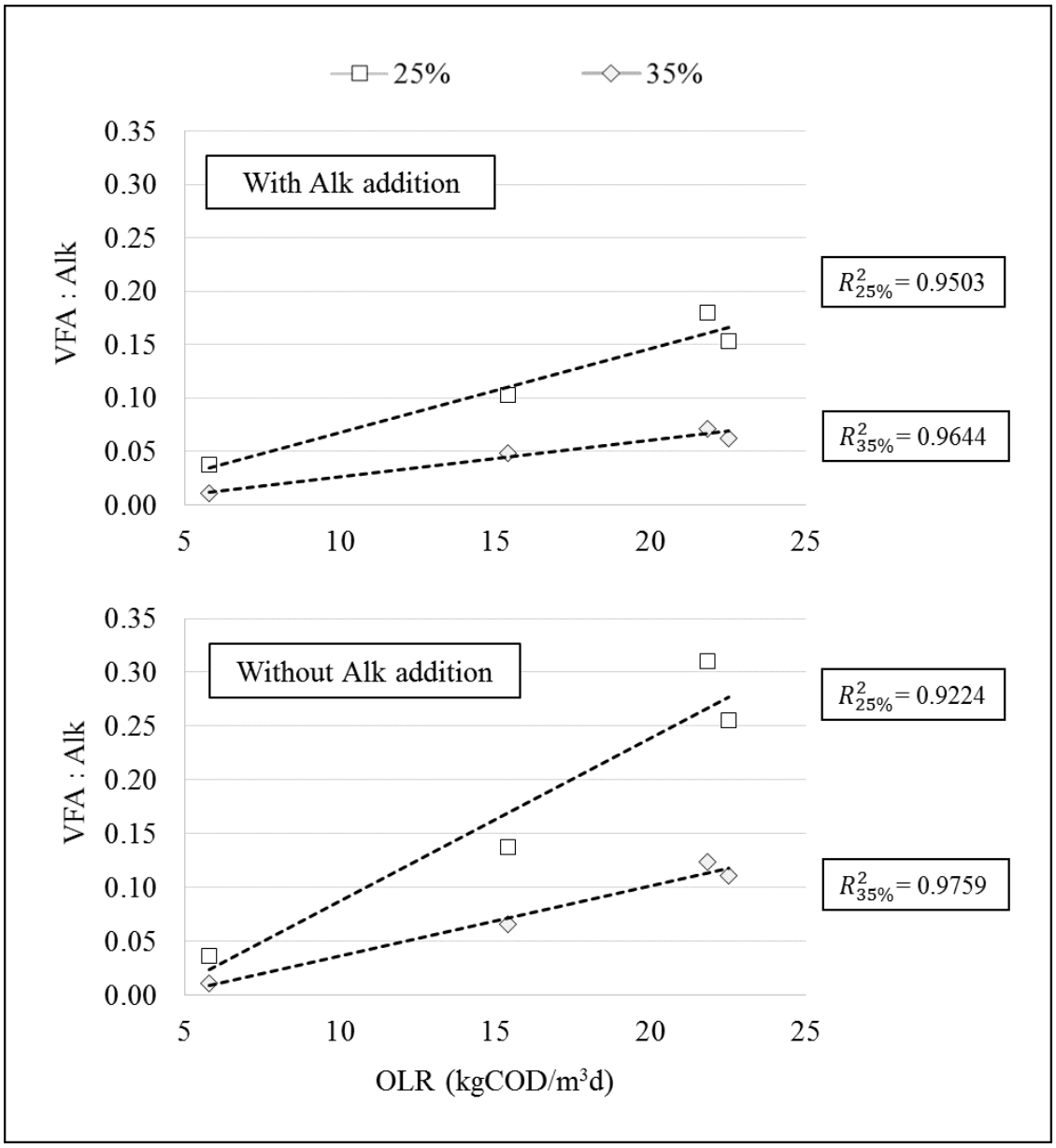


Figure 3-4 – OLR and VFA:Alk ratio correlations of both 25 % and 35 % reactors with (on top) and without (on the bottom) alkalinity addition

A strong correlation was found with R^2 greater than 0.9503 for all the data sets, while the curve for the 25 % reactor without any alkalinity addition had a lower linear regression coefficient (R^2 = 0.9224). Without alkalinity addition, the increase of OLR above 20 kg COD/m^3 d would likely have a negative impact on methanogenesis reaching higher values than the critical 0.3 VFA/Alk ratio.

3.1.4 Biogas production and characterization

Total biogas production and biogas composition were analysed throughout the study raw data are reported in Appendix 6.1.3. Data collection had begun when the reactors were determined to be performing at a steady state based on sCOD removal. Therefore, during the period in which loading rates were increased, it was possible to measure and record the biogas composition and the total biogas production. Gas chromatography showed an overall gas composition of 60 to 80 % CH_4, 25 to 30 % CO_2 and between 1 to 9 % N_2 during both synthetic and brewery feeding periods – Figure 3-5.

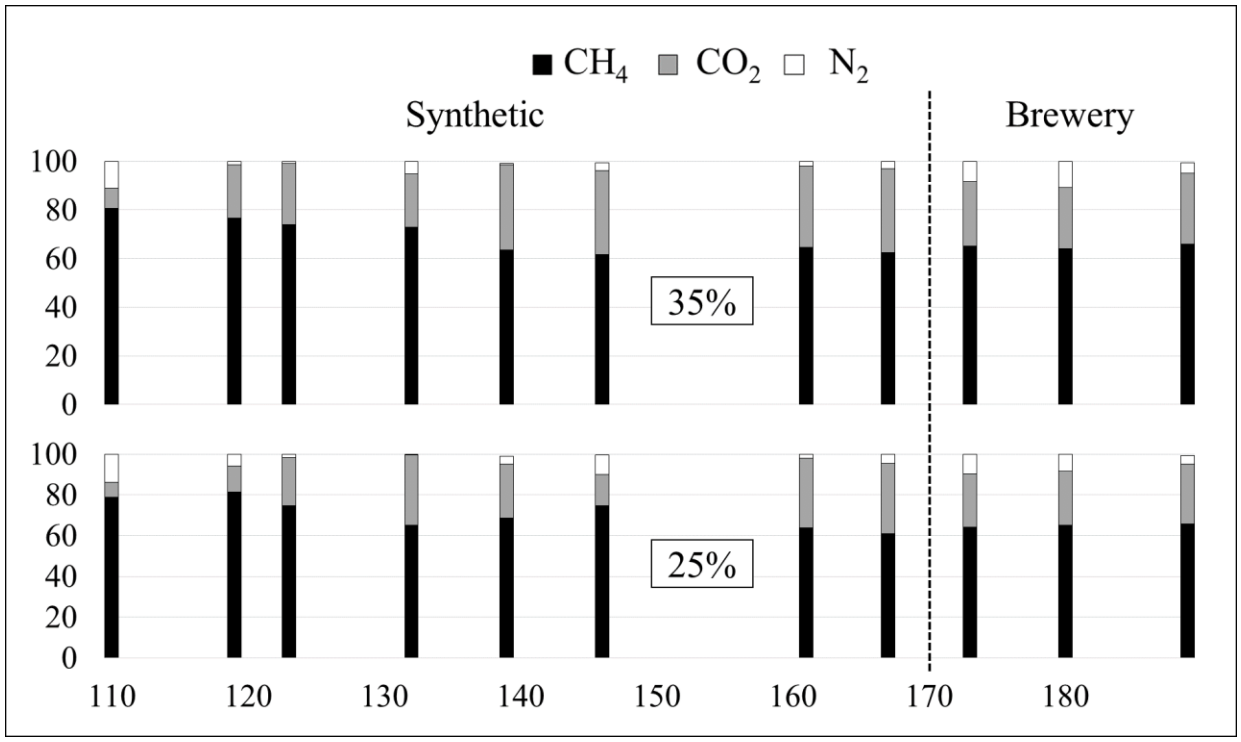


Figure 3-5 – Biogas composition on synthetic and brewery substrate in 25 % and 35 % reactors

The carbon dioxide yield, however, showed more fluctuation over time. Measurements revealed that during the synthetic feeding period the carbon dioxide content of the biogas was between 7 and 35 %. When the feed was switched from the synthetic to brewery wastewater, the biogas

produced had a more stable content of methane, between 64 and 66 % over time. Less fluctuation in carbon dioxide content was also observed treating brewery wastewater, within the range of 25 to 30 %.

With the biogas composition quantified, it was possible to plot both total biogas and methane production as a function of OLR (Figure 3-6). Both systems showed similar trends in total biogas production, which varied from 1 to 9 $m^3 CH_4/m^3$ d. The trend in total biogas production as a function of OLR was found to be comparable during synthetic and brewery wastewater treatment. This is supported by a strong linear correlation coefficient from the data sets for both total biogas ($R^2 = 0.9510$) and methane production ($R^2 = 0.9418$) observed for synthetic and brewery wastewater applications.

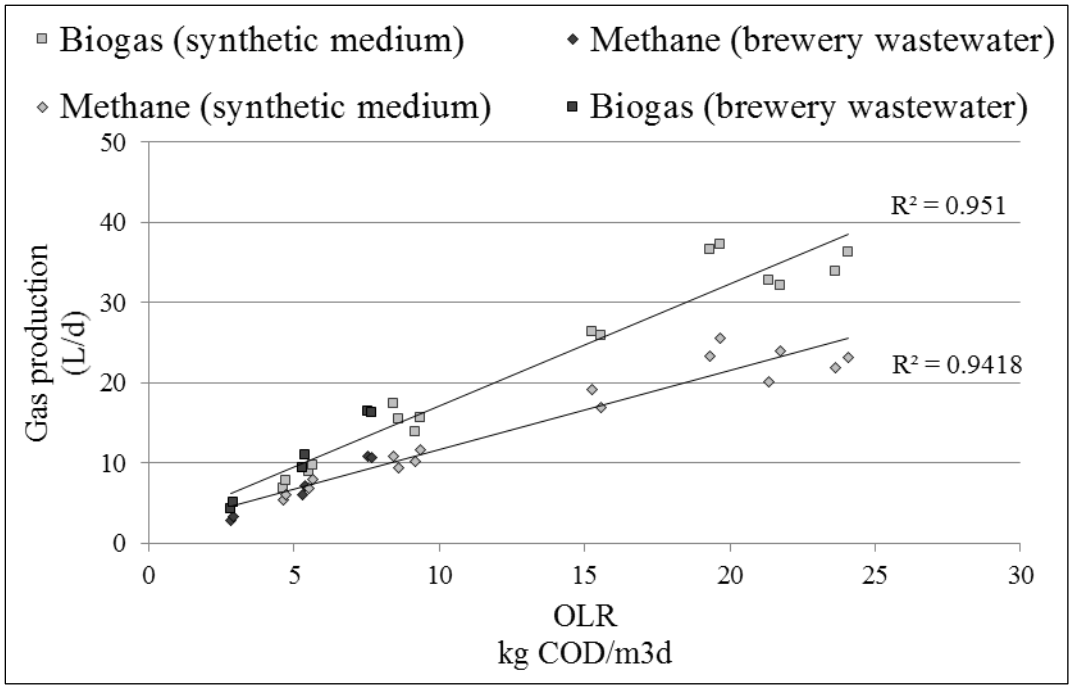


Figure 3-6 - Total biogas and methane production related to the OLR

From these results, it was possible to calculate methane yields of 0.36 ± 0.06 and 0.39 ± 0.03 $m^3 CH_4$/kg sCOD in the 25 and 35 % reactor, respectively, according to Equation 3. These values

obtained on synthetic industrial wastewater were close to the stoichiometric value of 0.40 m^3 CH_4/kg COD under mesophilic conditions between 30 - 38°C (Metcalf & Eddy, 2013).

3.2 Organic loading rate increasing via decreased HRT

3.2.1 Influent and effluent characteristics

The performance measures for the AMBBR influent and effluent for the entire study is presented in Table 3-1 and reported in Appendix 6.2.1.

Table 3-1 – AMMBR influent and effluent characteristics for the entire study

	Parameter	Minimum	Maximum	Average	Standard Deviation	Unit
Influent	pH	6.4	7.1	6.7	0.1	-
	TSS	0.1	3.6	0.5	0.6	kg TSS/m^3
	VSS	0.1	2.3	0.4	0.4	kg VSS/m^3
	sCOD	2.6	5.9	3.8	1.0	kg sCOD/m^3
	TCOD	2.8	9.2	5.0	1.6	kg TCOD/m^3
	BOD_5	2.1	7.1	3.8	1.2	kg BOD_5/m^3
Effluent	pH	6.7	7.3	7.0	0.1	-
	Alkalinity	1.3	2.0	1.6	0.1	kg $CaCO_3$/m^3
	TSS	0.1	0.8	0.3	0.2	kg TSS/m^3
	VSS	0.1	0.8	0.3	0.1	kg VSS/m^3
	sCOD	0.1	1.8	0.7	0.5	kg sCOD/m^3
	TCOD	0.3	2.6	1.1	0.6	kg TCOD/m^3
	BOD_5	0.2	2.0	0.9	0.5	kg BOD_5/m^3

Results presented neglecting the first week after a particular change of HRT was made.

Only the last two weeks of data for each interval of HRT were considered in the calculations to exclude the period of acclimation to higher loadings. Alkalinity concentration in the influent was initially supplied as 1.5 kg $CaCO_3$/m^3, but was increased to 2.0 kg $CaCO_3$/m^3 when the HRT was 6 h because a drop in pH to 6.5 was observed. The pH was maintained near neutral values for the entire study, and therefore non-acclimated methanogenic microorganisms should not have been

inhibited by acidic conditions (Taconi, Zappi, Todd French, & Brown, 2008). Effluent pH averaged 7.0 ± 0.1 while influent pH was 6.74 ± 0.05, demonstrating that the buffering capacity was capable of preventing acidification. Influent TSS and VSS were highly variable, with TSS ranging from 0.1 to 3.6 kg TSS/m^3. This was most likely due to fluctuating quantities of yeast and spent grains in the waste samples from the brewery. The AMBBR was able to reliably remove significant quantities of soluble and total COD in spite of inconsistent solids loading. Usually wastewater is treated with microscreens prior to MBBR processes such that solids do not interfere with removal rates (Josse, Theodoulou, & Scattergood, 2013). This suggests that the AMBBR is a robust system even though its biomass comprises slow growing microorganisms. Percent removal of soluble and total COD (Appendix 6.2.2) decreased as organic loading rate (OLR) increased – Figure 3-7.

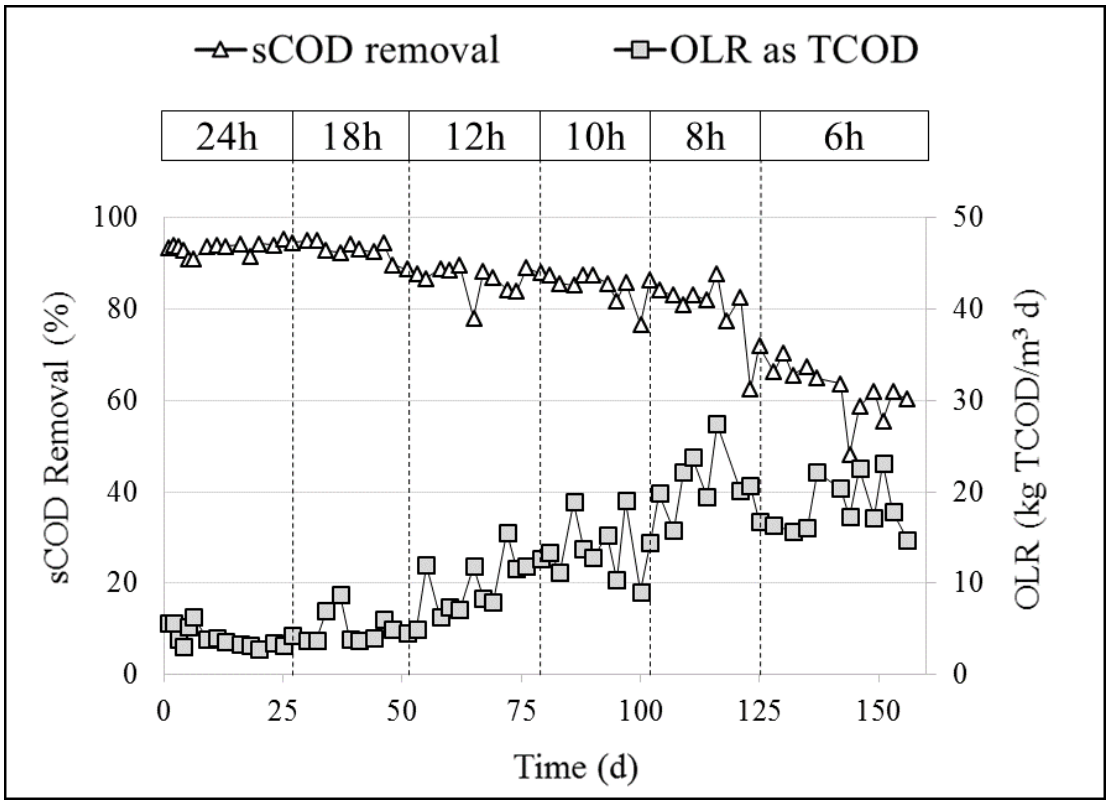


Figure 3-7 - Averaged soluble COD removal shown with averaged total OLR during the study. The different intervals of HRT are presented as blocks at the top of each graph

At an HRT of 24 h, which corresponds to an OLR of 3.8 kg TCOD/m^3 d, the AMBBR removed 93 % of sCOD and 84 % of TCOD. An 18 h HRT resulted in comparable COD removal even though the OLR was 5.5 kg TCOD/m^3 d. Performance was observed to decrease at each following interval of HRT, with sCOD removal recorded as 86 % during the 12 h HRT. At the shortest HRT applied, which was 6 h, sCOD removal was the lowest recorded at 60.8 % even though the OLR (18.7 kg TCOD/m^3 d) was actually lower than at an 8 h HRT. The discrepancy in OLR resulted from an error in feed preparation, where the feed was diluted more than required. This decreased the concentration of COD in the feed and thus the OLR decreased in spite of the hydraulic load increasing. The overall trend in performance demonstrated that 80 % sCOD removal was reliably attained until HRT decreased below 10 h. This suggested that surface area loading rates should not exceed 56 g TCOD/m^2 d to achieve 80 % removal of sCOD, given 40 % media filling fraction and a specific protected surface area of 680 m^2/m^3. A similar calculation can be made if 90 % removal of sCOD is desired since the AMBBR reliably removed more than 90 % of sCOD at HRT greater than or equal to 18 h. In this case the surface area loading rate should not exceed 24 g TCOD/m^2 d. Total COD removal rates, on the other hand, were more variable and ranged from 59 to 93 % during the study. This was attributed to the fluctuating solids concentration in the influent. Several studies on AMBBR processes treating industrial wastewater are summarized in Table 3-2.

Studies on winery wastewaters (Sheli and Moletta, 2010) demonstrated that to achieve 80 % sCOD removal at surface area loading rate (SALR) of 87 g sCOD/m^2 d using R9 media with an SAA of 320 m^2/m^3. On the other hand, Wang et al. (2009) investigated the capability of same R9 media treating a different substrate. The authors showed that an SALR of 58 g sCOD/m^2 d removing 85 % sCOD was achieved in treating dairy wastewater. Treating a more complex

substrate may have had a negative impact on the R9 media. The organic carbon present in the wine distillery wastewater mainly contains yeast and it is easier to breakdown than the milk permeate. The anaerobic biodegradability rates of fat-rich wastewaters is slower than the low fat ones because of the slower hydrolysis of the fat material (Vidal, Carvalho, Méndez, & Lema, 2000). Additionally the higher content in particulate material may hinder the properties of the R9 media and therefore limit its optimal mass transfer (Demirel, Yenigun, & Onay, 2005).

Table 3-2 – Review of AMBBR processes treating industrial wastewater

Wastewater treated	Carrier		sCOD rem (%)	SALR (g sCOD/m^2 d)	Reference
	Type	SAA (m^2/m^3)			
Wine distillery	R30*	320	85	87	Sheli et al. (2010)
Wine distillery	R9*	530	80	85	Sheli et al. (2014)
Dairy	R9*	530	85	58	Wang et al. (2009)
Brewery	AC920	680	70	87	This study
			80	56	
			85	40	
			90	24	

* Bioflow manufactured by Raushert Co.

Brewery wastewater composition is more comparable to the winery than dairy wastewater since both brewery and winery processes have similar production (e.g., fermenters), utilization of yeast, lower particulate content, and low fat characteristics. The yeast is the highest source of COD in these wastewaters. By comparing this study on the AC920 manufactured by Headworks Bio with the R9, different SALR values were obtained for similar percentage of sCOD removal. Particularly at 85 % sCOD removal, the AC920 media were able to achieve less than half the SALR than the R30. Similarly, approximately 30 g sCOD/m^2 d of difference in SALR were observed for 80 % sCOD removal. It was expected that the SALR was greater for the AC920

media compared to the others presented because of its physical properties. The higher SAA of the media used in this study (i.e., 680 m^2/m^3) was expected to reach a greater loading rate than the one obtained for 320 m^2/m^3.

The AC920 removal property was limited most likely because of the cubical shape of the media and small spacing between the different compartments of the media. Hence, mass transfer limitation was found to be the cause of unexpectedly decreased performance due to both the shape of the media and the mixing properties linked with the cubical feature in the small bench scale reactors. Increasing the mixing, and therefore the shear, had little impact on the mass transfer limitation observed.

3.2.2 Biomass activity

Biomass activity was quantified by specific surface area activity (SSAA; Equation 1), raw data are reported in Appendix 6.2.2. For the most part, SSAA increased with OLR as the HRT decreased – Figure 3-8. At an HRT of 24 h the SSAA was determined to be 11 ± 1 g TCOD/m^2 d, while at an 18 h HRT the SSAA was 13 ± 1 g sCOD/m^2 d. When the HRT decreased to 12 h the SSAA was calculated as 24 ± 1 g sCOD/m^2 d. The highest SSAA, 44 ± 2 g sCOD/m^2 d, was observed at a HRT of 10. Unexpectedly, the SSAA actually decreased to 32 ± 1 g sCOD/m^2 d when the HRT was changed to 6 h. Limitations in mass transfer that resulted in upset were most likely caused by high OLR since the biofilm could have outgrown the protected surface area provided by the media. The SSAA quantified for total COD followed similar trends compared to the ones for sCOD, although with higher fluctuation. This was expected since the influent solids were observed to vary significantly.

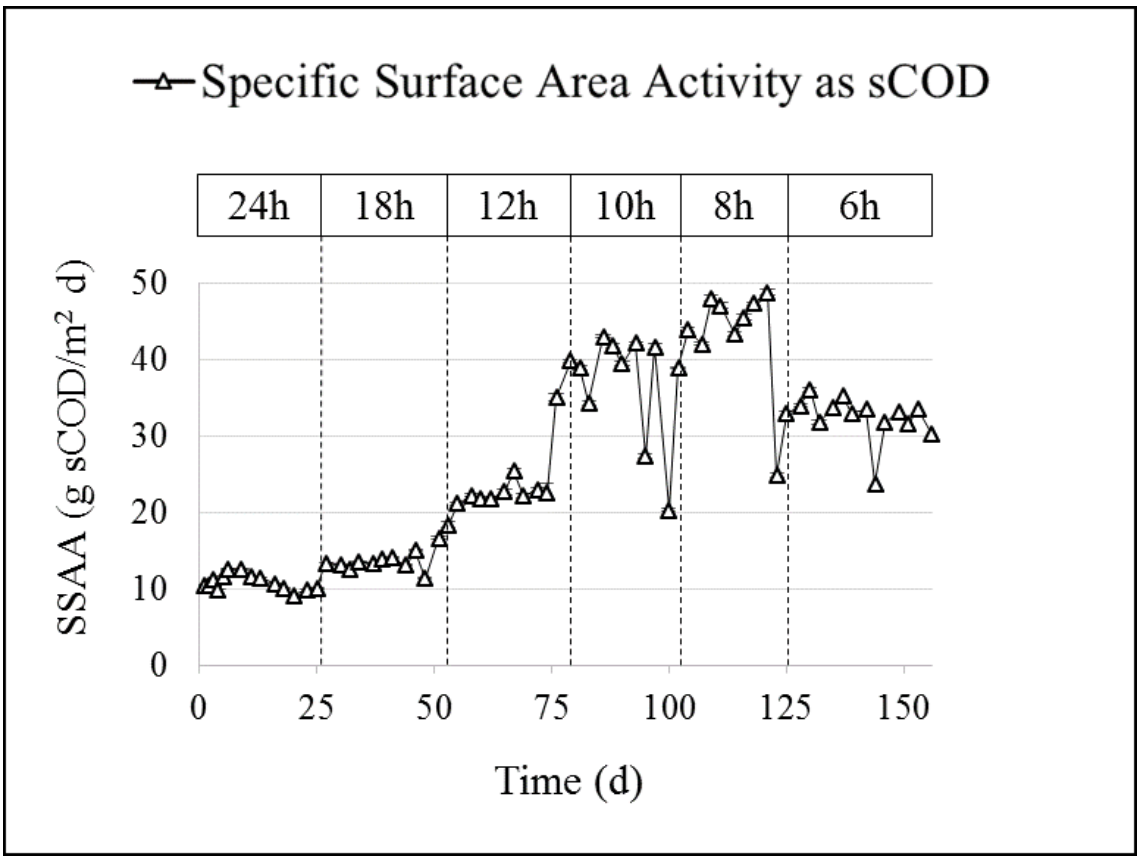


Figure 3-8 - Specific surface area activity (SSAA) measured on soluble COD at different HRTs

Kinetic tests were performed on the reactor liquor, without media, before each change in HRT (Appendix 6.2.3) to estimate the contribution of suspended biomass to the overall performance – Figure 3-9. The suspended biomass contribution was determined to be insignificant compared to the actual removal observed in the AMBBR. At a 24 h HRT the contribution of suspended biomass was approximately 0.4 % of the total COD removed, which suggested that activity was almost completely carried out by the established biofilm and the solids were spent inactive biomass. A positive linear correlation was observed between the suspended biomass activity and the applied OLR as the HRT decreased from 24 to 10 h. Estimations of the suspended biomass contribution increased from 0.4 % at 24 h to 2.6 %, 3.1 % and 5.4 % at 18, 12, 10 h HRT, respectively. The contribution of suspended biomass was observed to decrease between 10 to 6 h

HRT. Suspended biomass contribution was estimated to be 2.2 % and 0.4 % at 8 and 6 h HRT, respectively.

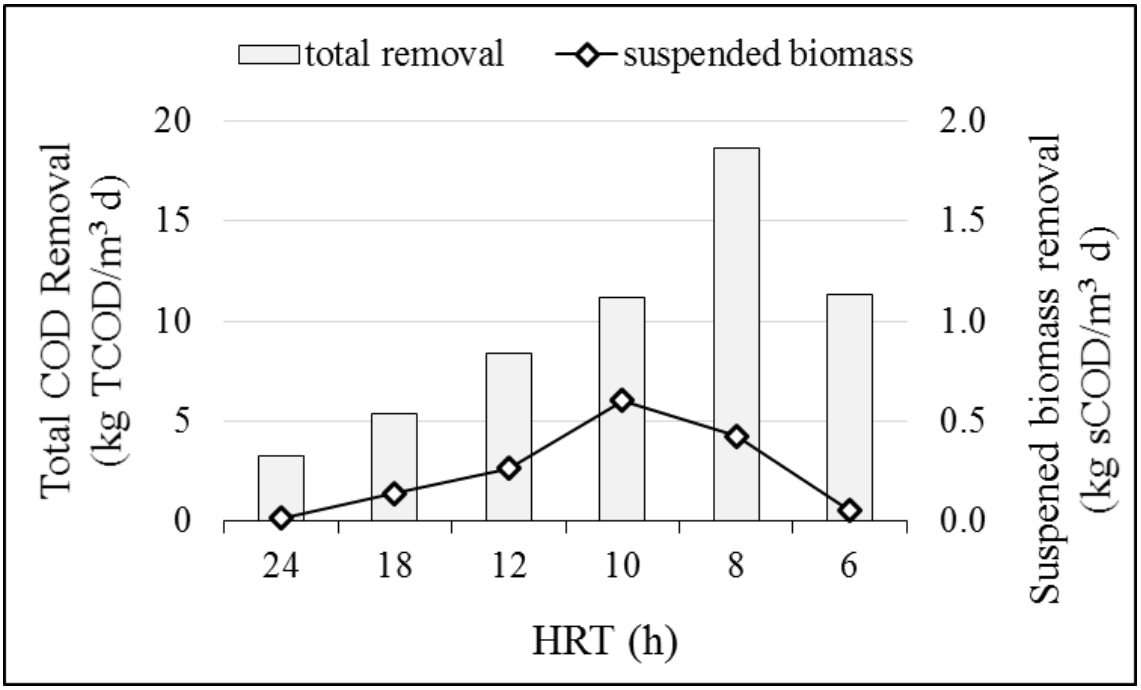


Figure 3-9 - Averaged total COD removal compared with removal by detached suspended biomass alone, at different HRT

With decreasing HRT, and thus increasing OLR, it is most likely that the biofilm experienced increased sloughing resulting in higher concentrations of active suspended solids. This would explain the increase in suspended biomass activity from 24 to 10 h HRT but not the decrease observed in suspended activity at 8 and 6 h HRT. It can be argued that at lower HRT, the SRT of suspended biomass would be too low for maintaining an active population. However, at higher OLR the increased biofilm production would augment the solids lost via the effluent. A more likely explanation for the observed decrease at 8 and 6 h HRT is that the void volume available for biofilm grown on the protected surface area became saturated with biomass, resulting in mass transfer limitations. This would result in less surface area available for biomass sloughing and also explain the significant drop in overall COD removal at a 6 h HRT, since substrate would

also have less surface area to penetrate the biofilm – Figure 3-10. This explanation is justified by physical observations of the biofilm.

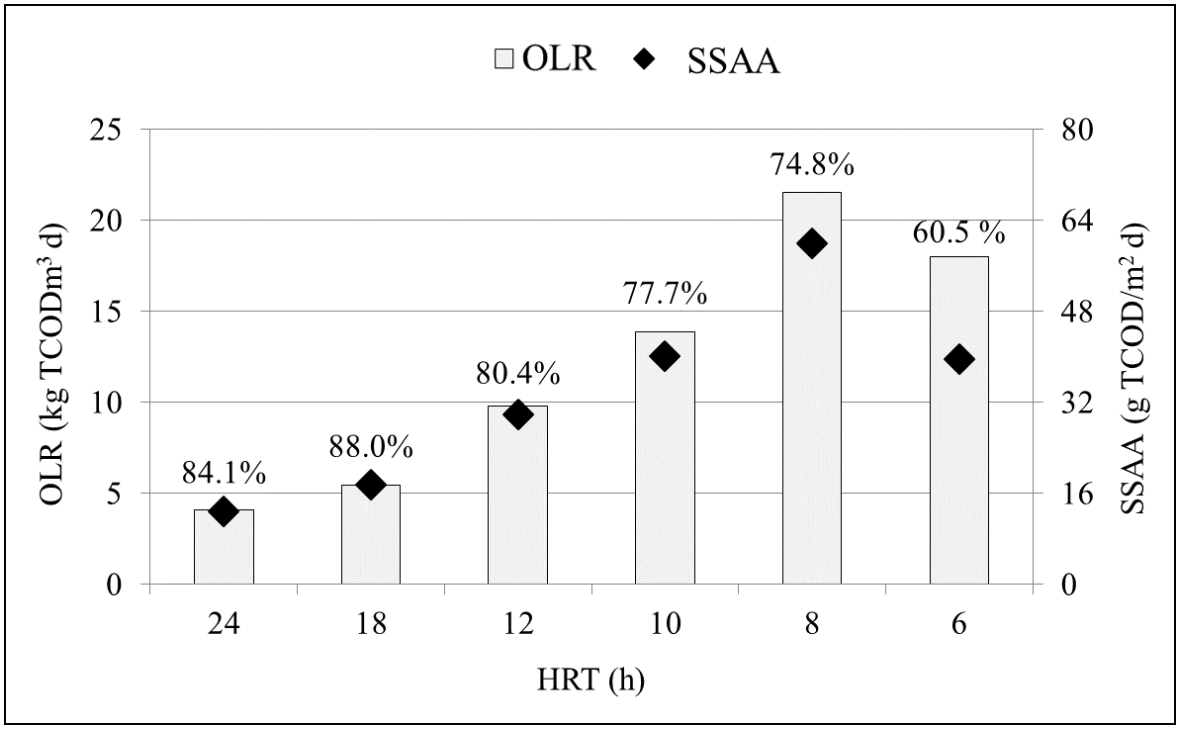


Figure 3-10 - Total organic loading rate (OLR) and specific surface area activity (SSAA) obtained at different HRT. Percentages represent the load of COD removed relative to the oncoming load

Specific surface area activity (Equation 1) increased as expected with the changes in loading rates at different HRT – Figure 3-10. However, the SSAA decreased significantly between 8 and 6 h HRT, which supports the previous claim that mass transfer limitations were brought about by the biofilm outgrowing the available protected surface area. The AC920 media were cubic in shape and divided into 9 square tunnels within which the biofilm can establish. Each square tunnel was approximately 4 by 4 mm, and thus the biofilm thickness should never reach 2 mm. Based on SSAA alone, it can be concluded that a SALR of 54 g TCOD/m^2d should not be exceeded for the AC920. However, suspended biomass kinetics suggested that mass transfer may have been limiting even at the 8 h HRT, and thus the SALR should not exceed 44 g TCOD/m^2d.

3.2.3 Biogas production and characterization

Biogas production and composition were monitored for the entire study. Raw data are reported in Appendix 6.2.4. The biogas compositions remained relatively consistent between 24h and 10h HRT, with 67 ± 3 % average methane and 23.0 ± 0.5 % carbon dioxide – Figure 3-11. Overall, percent carbon dioxide was observed to fluctuate slightly more than percent methane. The methane to carbon dioxide ratio presented in Figure 3-11, reviled that at HRT ranging from 24h to 10h the ratio was comparable. The CH_4:CO_2 ratio at these HRTs was around 3.0, indicating that fermentation and methanogenesis were balanced and stable. However, the ratio decreases to 2.3 ± 0.2 CH_4:CO_2 when the HRT was brought to 8h and at the lowest HRT of 6h reached 2.0 ± 0.2 CH_4:CO_2. These drop down suggested that stability of anaerobic digestion process was compromised and the fermentation process was occurring faster than methanogenesis. Calculations for unit methane generation (Equation 3), reported in Appendix 6.2.5, were performed using the OLR, biogas production, and percent methane. Unit methane generation was more stable when sCOD was considered, which was most likely due to fluctuating solids concentration in the feed. There was a slightly decreasing trend observed in unit methane generation with OLR. Overall, the unit methane generation did not exceed the theoretical value of 0.4 m^3 CH_4/kg sCOD (Viana, Freitas, Leitão, Pinto, & Santaella, 2012).

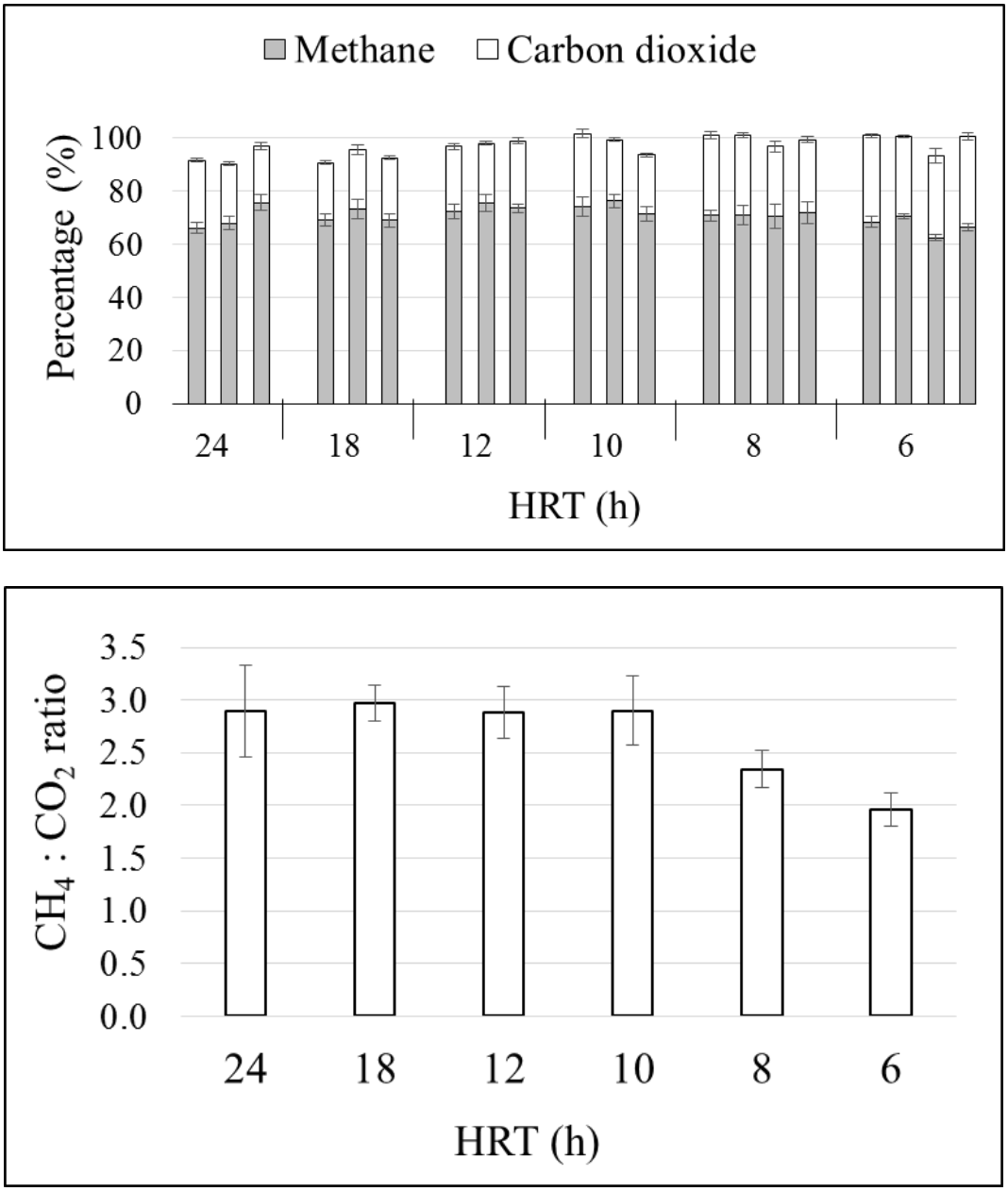


Figure 3-11 - Biogas composition as percent methane and carbon dioxide (on top). Methane and carbon dioxide ratio (on the bottom). Error bars represent one standard deviation between triplicates

CHAPTER 4 – ENGINEERING SIGNIFICANCE

4.1 Design considerations

The North American Brewery Association reports in its "Water and wastewater treatment and volume reduction manual" (Chastain et al., 2011) the annual beer production for different brewery size and the wastewater generate per unit of beer produced. This ratio varies depending on the policy adopted in the beer production chain and the effectiveness of the environmental management system if implemented. The manual reports different wastewater generation to beer produced ratios (i.e., 2:1, 4:1 and 10:1) typical of North American territories. Table 4-1 shows the wastewater generation rates per unit beer produced.

Table 4-1 – Beer production and wastewater generation rates

Size	Beer Production (m^3/d)	Wastewater : beer produced (m^3/d)		
		2 : 1	**4 : 1**	**10 : 1**
Small	3	5	11	27
Medium	137	274	548	1370
Large	411	822	1644	4110

North American brewery wastewater concentrations are reported to be equal to 1.8 kg COD/m^3, 3.65 kg COD/m^3 and 5.5 kg COD/m^3 as minimum, average and high value, respectively (Chastain et al., 2011). The three different concentrations were assumed to be representative of 2:1, 4:1 and 10:1 wastewater/beer production ratio. The breweries generating twice the wastewater rate than the beer produced were therefore assumed to have a higher COD concentration (i.e., 5.5 kg COD/m^3) while 4:1 was taken as the average and 10:1 as lower concentrations. Hence, it was possible to develop the COD loading rates of wastewater produced per day – Table 4-2.

Table 4-2 – Mass of COD produced in different breweries size

Size	Wastewater : beer produced (kg COD/d)		
	2 : 1	**4 : 1**	**10 : 1**
Small	30	40	49
Medium	1507	2000	2466
Large	4521	6000	7397

The small size breweries have a difference of around 10 kg COD/d as the ratio increases while approximately 500 kg COD/d is the difference in COD mass between 2:1, 4:1 and 10:1 wastewater/beer produced by a typical medium size. Large breweries however have a gap of 1400 kg COD/d as the ratio increases within the three considered.

The thesis presented on the AMBBR technology allowed developing design parameters for brewery wastewater application using the media manufactured by Headworks BIO. The SALR at particular sCOD removal is a key parameter to be estimated. Soluble COD was used for design since a microscreen will most likely be implemented before the AMBBR in full scale applications (Josse et al., 2013). An example of the equation used to quantify the SALR at different sCOD removal (e.g., 80 %) throughout the experimental data is presented in Equation 4.

$$\mathrm{SALR_{80\%}\ (g\ sCOD/m^2\ d)} = \frac{\mathrm{OLR_{exp}\ (g\ sCOD/}m^3\ \mathrm{d}) * \mathrm{V_R}(m^3)}{\mathrm{TSA_R\ (m^2)}} \qquad \text{(Equation 4)}$$

Where the OLR_{exp} represent the averaged OLR provided in the experimentation in which 80 % removal was achieved expressed in kg sCOD/m^3 d, and TSA_R is the total protected surface area (i.e., 1.008 m^2) in the 0.004 m^3 volume reactor (V_R) tested. Surface area loading rates were therefore calculated for each day analyzed at the OLR and sCOD percentage removal obtained in that particular day as reported in Appendix 6.3.1. A correlation curve of the data set obtained

from the HRT study was therefore developed and from the linear regression equation the SALR at a particular percentage of sCOD removal was extrapolated – Figure 4-1.

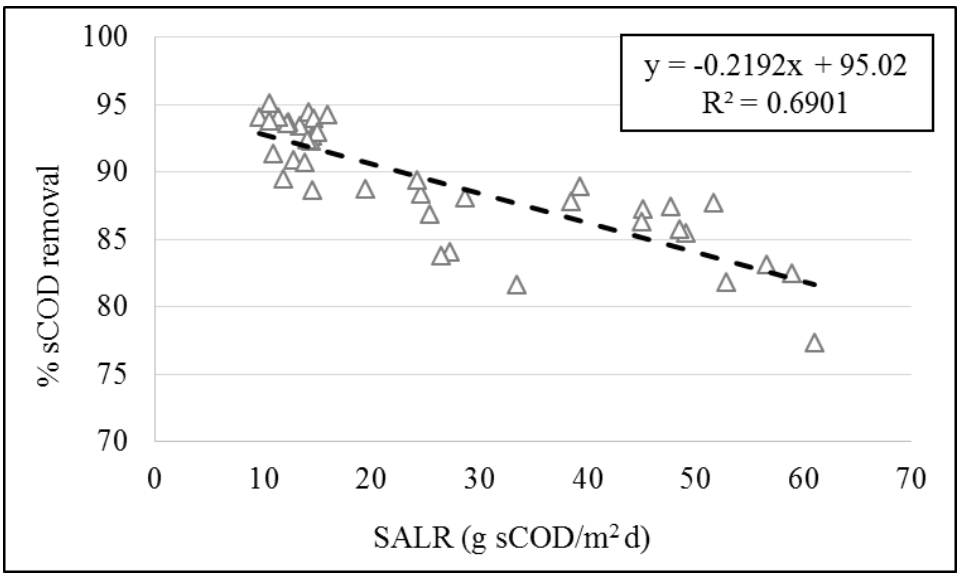


Figure 4-1 – Surface area loading rate and percentage removal of sCOD correlation from experimental data

The linear correlation curve was developed neglecting the results obtained from 6 h HRT. These results were not utilized because of the available protected surface area on the media was found to be saturated with biomass and therefore mass transfer limitations phenomena were observed. A good correlation, $R^2 = 0.6901$, for the data set allowed the extrapolation of the equation for different SALR for design considerations. Table 4-3 present the different SALR calculated from experimental data at particular sCOD removal (e.g., 80 %, 85 %, 90 %).

Table 4-3 – Surface area loading rates at particular sCOD percentage removal

sCOD % removal	SALR (g sCOD m^3/d)
80	69
85	46
90	23

The surface area loading rates reported in Table 4-3 allowed the estimation of the total surface area required (TSA_{req}) to achieve the desired COD removal by using the Equation 5.

$$\mathrm{TSA}_{req}(m^2) = \frac{\text{Organic Mass (kg sCOD/d)}}{\text{SALR (kg sCOD/m}^2\text{ d)}} \qquad \text{(Equation 5)}$$

Where the OLR is the organic loading rate of a particular brewery in kg sCOD/d and the SALR is the surface area loading rate (kg sCOD/m^2 d) at a targeted percentage of sCOD removal. Small, medium and large brewery loading rates were used to define the TSA required to achieve two different percentages of COD removal (i.e., 80 % and 90 %). The different values estimated for each brewery size and different wastewater to beer ratios are reported in Table 4-4.

Table 4-4 – Total surface area required to achieve COD 80 and 90 % removals

	> 80 COD rem			**> 90 COD rem**		
	Wastewater : beer production			**Wastewater : beer production**		
Brewery size	**2 : 1**	**4 : 1**	**10 : 1**	**2 : 1**	**4 : 1**	**10 : 1**
Small (m^2)	542	719	887	1245	1653	2038
Medium (m^2)	27102	35971	44348	62267	82645	101891
Large (m^2)	81305	107914	133044	186800	247934	305672

Different media were taken into consideration with a protected surface area ranging between 402 to 680 m^2/m^3 to compare media volume requirements for different shape and manufacture. In particular, the media chosen for the design were the AC450 (402 m^2/m^3), AC515 (485 m^2/m^3) used in the thesis experimentation and the AC920 (680 m^2/m^3) developed by Headworks Bio. The volume of media required to achieve the desire removal rates was therefore estimated for each type of media, targeted removal, brewery size and different wastewater production. The equation used to develop the volume requirements is reported in Equation 6.

$$\mathrm{V_M}(\mathrm{m^3}) = \frac{\mathrm{TSA_{req}}\ (\mathrm{m^2})}{\mathrm{SSA_{K1}}\ (m^2/m^3)} \qquad \text{(Equation 6)}$$

Where V_M is the volume media required in m^3, TSA_{req} is the total surface area required in m^2 (Table 4-4) and SSA is the specific surface area of the particular media (e.g., 680 m^2/m^3). The

volume of media required for the different type of supports selected is reported in the following table – Table 4-5.

Table 4-5 – Media volume (m^3) required for the three types of media selected

Size	Wastewater per beer ratio	> 80 COD rem			> 90 COD rem		
		AC450	AC515	AC920	AC450	AC515	AC920
Small	2 : 1	1.3	1.1	0.8	3.1	2.6	1.8
	4 : 1	1.8	1.5	1.1	4.1	3.4	2.4
	10 : 1	2.2	1.8	1.3	5.1	4.2	3.0
Medium	2 : 1	67	56	39.9	155	128	92
	4 : 1	89	74	52.9	206	170	122
	10 : 1	110	91	65.2	253	210	150
Large	2 : 1	202	168	119.6	465	385	275
	4 : 1	268	223	158.7	617	511	365
	10 : 1	331	274	195.7	760	630	450

The anaerobic moving-bed biofilm reactor volume is strictly dependent on the amount of available surface area for biomass development. From the table above it can be seen how the smaller is the surface area of the media (e.g. 402 m^2/m^3 for the AC450) the higher is the volume required to achieve the desired removal. In addition, the media filling volume plays a consistent role in the reactor volume requirement, depending on the percentage chosen (e.g., 30 %, 50 % or 70 %). Therefore, a comparison of reactor volume requirements for brewery wastewater treatment to achieve 80 % and 90 % COD removal was evaluated choosing media filling of 30 %, 50 %, and 70 % - Figure 4-2 and 4-3. Appendix 6.3.2 reports the raw data on which the figures were developed.

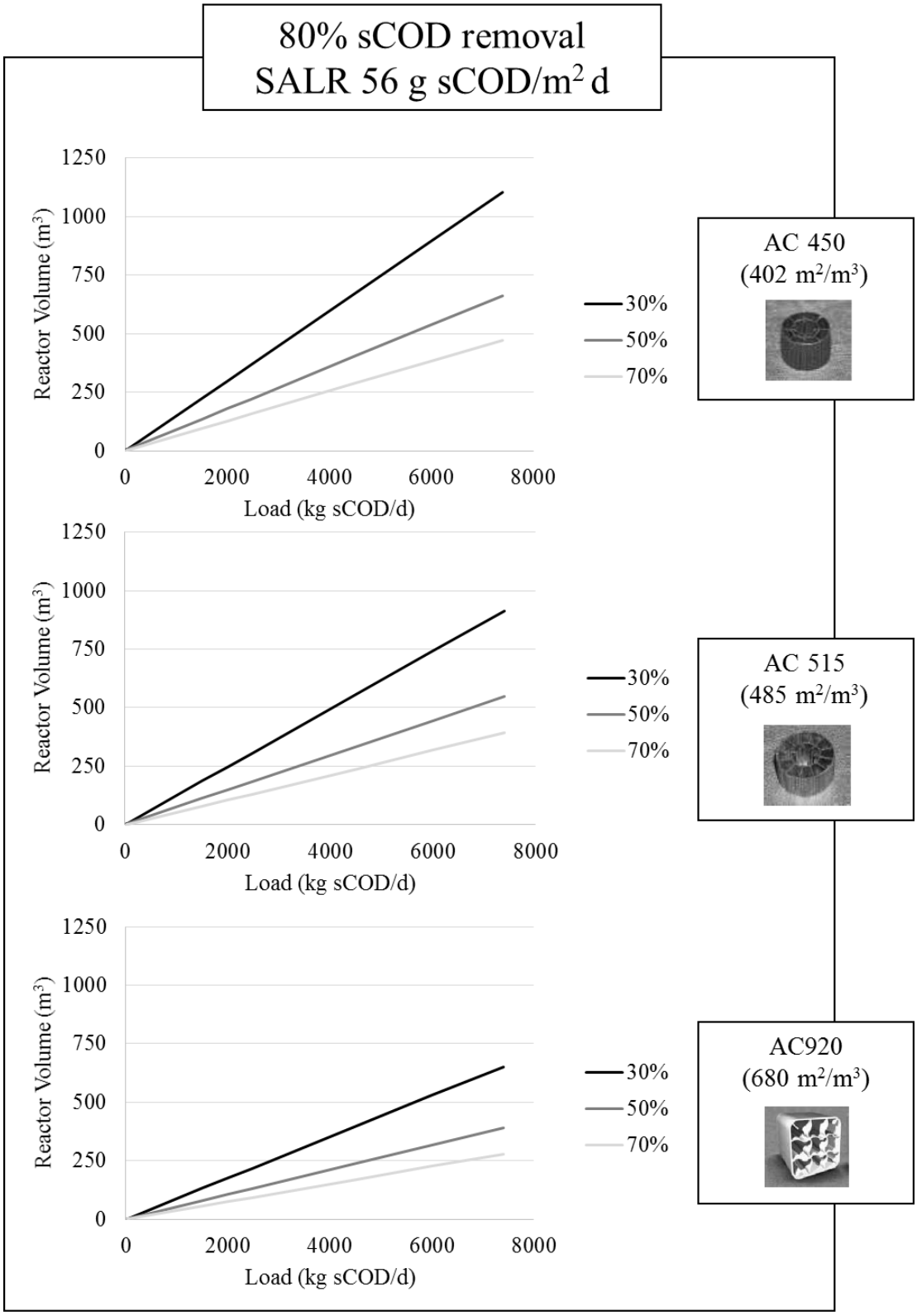


Figure 4-2 – Comparison of reactor volume requirement to achieve 80% removal by implementation of three media (i.e., AC450, AC515 and AC920)

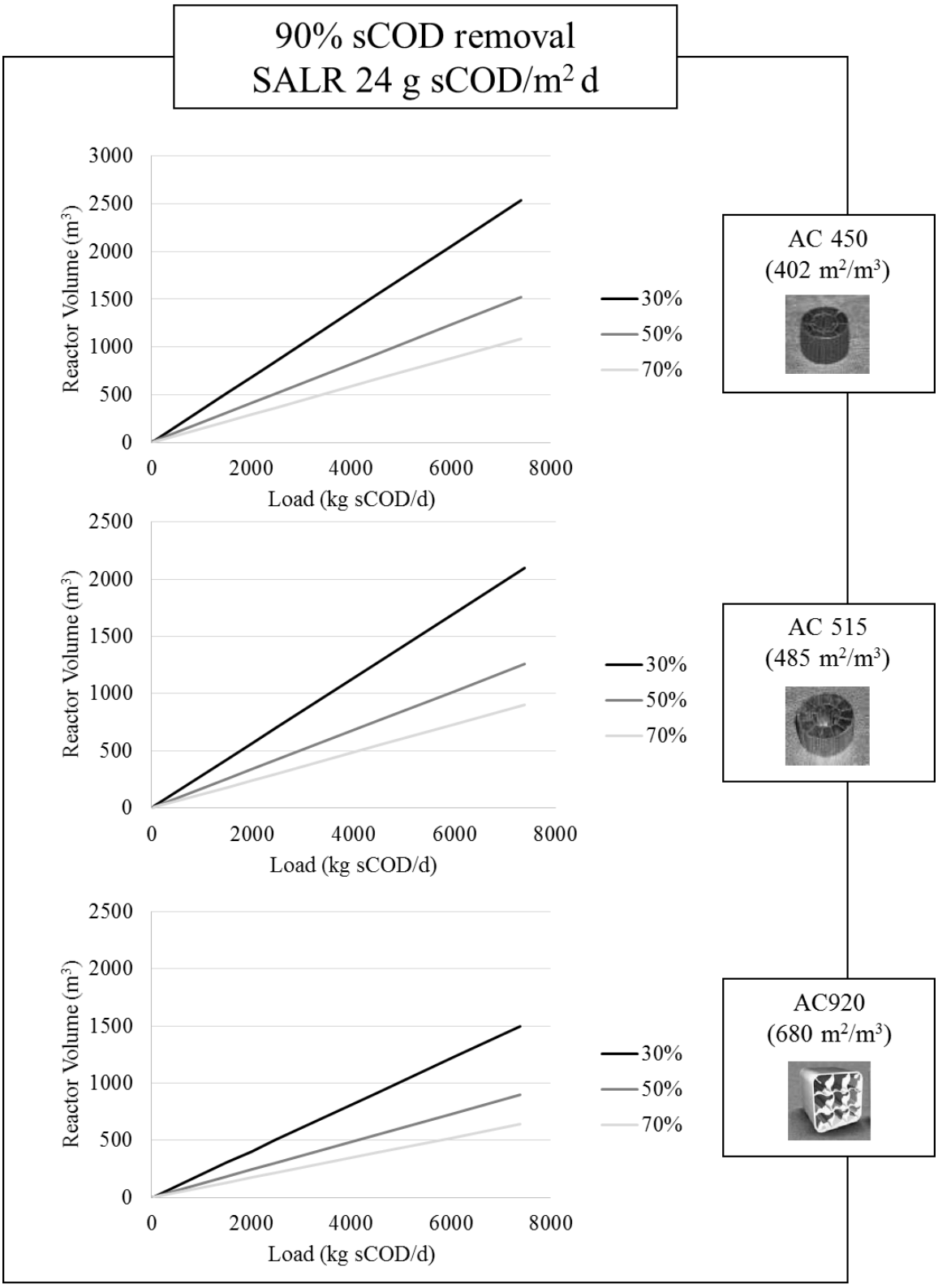


Figure 4-3 – Comparison of reactor volume requirement to achieve 90% removal by implementation of three media (i.e., AC450, AC515 and AC920)

The figures above are clearly showing how the media filling and the specific surface area related to the type of media used have a consistent impact in the total reactor volume required to achieve removal greater than 80 % and 90 %, respectively. The total volume required was calculated according to the following equation – Equation 7.

$$T_V = \frac{V_M}{\%_M} \qquad \text{(Equation 7)}$$

Where T_V is the total volume required, V_M is the volume of media and $\%_M$ is the percentage media filling by volume.

4.2 Case Study – Fort Garry Brewery

A case study of the Fort Garry Brewery (Winnipeg, Canada) was conducted to suggest a potential reactor volume required to treat the wastewater load produced. The brewery characterization information gathered through the thesis study shown that the brewery wastewater has an average sCOD concentration equal to 6 kg/m^3 with a daily flow of 150 m^3/d. Therefore, the loading rate calculated and used in the design is 900 kg sCOD/d.

Base on the experimental data it is not recommended to exceed SALR above 50 g sCOD/m^2d which will led to mass transfer limitation. Therefore using the linear correlation equation in Figure 4-1, allow to calculate SALR and relative removal suitable in designing an AMBBR for Fort Garry Brewery. Hence, an SALR < 50 g sCOD/m^2d was calculated to be at removal equal or greater than 85% (Table 4-3). Table 4-6 shows the design parameters, which allowed optimizing the AMBBR process for Fort Garry Brewery wastewater treatment. The city's industrial discharge into the sewer system law (City of Winnipeg sewer by-law No.92/2010) legislate a limit for BOD5 at 0.3 g BOD_5/L. Hence, for instance implementing one reactor with a volume of 41 m^3 at removal of 85%, will not comply while effluent concentration will be equal

to 0.675 g BOD_5/L (Table 4-6). Even 90% removal with SALR of 23 g sCOD/m^2d will not met the by-law, producing 0.675 g BOD_5/L effluent concentration.

Table 4-6 – Design parameters for AC920 media at 70% fillings

sCOD removal (%)	Total surface area required (m^2)	Volume media required (m^3)	Reactor volume (m^3)	Effluent concentration ($gBOD_5/L$)
85	19500	29	41	0.675
90	39100	58	82	0.45

However, by implementing a 41 m^3 tank with an in series 6 m^3 second tank both designed for 85% removal, the by-law limit will be met having effluent concentration of 0.1 g BOD_5/L. The effluents from the optimization study revealed that the BOD/COD ratio was not significantly changed from 0.77 of the raw wastewater. The effluent BOD, in fact, was greater than 70% the COD. For these reasons, it is assumed that the second reactor will achieve 85% COD removal with similar rate as per raw wastewater treatment.

SUMMARY AND CONCLUSIONS

Performance of AMBBR technology was investigated in its start-up on synthetic wastewater and brewery wastewater treatment application. In the start-up study two AMBBR reactors (4 L working volume) with different quantities of cube-shaped media (25 % and 35%) were examined. Reactors were fed synthetic protein-based feed and both were capable of treating up to 20 kg COD/m^3 d, achieving 80 % COD removal. Beyond that load, the performance began to decrease. The maximum biogas production observed was 0.37 m^3/d at 35°C and standard pressure and an OLR of 20 kg COD/m^3 d. The methane composition varied between 60 % to 70%. Methane yields between 0.28 and 0.45 m^3 CH_4/kg sCOD removed and specific removal rates from 0.4 to 20 kg COD/ m^3 d at standard temperature and pressure were achieved. Kinetic tests performed at the maximum OLR revealed that free-floating suspended biomass contributed to no more than 2.5 % of the total removal. The reactors were also capable of treating brewery wastewater, reliably attaining 80 % COD removal with methane yields between 0.30 to 0.42 m^3 CH_4/kg COD removed at an OLR of 9 kg/m^3 d.

Two months were required for a mature biofilm to develop and start performing to desired levels of above 80 % removal of COD. Results indicated that both synthetic and real wastewater had a strong correlation between total biogas, where 60 – 70 % was methane, and the OLRs applied. The difference in media content of 10 % by volume did not show a significant difference in terms of overall performances.

In the second part of the thesis work, the analysis of AMBBR treating brewery wastewater when subjected to increasing organic load via decreased HRT was investigated. Three reactors with 40 % media filling by volume were used to test HRT of 24, 18, 12, 10, 8, 6 h treating brewery wastewater with average concentrations of 4.0 ± 1 kg sCOD/m^3 and 5.2 ± 2.1 kg COD/m^3.

Results revealed that with an influent BOD/COD ratio equal to 0.77 ± 0.02 the HRT of 18 h has the highest performance as COD removal. At 18 h HRT, 88 ± 2.5 % of the COD and 92.6 ± 1.5 % as sCOD were removed with OLRs 5.5 ± 1.9 kg COD/m^3 d and 3.9 ± 0.3 kg sCOD/ m^3 d. Moreover at this HRT, the highest soluble carbon conversion into methane of 0.36 ± 0.06 m^3 CH_4/kg sCOD was measured. The biogas produced at stable HRTs (24-10 h) was characterized as 67 ± 3 % and 23.0 ± 0.5 % as methane and carbon dioxide, respectively.

Anaerobic moving bed biofilm reactors were demonstrated to be a viable option for the treatment of brewery wastewater under mesophilic conditions. By decreasing the HRT, and therefore increasing the OLR, performance measures were determined for a wastewater of fixed characteristics. The measure of most interest was the SALR, determined to be 87 g sCOD/m^2 d for 70 % removal, 56 g sCOD/m^2 d for 80 % removal, and 24 g sCOD/m^2 d for 90 % removal. Higher SALR than 50 g sCOD/m^2 d resulted in mass transfer limitations due to the biofilm outgrowing the available protected surface area. Furthermore, significant biogas production was attained, with unit methane generations observed to be as high as 0.36 ± 0.06 m^3 CH_4/kg sCOD under mesophilic conditions. Thus, an AMBBR could provide high-rate anaerobic treatment with the potential for energy recovery from waste.

This thesis study proves that brewery wastewater can be reliable treated with AMBBR technology. The design consideration for full-scale application at the Fort Garry brewery shows the capability of treating high strength wastewater in a small footprint process. Therefore, it is feasible to comply with the discharge into sewer limit regarding pH, BOD and TSS with a two tank in series configuration. This setup will therefore optimize at its maximum the process and will required reactor volume of 41 m^3 for the primary and 6 m^3 for the secondary tanks to comply with the by-law resulting in effluent concentration of 0.1 g BOD_5/L.

REFERENCE

Acharya, B. K., Mohana, S., & Madamwar, D. (2008). Anaerobic treatment of distillery spent wash - A study on upflow anaerobic fixed film bioreactor. *Bioresource Technology*, *99*(11), 4621–4626. doi:10.1016/j.biortech.2007.06.060

Acharya, B. K., Pathak, H., Mohana, S., Shouche, Y., Singh, V., & Madamwar, D. (2011). Kinetic modelling and microbial community assessment of anaerobic biphasic fixed film bioreactor treating distillery spent wash. *Water Research*, *45*(14), 4248–4259. doi:10.1016/j.watres.2011.05.048

Alvarado-Lassman, a, Rustrián, E., García-Alvarado, M. a, Rodríguez-Jiménez, G. C., & Houbron, E. (2008). Brewery wastewater treatment using anaerobic inverse fluidized bed reactors. *Bioresource Technology*, *99*(8), 3009–15. doi:10.1016/j.biortech.2007.06.022

APHA. (2012). *Standard Methods - Examination of water and wastewater*. (E. E.W. Rice, R.B. Baird, A.D. Eaton, L.S. Clesceri, Ed.). American Public Health Association, American Water Works Association, Water Environment Federation.

Bacenetti, J., Negri, M., Fiala, M., & González-García, S. (2013). Anaerobic digestion of different feedstocks: Impact on energetic and environmental balances of biogas process. *Science of the Total Environment*, *463-464*, 541–551. doi:10.1016/j.scitotenv.2013.06.058

Barber, W. P., & Stuckey, D. C. (1999). The use of the anaerobic baffled reactor (ABR) for wastewater treatment: A review. *Water Research*, *33*(7), 1559–1578. doi:10.1016/S0043-1354(98)00371-6

Barwal, A., & Chaudhary, R. (2014). To study the performance of biocarriers in moving bed biofilm reactor (MBBR) technology and kinetics of biofilm for retrofitting the existing aerobic treatment systems: a review. *Reviews in Environmental Science and*

Bio/Technology, *13*(3), 285–299. doi:10.1007/s11157-014-9333-7

BeerCanada. (2016). Beer Canada. Retrieved May 17, 2016, from http://www.beercanada.com/statistics

Bertin, L., Lampis, S., Todaro, D., Scoma, A., Vallini, G., Marchetti, L., … Fava, F. (2010). Anaerobic acidogenic digestion of olive mill wastewaters in biofilm reactors packed with ceramic filters or granular activated carbon. *Water Research*, *44*(15), 4537–4549. doi:10.1016/j.watres.2010.06.025

Biswas, K., Taylor, M. W., & Turner, S. J. (2014). Successional development of biofilms in moving bed biofilm reactor (MBBR) systems treating municipal wastewater. *Applied Microbiology and Biotechnology*, *98*(3), 1429–40. doi:10.1007/s00253-013-5082-8

Brewers Association of Canada. (2011). *Guide to Energy Efficiency Opportunities in the Canadian Brewing Industry*.

Brito, A. G., Peixoto, J., Oliveira, J. M., Oliveira, J. a, Costa, C., Nogueira, R., & Rodrigues, A. (2004). Brewery and winery wastewater treatment: some focal points of design and operation. *Water Science & Technology*, 1–22.

Britz, T., & Robinson, R. (2008). *Advanced dairy science and technology*. (S. A. University of Stellenbosch & U. Consultant in Food Science and Technology, Eds.). Blackwell Publishing Ltd, Oxford, UK. doi:10.1002/9780470697634.fmatter

Büyükkamaci, N., & Filibeli, A. (2002). Concentrated wastewater treatment studies using an anaerobic hybrid reactor. *Process Biochemistry*, *38*(5), 771–775. doi:10.1016/S0032-9592(02)00221-2

Cai, M., Wilkins, D., Chen, J., Ng, S.-K., Lu, H., Jia, Y., & Lee, P. K. H. (2016). Metagenomic Reconstruction of Key Anaerobic Digestion Pathways in Municipal Sludge and Industrial

Wastewater Biogas-Producing Systems. *Frontiers in Microbiology*, *7*(May), 1–12. doi:10.3389/fmicb.2016.00778

Chastain, C., Crandall, S., Del Grande, D., Flores, T., Gilliland, M., Horwitz, L., … Utz, M. (2011). *Water and Wastewater : Treatment / Volume Reduction Manual*. Brewers Association Report.

Cronin, C., & Lo, K. V. (1998). Anaerobic treatment of brewery wastewater using UASB reactors seeded with activated sludge. *Bioresource Technology*, *64*(1), 33–38. doi:10.1016/S0960-8524(97)00154-5

Demirel, B., Yenigun, O., & Onay, T. T. (2005). Anaerobic treatment of dairy wastewaters: A review. *Process Biochemistry*, *40*(8), 2583–2595. doi:10.1016/j.procbio.2004.12.015

Environsim. (2015). Biowin 5.0.

Ertem, F. C. (2011). *Improving biogas production by anaerobic digestion of different substrates - Calulation of potential energy outcomes*. Halmstad University.

Faisal, M., & Unno, H. (2001). Kinetic analysis of palm oil mill wastewater treatment by a modified anaerobic baffled reactor. *Biochemical Engineering Journal*, *9*(1), 25–31. doi:10.1016/S1369-703X(01)00122-X

Fang, H. H. P., & Chan, O. C. (1997). Toxicity of phenol towards anaerobic biogranules. *Water Research*, *31*(9), 2229–2242. doi:10.1016/S0043-1354(97)00069-9

Fernández, N., Díaz, E. E., Amils, R., & Sanz, J. L. (2008). Analysis of microbial community during biofilm development in an anaerobic wastewater treatment reactor. *Microbial Ecology*, *56*(1), 121–32. doi:10.1007/s00248-007-9330-2

Fernandez, N., Montalvo, S., Borja, R., Guerrero, L., S??nchez, E., Cort??s, I., … Raposo, F. (2008). Performance evaluation of an anaerobic fluidized bed reactor with natural zeolite as

support material when treating high-strength distillery wastewater. *Renewable Energy*, *33*(11), 2458–2466. doi:10.1016/j.renene.2008.02.002

Ferreira, M., Marques, I. P., & Malico, I. (2012). Biogas in Portugal: Status and public policies in a European context. *Energy Policy*, *43*, 267–274. doi:10.1016/j.enpol.2012.01.003

Gangagni Rao, A., Venkata Naidu, G., Krishna Prasad, K., Chandrasekhar Rao, N., Venkata Mohan, S., Jetty, A., & Sarma, P. N. (2005). Anaerobic treatment of wastewater with high suspended solids from a bulk drug industry using fixed film reactor (AFFR). *Bioresource Technology*, *96*(1), 87–93. doi:10.1016/j.biortech.2003.05.007

Garcia-Calderon, D., Buffiere, P., Moletta, R., & Elmaleh, S. (1998). Anaerobic digestion of wine distillery wastewater in down-flow fluidized bed. *Water Research*, *32*(12), 3593–3600. doi:10.1016/S0043-1354(98)00134-1

Gerardi, M. H. (2003). *The Microbiology of Anaerobic Digesters*. John Wiley & Sons, Inc., Hoboken, NJ, USA. doi:10.1002/0471468967.ch7

Grobicki, A. M. W., & Stuckey, D. C. (1991). Performance of the anaerobic baffled reactor under steady state and shock loading conditions. *Biotechnology Bioeng*, *37*, 344–355.

Grover, R., Marwaha, S. S., & Kennedy, J. F. (1999). Studies on the use of an anaerobic baffled reactor for the continuous anaerobic digestion of pulp and paper mill black liquors. *Process Biochemistry*, *34*(6-7), 653–657. doi:10.1016/S0032-9592(98)00138-1

Guo, J., Peng, Y., Ni, B.-J., Han, X., Fan, L., & Yuan, Z. (2015). Dissecting microbial community structure and methane-producing pathways of a full-scale anaerobic reactor digesting activated sludge from wastewater treatment by metagenomic sequencing. *Microbial Cell Factories*, *14*(1), 33. doi:10.1186/s12934-015-0218-4

Hassan, S., & Dahlan, I. (2013). Anaerobic wastewater treatment using anaerobic baffled

bioreactor: a review. *Central European Journal of Engineering*, *3*(3), 389–399. doi:10.2478/s13531-013-0107-8

Hidaka, T., Wang, F., & Tsumori, J. (2015). Comparative evaluation of anaerobic digestion for sewage sludge and various organic wastes with simple modeling. *Waste Management*, *43*, 144–151. doi:10.1016/j.wasman.2015.04.026

Jafarzadeh, M. T., Mehrdadi, N., & Hashemian, S. J. (2012). Application of an anaerobic hybrid reactor for petrochemical effluent treatment. *Water Science and Technology*, *65*(12), 2098–2105. doi:10.2166/wst.2012.088

Jahren, S. J., Rintala, J. a, & Odegaard, H. (2002). Aerobic moving bed biofilm reactor treating thermomechanical pulping whitewater under thermophilic conditions. *Water Research*, *36*(4), 1067–75. Retrieved from http://www.ncbi.nlm.nih.gov/pubmed/11848344

Josse, J. C., Theodoulou, M. D., & Scattergood, S. (2013). Wastewater treatment process with moving bed bioreactor (MBBR).

Kalyuzhnyi, S. V., Perez Martinez, E., & Rodriguez Martinez, J. (1997). Anaerobic treatment of high-strength cheese-whey wastewaters in laboratory and pilot UASB-reactors. *Bioresource Technology*, *60*(1), 59–65. doi:10.1016/S0960-8524(96)00176-9

Kennedy, K. J. (1985). *Startup and steady state kinetics of anaerobic downflow stationary fixed film reactors*.

Kermani, M., Bina, B., Movahedian, H., Amin, M. M., & Nikaein, M. (2008). Application of Moving Bed Biofilm Process for Biological Organics and Nutrients Removal from Municipal Wastewater. *American Journal of Environmental Sciences*, *4*(6), 675–682.

Kleerebezem, R., Joosse, B., Rozendal, R., & Van Loosdrecht, M. C. M. (2015). Anaerobic digestion without biogas? *Reviews in Environmental Science and Bio/Technology*.

doi:10.1007/s11157-015-9374-6

Kumar, A., Yadav, A. K., Sreekrishnan, T. R., Satya, S., & Kaushik, C. P. (2008). Treatment of low strength industrial cluster wastewater by anaerobic hybrid reactor. *Bioresource Technology*, *99*(8), 3123–3129. doi:10.1016/j.biortech.2007.05.056

Kundu, K., Sharma, S., & Sreekrishnan, T. R. (2013). Changes in microbial communities in a hybrid anaerobic reactor with organic loading rate and temperature. *Bioresource Technology*, *129*, 538–547. doi:10.1016/j.biortech.2012.11.118

Latif, M. A., Ghufran, R., Wahid, Z. A., & Ahmad, A. (2011). Integrated application of upflow anaerobic sludge blanket reactor for the treatment of wastewaters. *Water Research*, *45*(16), 4683–4699. doi:10.1016/j.watres.2011.05.049

Li, Y., Park, S. Y., & Zhu, J. (2011). Solid-state anaerobic digestion for methane production from organic waste. *Renewable and Sustainable Energy Reviews*, *15*(1), 821–826. doi:10.1016/j.rser.2010.07.042

Lim, S. J., & Kim, T. H. (2014). Applicability and trends of anaerobic granular sludge treatment processes. *Biomass and Bioenergy*, *60*, 189–202. doi:10.1016/j.biombioe.2013.11.011

Lu, X., Zhen, G., Chen, M., Kubota, K., & Li, Y. Y. (2015). Biocatalysis conversion of methanol to methane in an upflow anaerobic sludge blanket (UASB) reactor: Long-term performance and inherent deficiencies. *Bioresource Technology*, *198*(October), 691–700. doi:10.1016/j.biortech.2015.09.073

Lusk, P. (1999). Latest progress in anaerobic digestion. *BioCycle*, *40*(7).

Madigan, M. T., Martinko, H. M., Stahl, D., & Clark, D. P. (2012). *Brock Biology and Microorganisms*.

Mahony, T., O'Flaherty, V., Colleran, E., Killilea, E., Scott, S., & Curtis, J. (2002). *Feasibility*

Study for Centralised Anaerobic Digestion for Treatment of Various Wastes and Wastewaters in Sensitive Catchment Areas: Final Project Report.

Mata-Alvarez, J. (2000). Anaerobic digestion of organic solid wastes . An overview of research achievements and perspectives. *Bioresource Technology*, *74*, 3–16.

Maunoir, S., Sabil, N., Rambaud, a., Philip, H., & Coletti-Previero, M. a. (1991). Role of insoluble enzymes in anaerobic wastewater treatment and enzyme-bioactivator interactions. *Environmental Technology*, *12*(4), 313–323. doi:10.1080/09593339109385011

McQuarrie, J. P., & Boltz, J. P. (2011). Moving Bed Biofilm Reactor Technology: Process Applications, Design, and Performance. *Water Environment Research*, *83*(6), 560–575. doi:10.2175/106143010X12851009156286

Metcalf, & Eddy. (2013). *Wastewater Engineering: Treatment and Resource Recovery* (5 edition.). McGraw-Hill Education, 2 Penn Plaza, New York, NY 10121.

Meyer, T., & Edwards, E. A. (2014). Anaerobic digestion of pulp and paper mill wastewater and sludge. *Water Research*, *65*, 321–349. doi:10.1016/j.watres.2014.07.022

Mijaylova Nacheva, P., Moeller Chávez, G., Matías Chacón, J., & Canul Chuil, A. (2009). Treatment of cane sugar mill wastewater in an upflow anaerobic sludge bed reactor. *Water Science and Technology*, *60*(5), 1347–1352. doi:10.2166/wst.2009.402

Monteiro, E., Mantha, V., & Rouboa, A. (2011). Prospective application of farm cattle manure for bioenergy production in Portugal. *Renewable Energy*, *36*(2), 627–631. doi:10.1016/j.renene.2010.08.035

Moody, L., Burns, R., Wu-Haan, W., & Spajić, R. (2009). Use of biochemical methane potential (BMP) assays for predicting and enhancing anaerobic digester performance. *Proceedings of The 4th International and 44th Croatian Symposium of Agriculture*, 930–934. Retrieved

from http://sa.agr.hr/pdf/2009/sa2009_p1009.pdf

Morgavi, D. P., Martin, C., Jouany, J.-P., & Ranilla, M. J. (2012). Rumen protozoa and methanogenesis: not a simple cause–effect relationship. *British Journal of Nutrition*, *107*(03), 388–397. doi:10.1017/S0007114511002935

Mosquera-Corral, A., Sanchez, M., Campos, J. L., Mendez, R., & Lema, J. M. (2001). Simultaneous methanogenesis and denitrification of pretreated effluents from a fish canning industry. *Water Research*, *35*(2), 411–418. doi:10.1016/S0043-1354(00)00288-8

Najafpour, G. D., Zinatizadeh, A. A. L., Mohamed, A. R., Hasnain Isa, M., & Nasrollahzadeh, H. (2006). High-rate anaerobic digestion of palm oil mill effluent in an upflow anaerobic sludge-fixed film bioreactor. *Process Biochemistry*, *41*(2), 370–379. doi:10.1016/j.procbio.2005.06.031

Ødegaard. (1994). A new moving bed biofilm reactor - application and results.pdf. *Water Science & Technology*, *29*, 157–165.

Ødegaard, H. (1999). The Moving Bed Biofilm Reactor. *Water Environmental Engineering and Reuse of Water*, (0575314), 205–305.

Olajire, A. a. (2012). The brewing industry and environmental challenges. *Journal of Cleaner Production*, 1–21. doi:10.1016/j.jclepro.2012.03.003

Oleszkiewicz, J. A., & Mavinic, D. S. (2001). Wastewater biosolids: an overview of processing, treatment, and management. *Canadian Journal of Civil Engineering*, *28*(S1), 102–114. doi:10.1139/cjce-28-S1-102

Oleszkiewicz, J. A., & Romanek, A. (1989). Granulation in anaerobic sludge bed reactors treating food industry wastes. *Biological Wastes*, *27*(3), 217–235. doi:10.1016/0269-7483(89)90003-7

Ozuolmez, D., Na, H., Lever, M. A., Kjeldsen, K. U., Jørgensen, B. B., & Plugge, C. M. (2015). Methanogenic archaea and sulfate reducing bacteria co-cultured on acetate: Teamwork or coexistence? *Frontiers in Microbiology*, *6*(MAY), 1–12. doi:10.3389/fmicb.2015.00492

Patel, H., & Madamwar, D. (2001). Single and multichamber fixed film anaerobic reactors for biomethanation of acidic petrochemical wastewater-systems performance. *Process Biochemistry*, *36*(7), 613–619. doi:10.1016/S0032-9592(00)00249-1

Petrillo, N. (2015). *Going flat : Changing drinking patterns and low international demand will limit revenue growth Breweries in Canada About this Industry*.

Pol, L. H., & Lettinga, G. (1986). New technologies for anaerobic wastewater treatment. *Water Science & Technology*, *18*(12), 41–53.

Pradeep, N. V., Anupama, S., Arun Kumar, J. M., Vidyashree, K. G., Lakshmi, P., Ankitha, K., & Pooja, J. (2014). Treatment of Sugar Industry Wastewater in Anaerobic Downflow Stationary Fixed Film (DSFF) Reactor. *Sugar Tech*, *16*(1), 9–14. doi:10.1007/s12355-013-0227-8

Rajeshwari, K. ., Balakrishnan, M., Kansal, a, Lata, K., & Kishore, V. V. . (2000). State-of-the-art of anaerobic digestion technology for industrial wastewater treatment. *Renewable and Sustainable Energy Reviews*, *4*(2), 135–156. doi:10.1016/S1364-0321(99)00014-3

Ratanatamskul, C., & Siritiewsri, T. (2014). A compact on-site UASB-EGSB system for organic and suspended solid digestion and biogas recovery from department store wastewater. *International Biodeterioration and Biodegradation*, *102*, 24–30. doi:10.1016/j.ibiod.2015.04.002

Rodrigues, L. S., Silva, I. J., Oliveira, P. R., Pinto, A. C. A., & Lima, C. A. (2014). Influence of the Ratio IA/PA and Volatile Acids in the Monitoring of UASB Reactor in the Treating of

Swine Waste Water. *Journal of Water Resource and Protection*, *6*(April), 526–531.

Rosenzweig, A., & Ragsdale, S. W. (2011). *Methods in Enzymology, Volume 494, Methods in Methane Metabolism, Part A* (First.). Elsevier.

Rusten, B., Eikebrokk, B., Ulgenes, Y., & Lygren, E. (2006). Design and operations of the Kaldnes moving bed biofilm reactors. *Aquacultural Engineering*, *34*(3), 322–331. doi:10.1016/j.aquaeng.2005.04.002

Seghezzo, L., Zeeman, G., Van Lier, J. B., Hamelers, H. V. M., & Lettinga, G. (1998). A review: The anaerobic treatment of sewage in UASB and EGSB reactors. *Bioresource Technology*, *65*(3), 175–190. doi:10.1016/S0960-8524(98)00046-7

Sevilla-Espinosa, S., Solórzano-Campo, M., & Bello-Mendoza, R. (2010). Performance of staged and non-staged up-flow anaerobic sludge bed (USSB and UASB) reactors treating low strength complex wastewater. *Biodegradation*, *21*(5), 737–751. doi:10.1007/s10532-010-9339-y

Sheli, C., Guo, J., Yuan, C., Jing, C., & Lina, G. (2014). Anaerobic treatment of winery wastewater in moving bed biofilm reactors. *Desalination and Water Treatment*, *52*, 1841–1849. doi:10.1007/s00449-009-0387-9

Sheli, C., & Moletta, R. (2007). Anaerobic treatment of vinasses by a sequentially mixed moving bed biofilm reactor. *Water Science & Technology*, *56*(2), 1. doi:10.2166/wst.2007.465

Sheli, C., & Moletta, R. (2010). Treatment of wine distillery wastewater using an anaerobic moving bed biofilm reactor with low density of polyethylene support. *International Conference on Challenges in Environmental Science and Computer Engineering, CESCE 2010*, *2*, 232–235. doi:10.1109/CESCE.2010.229

Shen, Y., Linville, J. L., Urgun-Demirtas, M., Mintz, M. M., & Snyder, S. W. (2015). An

overview of biogas production and utilization at full-scale wastewater treatment plants (WWTPs) in the United States: Challenges and opportunities towards energy-neutral WWTPs. *Renewable and Sustainable Energy Reviews*, *50*, 346–362. doi:10.1016/j.rser.2015.04.129

Somasiri, W., Li, X. F., Ruan, W. Q., & Jian, C. (2008). Evaluation of the efficacy of upflow anaerobic sludge blanket reactor in removal of colour and reduction of COD in real textile wastewater. *Bioresource Technology*, *99*(9), 3692–3699. doi:10.1016/j.biortech.2007.07.024

Sunil Kumar, G., Gupta, S. K., & Singh, G. (2007). Biodegradation of distillery spent wash in anaerobic hybrid reactor. *Water Research*, *41*(4), 721–730. doi:10.1016/j.watres.2006.11.039

Taconi, K. a., Zappi, M. E., Todd French, W., & Brown, L. R. (2008). Methanogenesis under acidic pH conditions in a semi-continuous reactor system. *Bioresource Technology*, *99*(17), 8075–8081. doi:10.1016/j.biortech.2008.03.068

Tauseef, S. M., Abbasi, T., & Abbasi, S. A. (2013). Energy recovery from wastewaters with high-rate anaerobic digesters. *Renewable and Sustainable Energy Reviews*, *19*, 704–741. doi:10.1016/j.rser.2012.11.056

Tawfik, a, & El-Kamah, H. (2012). Treatment of fruit-juice industry wastewater in a two-stage anaerobic hybrid (AH) reactor system followed by a sequencing batch reactor (SBR). *Environmental Technology*, *33*(4-6), 429–36. doi:10.1080/09593330.2011.579178

Tilche, A., & Vieira, M. M. (1991). Discussion report on reactor design of anaerobic filters and sludge bed reactors. *Water Science & Technology*, *24*(8), 193–206.

van den Berg, L., Kennedy, K. ., & Samson, R. (1985). Anaerobic Downflow Stationary Fixed

Film Reactors: Performance Under Steady-State and Non-Steady State Conditions. *Water Science & Technology*, *17*(1), 89–102.

Van Lier, J. B., Boersma, F., Debets, M. M. W. H., & Lettinga, G. (1994). High rate thermophilic anaerobic wastewater treatment in compartmentalized upflow reactors. *Water Science & Technology*, *30*, 251–261.

Viana, M. B., Freitas, A. V., Leitão, R. C., Pinto, G. A. S., & Santaella, S. T. (2012). Anaerobic digestion of crude glycerol: a review. *Environmental Technology Reviews*, *1*(1), 81–92. doi:10.1080/09593330.2012.692723

Vidal, G., Carvalho, A., Méndez, R., & Lema, J. M. (2000). Influence of the content in fats and proteins on the anaerobic biodegradability of dairy wastewaters. *Bioresource Technology*, *74*(3), 231–239. doi:10.1016/S0960-8524(00)00015-8

Vlyssides, A., Mai, S., & Barampouti, E. M. (2015). Energy Generation Potential in Greece From Agricultural Residues and Livestock Manure by Anaerobic Digestion Technology. *Waste and Biomass Valorization*, *6*(5), 747–757. doi:10.1007/s12649-015-9400-5

Wahab, M. A., Habouzit, F., Bernet, N., Steyer, J. P., Jedidi, N., & Escudi??, R. (2014). Sequential operation of a hybrid anaerobic reactor using a lignocellulosic biomass as biofilm support. *Bioresource Technology*, *172*, 150–155. doi:10.1016/j.biortech.2014.08.127

Wang, K., Yin, J., Shen, D., & Li, N. (2014). Anaerobic digestion of food waste for volatile fatty acids (VFAs) production with different types of inoculum: Effect of pH. *Bioresource Technology*, *161*, 395–401. doi:10.1016/j.biortech.2014.03.088

Wang, S., Rao, N. C., Qiu, R., & Moletta, R. (2009). Performance and kinetic evaluation of anaerobic moving bed biofilm reactor for treating milk permeate from dairy industry.

Bioresource Technology, *100*(23), 5641–7. doi:10.1016/j.biortech.2009.06.028

Xu, F., Wang, F., Lin, L., & Li, Y. (2016). Comparison of digestate from solid anaerobic digesters and dewatered effluent from liquid anaerobic digesters as inocula for solid state anaerobic digestion of yard trimmings. *Bioresource Technology*, *200*, 753–760. doi:10.1016/j.biortech.2015.10.103

Yang, Y., Yu, K., Xia, Y., Lau, F. T. K., Tang, D. T. W., Fung, W. C., … Zhang, T. (2014). Metagenomic analysis of sludge from full-scale anaerobic digesters operated in municipal wastewater treatment plants. *Applied Microbiology and Biotechnology*, *98*(12), 5709–5718. doi:10.1007/s00253-014-5648-0

Zhang, Z. Y., & Maekawa, T. (1996). Effects of sulfur-containing compounds on the growth and methane production of acclimated-mixed methanogens. *Biomass and Bioenergy*, *10*(1), 45–56. doi:10.1016/0961-9534(95)00054-2

APPENDIX

6.1 Start-up and transition to brewery wastewater

6.1.1 pH, alkalinity, influent and effluent COD, and OLR

35% Reactor							
	pH	Alkalinity	CODin	CODout	OLR	COD rem efficiency	COD rem efficiency
Day	-	g/L	g/L	g/L	kgCOD/m3d	g/L	%
0	7.1	1.8	2.02	1.52	1.91	0.50	25.0
3	7.3	1.9	2.02	1.43	1.91	0.59	29.0
5	7.2	2.0	1.29	0.72	0.37	0.57	44.5
7	7.3	1.8	0.87	0.42	0.25	0.46	52.3
10	7.5	1.8	0.49	0.35	0.46	0.13	27.6
12	7.5	1.8	0.49	0.28	0.46	0.20	41.6
14	7.5	1.8	0.50	0.30	0.47	0.20	40.3
17	7.6	1.8	0.50	0.22	0.47	0.28	55.8
19	7.6	1.8	0.51	0.21	0.48	0.30	58.0
24	7.8	1.7	0.52	0.18	0.49	0.35	66.5
26	7.9	1.8	0.49	0.14	0.46	0.36	72.3
28	7.8	1.6	0.50	0.14	0.48	0.36	71.4
31	7.9	1.7	0.49	0.14	0.46	0.36	72.5
33	7.8	1.7	0.52	0.12	0.49	0.40	77.7
35	7.9	1.6	0.44	0.09	0.42	0.35	79.2
38	7.6	1.6	0.49	0.13	0.46	0.36	74.4
40	7.6	1.7	0.54	0.11	0.51	0.43	80.2
42	7.7	1.7	0.49	0.11	0.46	0.38	77.5
45	7.7	1.6	0.49	0.11	0.46	0.38	77.6
47	7.6	1.7	0.51	0.11	0.48	0.40	78.7
49	7.9	1.7	0.48	0.10	0.46	0.38	79.5
52	7.9	1.6	0.48	0.11	0.45	0.37	77.4
54	7.7	1.7	0.76	0.15	0.72	0.61	79.8
56	7.8	1.9	0.75	0.15	0.71	0.60	79.8
59	7.7	1.7	0.76	0.16	0.72	0.61	79.6
61	7.7	1.8	0.80	0.19	0.75	0.61	76.8
63	7.7	1.7	0.73	0.21	0.69	0.53	72.0
66	7.8	1.8	0.75	0.15	0.71	0.60	80.4
68	7.7	1.8	0.75	0.10	0.71	0.65	86.3
70	7.7	1.7	0.75	0.09	0.71	0.66	87.6
73	7.6	1.8	0.93	0.13	0.88	0.80	86.0
75	7.7	1.9	0.97	0.15	0.92	0.82	84.1
77	7.7	1.9	0.98	0.11	0.93	0.87	88.4

35% Reactor							
	pH	Alkalinity	CODin	CODout	OLR	COD rem efficiency	COD rem efficiency
Day	-	g/L	g/L	g/L	kgCOD/m3d	g/L	%
80	7.7	1.9	0.99	0.13	0.93	0.86	87.3
82	7.7	2.0	1.69	0.28	1.60	1.41	83.5
84	7.7	2.0	1.74	0.22	1.65	1.52	87.6
87	7.7	2.0	1.63	0.21	1.54	1.42	86.9
89	7.7	2.0	1.67	0.18	1.58	1.49	89.0
91	7.7	2.1	1.66	0.20	1.57	1.46	88.0
94	7.6	2.0	1.66	0.20	1.57	1.46	87.8
96	7.6	1.5	2.17	0.19	1.80	1.98	91.4
98	7.6	1.9	2.23	0.25	2.16	1.98	88.9
101	7.5	1.8	2.37	0.27	2.30	2.10	88.6
103	7.4	1.8	3.07	0.37	2.98	2.70	87.8
105	7.4	1.9	2.90	0.37	2.82	2.54	87.4
108	7.4	1.9	3.04	0.40	2.96	2.65	87.0
110	7.5	2.3	4.78	0.55	4.64	4.23	88.6
112	7.5	2.5	4.50	0.56	4.37	3.94	87.5
115	7.5	2.5	4.45	0.38	4.33	4.07	91.4
117	7.7	2.6	5.89	0.40	5.73	5.49	93.2
119	7.5	2.7	5.69	0.48	5.53	5.21	91.5
122	7.5	2.7	5.89	0.41	5.73	5.48	93.0
124	7.5	3.1	9.46	0.77	9.20	8.69	91.9
126	7.5	2.6	8.48	0.67	8.24	7.81	92.1
129	7.3	2.4	6.47	0.57	6.29	5.90	91.2
131	7.5	3.7	15.72	1.63	15.28	14.09	89.6
133	7.4	3.8	14.13	1.15	13.73	12.98	91.9
136	7.4	3.8	17.28	1.69	16.79	15.58	90.2
138	7.4	4.7	22.28	4.11	21.65	18.16	81.5
140	7.5	4.8	19.88	3.51	19.32	16.37	82.3
143	7.5	5.1	23.38	2.91	22.72	20.47	87.6
145	7.5	4.8	22.48	3.51	21.85	18.97	84.4
147	7.5	4.9	21.95	3.08	21.34	18.87	86.0
150	7.7	4.9	23.13	3.70	22.48	19.43	84.0
154	7.4	3.0	11.40	1.77	11.08	9.63	84.5
157	7.4	3.2	13.83	2.54	13.44	11.30	81.7
159	7.3	2.9	11.78	2.77	11.45	9.01	76.5
161	7.4	3.8	24.32	5.09	23.64	19.23	79.1
164	7.5	4.8	22.74	6.69	22.10	16.05	70.6
166	7.0	2.1	9.58	3.25	9.31	6.33	66.1
168	7.1	1.8	8.68	2.46	8.44	6.22	71.7
171	6.9	1.1	4.01	0.76	3.90	3.25	81.1
173	7.1	1.1	2.93	0.72	2.85	2.21	75.5
175	6.9	1.2	2.93	0.60	2.85	2.33	79.7
178	6.9	1.1	4.86	0.84	4.72	4.02	82.7
180	6.9	1.2	5.43	0.90	5.28	4.53	83.4

25% Reactor							
	pH	Alkalinity	CODin	CODout	OLR	COD rem efficiency	COD rem efficiency
Day	-	g/L	g/L	g/L	kgCOD/m3d	g/L	%
0	6.9	1.8	2.02	1.72	1.91	0.30	14.9
3	7.1	1.9	2.02	1.64	1.91	0.38	18.8
5	7.1	2.0	1.43	1.12	0.41	0.31	21.5
7	7.2	1.8	1.05	0.68	0.30	0.38	35.6
10	7.5	1.8	0.49	0.32	0.46	0.16	33.4
12	7.5	1.8	0.49	0.34	0.46	0.14	29.3
14	7.5	1.8	0.50	0.35	0.47	0.15	30.9
17	7.6	1.8	0.50	0.33	0.47	0.17	33.3
19	7.5	1.8	0.51	0.28	0.48	0.24	46.1
24	7.7	1.7	0.52	0.22	0.49	0.30	58.0
26	7.8	1.8	0.49	0.17	0.46	0.33	66.4
28	7.7	1.6	0.50	0.16	0.48	0.34	68.4
31	7.9	1.7	0.49	0.15	0.46	0.34	69.2
33	7.9	1.7	0.52	0.14	0.49	0.38	73.3
35	7.9	1.6	0.44	0.13	0.42	0.32	71.6
38	7.6	1.6	0.49	0.15	0.46	0.34	69.1
40	7.6	1.7	0.54	0.11	0.51	0.43	80.0
42	7.6	1.7	0.49	0.12	0.46	0.36	75.1
45	7.6	1.6	0.49	0.12	0.46	0.37	75.2
47	7.7	1.7	0.51	0.09	0.48	0.42	82.5
49	7.8	1.7	0.48	0.10	0.46	0.38	79.2
52	7.8	1.6	0.48	0.10	0.45	0.38	78.9
54	7.7	1.7	0.76	0.18	0.72	0.58	76.1
56	7.7	1.9	0.75	0.16	0.71	0.60	79.3
59	7.7	1.7	0.76	0.19	0.72	0.57	75.1
61	7.6	1.8	0.80	0.21	0.75	0.59	73.8
63	7.7	1.7	0.73	0.21	0.69	0.53	71.7
66	7.7	1.8	0.75	0.16	0.71	0.59	78.9
68	7.7	1.8	0.75	0.17	0.71	0.58	77.0
70	7.7	1.7	0.75	0.14	0.71	0.61	81.6
73	7.6	1.8	0.93	0.20	0.88	0.72	78.1
75	7.7	1.9	0.97	0.22	0.92	0.76	77.8
77	7.7	1.9	0.98	0.20	0.93	0.78	79.9
80	7.7	1.9	0.99	0.19	0.93	0.79	80.4
82	7.7	2.0	1.69	0.40	1.60	1.29	76.1
84	7.7	2.0	1.74	0.37	1.65	1.37	78.9
87	7.7	2.0	1.63	0.28	1.54	1.35	82.8
89	7.7	2.0	1.67	0.25	1.58	1.43	85.3
91	7.6	2.1	1.66	0.18	1.57	1.48	89.3

25% Reactor							
	pH	Alkalinity	CODin	CODout	OLR	COD rem efficiency	COD rem efficiency
Day	-	g/L	g/L	g/L	kgCOD/m3d	g/L	%
94	7.6	2.0	1.66	0.20	1.57	1.46	87.8
96	7.6	1.5	2.17	0.19	1.80	1.98	91.4
98	7.6	1.9	2.23	0.25	2.16	1.98	88.9
101	7.5	1.8	2.37	0.27	2.30	2.10	88.6
103	7.4	1.8	3.07	0.37	2.98	2.70	87.8
105	7.4	1.9	2.90	0.37	2.82	2.54	87.4
108	7.4	1.9	3.04	0.40	2.96	2.65	87.0
110	7.5	2.3	4.78	0.55	4.64	4.23	88.6
112	7.5	2.5	4.50	0.56	4.37	3.94	87.5
115	7.5	2.5	4.45	0.38	4.33	4.07	91.4
117	7.7	2.6	5.89	0.40	5.73	5.49	93.2
119	7.5	2.7	5.69	0.48	5.53	5.21	91.5
122	7.5	2.7	5.89	0.41	5.73	5.48	93.0
124	7.5	3.1	9.46	0.77	9.20	8.69	91.9
126	7.5	2.6	8.48	0.67	8.24	7.81	92.1
129	7.3	2.4	6.47	0.57	6.29	5.90	91.2
131	7.5	3.7	15.72	1.63	15.28	14.09	89.6
133	7.4	3.8	14.13	1.15	13.73	12.98	91.9
136	7.4	3.8	17.28	1.69	16.79	15.58	90.2
138	7.4	4.7	22.28	4.11	21.65	18.16	81.5
140	7.5	4.8	19.88	3.51	19.32	16.37	82.3
143	7.5	5.1	23.38	2.91	22.72	20.47	87.6
145	7.5	4.8	22.48	3.51	21.85	18.97	84.4
147	7.5	4.9	21.95	3.08	21.34	18.87	86.0
150	7.7	4.9	23.13	3.70	22.48	19.43	84.0
154	7.4	3.0	11.40	1.77	11.08	9.63	84.5
157	7.4	3.2	13.83	2.54	13.44	11.30	81.7
159	7.3	2.9	11.78	2.77	11.45	9.01	76.5
161	7.4	3.8	24.32	5.09	23.64	19.23	79.1
164	7.5	4.8	22.74	6.69	22.10	16.05	70.6
166	7.0	2.1	9.58	3.25	9.31	6.33	66.1
168	7.1	1.8	8.68	2.46	8.44	6.22	71.7
171	6.9	1.1	4.01	0.76	3.90	3.25	81.1
173	7.1	1.1	2.93	0.72	2.85	2.21	75.5
175	6.9	1.2	2.93	0.60	2.85	2.33	79.7
178	6.9	1.1	4.86	0.84	4.72	4.02	82.7
180	6.9	1.2	5.43	0.90	5.28	4.53	83.4

6.1.2 VFA raw data in mg/L

Day	Sample	Formate	Acetate	Propionate	Isobutyrate	Butyrate	Isovalerate	Valerate
122	Feed	303.5	592.9	0	0	0	0	0
	35%	0	28.4	0	0	0	0	0
	25%	0	88.1	9.5	0	0	0	0
131	Feed	783.5	687.9	0	0	0	0	0
	35%	0	156.3	20.8	0	0	0	0
	25%	0	302.1	68.7	0	0	0	0
138	Feed	942.1	1000.1	0	0	0	0	0
	35%	0	274.5	60.1	0	0	0	0
	25%	0	620.2	198.9	0	9.8	7.7	0
145	Feed	821.6	934.8	0	0	0	0	0
	35%	0	252.2	48.0	0	0	0	0
	25%	0	494.1	161.5	0	26.1	8.2	0
Standards	**mM**	46.0	60.1	75.0	88.1	88.1	102.1	102.1

6.1.3 Total gas production and composition

35% Reactor						
	OLR	sCOD rem	Methane	Biogas	Total gas production	Methane production
Day	kg/m3 d	g/d	L/d	L/d	L/d	L/d
110	4.6	16.4	5.5	6.8	6.8	5.5
119	5.5	20.2	6.9	9.0	9.0	6.9
124	9.2	33.8	10.3	14.0	14.0	10.3
132	15.3	54.8	19.2	26.4	26.4	19.2
138	19.3	63.6	23.3	36.6	36.6	23.3
147	21.3	73.4	20.2	32.8	32.8	20.2
161	23.6	74.8	21.9	34.0	34.0	21.9
167	8.4	24.2	10.9	17.5	17.5	10.9
173	2.8	9.4	2.8	4.4	4.4	2.8
178	5.3	17.6	6.1	9.5	9.5	6.1
185	7.5	25.6	10.8	16.4	16.4	10.8

25% Reactor						
	OLR	sCOD rem	Methane	Biogas	Total gas production	Methane production
Day	kg/m3 d	g/d	L/d	L/d	L/d	L/d
110	4.7	15.9	6.1	7.8	7.8	6.1
119	5.6	19.8	7.9	9.8	9.8	7.9
124	9.4	33.5	11.7	15.6	15.6	11.7
132	15.6	55.1	16.9	26.0	26.0	16.9
138	19.7	62.0	25.5	37.2	37.2	25.5
147	21.7	72.4	24.0	32.2	32.2	24.0
161	24.1	56.8	23.2	36.3	36.3	23.2
167	8.6	24.5	9.5	15.6	15.6	9.5
173	2.9	9.3	3.3	5.2	5.2	3.3
178	5.4	17.3	7.2	11.0	11.0	7.2
185	7.7	26.1	10.8	16.4	16.4	10.8

	CH_4		CO_2		N_2	
	%		%		%	
Day	**35%**	**25%**	**35%**	**25%**	**35%**	**25%**
110	80.5	78.8	8.5	7.3	10.9	13.9
119	76.5	81.3	22.1	12.9	1.3	5.9
123	73.8	74.7	25.2	23.7	1.0	1.6
132	72.8	65.2	22.2	34.2	5.0	0.6
139	63.6	68.6	35.0	26.5	0.6	3.7
146	61.6	74.7	34.5	15.2	3.4	9.8
161	64.5	63.9	33.5	34.0	2.1	2.0
167	62.4	61.0	34.5	34.5	3.1	4.6
173	65.0	64.0	26.7	26.4	8.2	9.5
180	64.0	65.2	25.1	26.3	10.9	8.3
189	65.8	65.8	29.3	29.3	4.3	4.3

6.2 Appendix B – HRT study

6.2.1 Influent and effluent solids, total and soluble COD, Alkalinity and pH

6.2.1.1 Influent and effluent solids

Solids Feed						
	TSS (g/L)			VSS (g/L)		
Day	F1	F2	F3	F1	F2	F3
0	0.7	0.7	0.7	0.7	0.1	0.7
2	3.7	3.4	3.7	3.4	0.1	3.4
4	0.2	0.2	0.2	0.2	0.1	0.2
7	0.1	0.1	0.1	0.1	0.1	0.1
9	1.7	0.6	0.7	1.6	0.1	0.7
11	0.7	0.3	0.3	0.7	0.1	0.3
14	0.3	0.2	0.2	0.3	0.1	0.2
17	0.1	0.1	0.1	0.1	0.1	0.1
21	0.1	0.1	0.1	0.1	0.1	0.1
24	0.1	0.1	0.1	0.1	0.1	0.1
28	0.2	0.3	0.2	0.2	0.1	0.2
31	0.1	0.1	0.1	0.1	0.1	0.1
35	0.2	0.1	0.2	0.2	0.1	0.2
38	0.3	0.3	0.3	0.3	0.1	0.2
42	0.2	0.1	0.3	0.2	0.1	0.3
45	0.1	0.1	0.3	0.1	0.1	2.3
49	0.2	0.1	0.5	0.2	0.1	0.5
52	0.2	0.1	0.2	0.1	0.1	0.1
56	0.3	0.2	0.1	0.2	0.1	0.1
59	0.3	0.2	0.2	0.3	0.1	0.2
63	0.3	0.0	0.3	0.3	0.1	0.2
66	0.1	0.1	0.1	0.1	0.2	0.1
70	2.2	0.7	0.8	2.0	0.1	0.8
73	0.2	0.1	0.3	0.2	0.1	0.3
77	1.4	0.2	0.9	1.3	0.1	0.9
80	0.4	0.1	0.1	0.4	0.1	0.0
84	0.2	0.2	0.7	0.2	0.1	0.6
87	0.1	0.1	0.1	0.1	0.1	0.2
91	0.9	0.3	0.6	0.8	0.1	0.6
94	0.1	0.1	0.1	0.1	0.1	0.1
98	0.3	0.3	0.3	0.3	0.1	0.2
101	0.3	0.1	0.2	0.3	0.1	0.2
105	1.2	0.2	0.2	1.1	0.1	0.2

Solids Feed						
	TSS (g/L)			VSS (g/L)		
Day	F1	F2	F3	F1	F2	F3
108	0.2	0.1	0.2	0.1	0.2	0.3
112	0.5	0.1	0.4	0.4	0.1	0.3
115	0.1	0.6	0.4	0.0	0.1	0.5
119	0.9	0.2	0.4	0.9	0.1	0.4
122	1.6	1.1	1.3	1.5	0.1	1.2
129	2.6	0.2	0.7	2.4	0.1	0.7
133	0.5	0.2	0.3	0.4	0.1	0.3
136	0.2	0.1	0.2	0.2	0.1	0.2
140	0.4	0.3	0.2	0.5	0.1	0.2
143	1.1	1.0	1.4	1.0	0.1	1.1
147	0.5	0.4	1.0	0.5	0.1	0.9
150	0.4	0.1	0.4	0.3	0.1	0.3
154	0.6	0.3	0.3	0.5	0.1	0.2
157	1.5	0.4	0.5	1.4	0.1	0.5
161	0.1	0.0	0.2	0.1	0.0	0.1

6.2.1.2 Effluent solids

Solids Effluent						
	TSS (g/L)			VSS (g/L)		
Day	R1	R2	R3	R1	R2	R3
0	0.4	0.1	0.2	0.4	0.2	0.2
2	0.1	0.1	0.1	0.1	0.1	0.1
4	0.2	0.1	0.2	0.1	0.1	0.2
7	0.2	0.1	0.2	0.2	0.1	0.2
9	0.5	0.4	1.0	0.4	0.4	0.9
11	0.2	0.2	0.4	0.3	0.3	0.4
14	0.2	0.2	0.2	0.2	0.2	0.2
17	0.2	0.2	0.2	0.1	0.2	0.1
21	0.1	0.1	0.1	0.1	0.1	0.1
24	0.1	0.2	0.5	0.1	0.1	0.5
28	0.3	0.4	0.2	0.3	0.4	0.2
31	0.1	0.1	0.2	0.1	0.1	0.2
35	0.1	0.1	0.1	0.1	0.1	0.1
38	0.2	0.1	0.1	0.2	0.1	0.1

Solids Effluent						
	TSS (g/L)			VSS (g/L)		
Day	R1	R2	R3	R1	R2	R3
42	0.1	0.3	0.3	0.2	0.3	0.3
45	0.1	0.1	0.1	0.1	0.1	0.1
49	0.2	0.2	0.2	0.2	0.2	0.2
52	0.3	0.2	0.2	0.2	0.2	0.1
56	0.2	0.2	0.2	0.2	0.2	0.2
59	0.2	0.2	0.2	0.2	0.1	0.1
63	0.2	0.2	0.2	0.1	0.1	0.1
66	0.1	0.2	0.2	0.2	0.2	0.2
70	0.3	0.2	0.3	0.3	0.2	0.3
73	0.3	0.2	0.3	0.3	0.2	0.3
77	0.4	0.1	0.2	0.4	0.1	0.2
80	0.2	0.2	0.1	0.2	0.2	0.2
84	0.2	0.4	0.4	0.2	0.4	0.4
87	0.2	0.1	0.1	0.2	0.1	0.1
91	0.3	0.2	0.4	0.3	0.2	0.4
94	0.2	0.1	0.2	0.2	0.1	0.2
98	0.5	0.2	0.3	0.4	0.2	0.3
101	0.2	0.2	0.2	0.2	0.2	0.2
105	0.6	0.8	0.5	0.5	0.5	0.4
108	0.2	0.1	0.2	0.2	0.1	0.2
112	0.3	0.3	0.3	0.2	0.3	0.2
115	0.3	0.5	1.7	0.3	0.6	1.4
119	0.3	0.3	0.5	0.3	0.4	0.5
122	0.7	0.3	0.4	0.6	0.2	0.4
129	0.3	0.5	0.8	0.3	0.5	0.7
133	0.2	0.3	0.2	0.2	0.3	0.2
136	0.3	0.2	0.3	0.2	0.2	0.3
140	0.2	0.1	0.2	0.2	0.1	0.2
143	0.3	0.3	0.7	0.3	0.3	0.6
147	0.2	0.2	0.2	0.2	0.2	0.2
150	0.2	0.2	0.4	0.2	0.1	0.4
154	0.3	0.4	0.3	0.3	0.3	0.3
157	0.4	0.3	0.4	0.4	0.3	0.4
161	0.2	0.1	0.3	0.2	0.1	0.2

6.2.1.3 Influent soluble and total COD

	sCODin				tCODin			
	[g/L]	[g/L]	[g/L]		[g/L]	[g/L]	[g/L]	
Day	R1	R2	R3	ave	R1	R2	R3	ave
0	3.0	3.0	3.0	3.0	5.7	5.7	5.7	5.7
2	3.0	3.0	3.0	3.0	5.7	5.7	5.7	5.7
4	3.2	3.2	3.2	3.2	3.9	3.9	3.9	3.9
7	2.9	2.9	2.9	2.9	3.1	3.1	3.1	3.1
9	3.5	3.5	3.4	3.5	5.8	4.8	5.1	5.2
11	3.8	3.7	3.7	3.8	7.3	6.5	5.2	6.3
14	3.6	3.6	3.7	3.6	4.0	3.8	4.0	3.9
16	3.3	3.4	3.4	3.3	4.5	3.7	3.7	4.0
18	3.3	3.4	3.3	3.3	3.9	3.5	3.5	3.6
21	3.0	3.1	3.1	3.1	3.3	3.3	3.3	3.3
23	3.0	2.9	3.0	3.0	3.0	3.3	3.2	3.2
25	2.6	2.6	2.6	2.6	2.8	2.8	2.7	2.8
28	3.0	2.8	2.8	2.9	3.3	4.1	3.0	3.4
30	2.9	2.8	2.8	2.9	3.1	3.1	3.5	3.3
32	2.9	2.9	2.9	2.9	3.5	3.1	3.1	3.2
35	2.9	2.8	2.8	2.8	2.9	2.9	2.9	2.9
37	2.6	2.7	2.7	2.7	2.8	2.8	2.9	2.8
39	2.9	3.1	2.9	3.0	4.3	6.7	5.7	5.5
42	3.0	2.9	2.9	2.9	4.1	6.5	6.5	5.7
44	3.0	3.0	3.0	3.0	3.9	3.2	3.3	3.5
46	3.1	3.1	3.0	3.1	3.9	3.2	3.2	3.5
49	2.9	2.8	3.0	2.9	3.2	2.8	3.0	3.0
51	3.1	3.3	3.4	3.2	4.3	5.0	4.3	4.6
53	2.7	2.5	2.6	2.6	4.7	3.6	3.6	4.0
56	2.4	2.6	2.7	2.6	3.0	3.1	2.8	3.0
58	3.2	2.8	2.6	2.9	3.3	3.1	3.3	3.2
60	3.5	3.3	3.3	3.3	7.5	7.9	8.0	7.8
63	3.4	3.3	3.6	3.4	3.8	3.6	4.8	4.1
65	3.5	3.3	3.4	3.4	3.6	3.6	3.9	3.7
67	3.2	3.4	3.4	3.3	3.6	3.7	3.4	3.6
70	4.1	3.6	4.2	4.0	7.1	4.8	6.1	6.0
72	3.7	3.9	4.3	3.9	4.5	3.9	4.3	4.3
74	3.5	3.5	3.5	3.5	4.3	3.8	3.9	4.0
77	3.8	3.7	3.7	3.7	10.4	4.7	8.6	7.9
79	3.7	4.7	2.4	3.6	4.5	7.9	5.1	5.8
81	5.6	5.3	5.3	5.4	6.2	5.9	5.9	6.0
84	5.2	5.3	5.3	5.3	8.0	5.6	5.7	6.4

	sCODin				tCODin			
	[g/L]	[g/L]	[g/L]		[g/L]	[g/L]	[g/L]	
Day	R1	R2	R3	ave	R1	R2	R3	ave
86	5.3	5.1	5.1	5.2	5.6	5.9	5.6	5.7
88	4.7	4.6	4.7	4.7	4.8	4.8	4.9	4.8
91	5.9	6.0	5.8	5.9	9.1	7.3	7.9	8.1
93	5.7	5.6	5.4	5.6	5.9	6.0	5.8	5.9
95	5.2	5.3	5.3	5.3	5.5	5.5	5.5	5.5
98	5.6	5.8	5.8	5.7	6.7	6.5	6.6	6.6
100	3.9	3.9	3.9	3.9	4.7	4.2	4.3	4.4
102	5.6	5.7	5.7	5.7	8.4	7.9	8.1	8.2
105	3.2	3.0	3.0	3.1	4.8	3.5	3.3	3.9
107	4.2	4.1	4.0	4.1	5.0	4.4	5.1	4.8
109	4.8	4.9	4.5	4.8	6.8	6.9	6.2	6.6
112	4.6	4.7	4.6	4.6	5.5	4.7	5.6	5.3
114	5.8	5.1	5.3	5.4	8.6	7.0	6.7	7.4
116	5.0	5.2	5.2	5.1	9.7	6.0	8.1	8.0
119	4.6	4.9	4.9	4.8	7.7	5.5	6.4	6.5
121	4.5	4.5	5.1	4.7	9.0	9.1	9.5	9.2
123	5.6	5.4	5.7	5.6	11.1	10.1	10.5	10.6
126	5.2	5.3	5.6	5.4	9.1	5.3	5.8	6.7
128	3.8	3.4	3.6	3.6	10.1	4.7	5.9	6.9
130	3.5	3.1	2.9	3.2	4.9	3.7	4.3	4.3
133	3.7	3.4	3.6	3.6	4.7	4.0	3.9	4.2
135	3.7	3.6	3.4	3.6	3.6	3.6	3.6	3.6
137	3.6	3.1	3.3	3.4	4.1	3.8	4.1	4.0
140	3.5	3.5	3.4	3.5	4.0	4.2	4.2	4.1
142	3.8	3.5	4.0	3.8	6.3	5.1	5.6	5.7
144	3.8	3.3	3.7	3.6	15.8	15.7	15.1	15.5
147	3.8	3.5	3.6	3.7	5.3	4.8	5.5	5.2
149	3.3	3.4	3.6	3.4	4.8	4.2	4.3	4.4
151	3.8	3.5	3.9	3.7	5.2	6.2	5.9	5.8
154	4.0	3.6	3.5	3.7	5.0	4.2	4.1	4.4
156	4.2	3.6	3.9	3.9	7.8	4.8	5.1	5.9
158	3.8	3.6	3.9	3.8	4.4	4.8	4.5	4.6
161	3.7	3.4	3.4	3.5	4.1	3.6	3.8	3.8

6.2.1.4 Effluent soluble and total COD

	sCODout				tCODout			
	[g/L]	[g/L]	[g/L]		[g/L]	[g/L]	[g/L]	
Day	R1	R2	R3	ave	R1	R2	R3	ave
0	0.2	0.3	0.2	0.2	0.9	0.5	0.6	0.7
2	0.2	0.2	0.2	0.2	0.5	0.5	0.4	0.5
4	0.2	0.2	0.2	0.2	0.5	0.5	0.5	0.5
7	0.2	0.3	0.2	0.2	0.4	0.4	0.5	0.4
9	0.3	0.4	0.3	0.3	1.1	1.2	1.9	1.4
11	0.3	0.4	0.3	0.3	0.7	0.9	1.1	0.9
14	0.2	0.3	0.2	0.2	0.5	0.7	0.5	0.5
16	0.2	0.3	0.2	0.2	0.4	0.7	0.5	0.5
18	0.2	0.3	0.2	0.2	0.3	0.7	0.4	0.5
21	0.2	0.2	0.2	0.2	0.4	0.5	0.4	0.4
23	0.2	0.4	0.2	0.3	0.3	0.6	0.7	0.5
25	0.1	0.2	0.2	0.2	0.2	0.3	0.4	0.3
28	0.2	0.2	0.1	0.2	0.7	0.9	0.5	0.7
30	0.1	0.2	0.1	0.1	0.5	0.5	0.9	0.6
32	0.1	0.2	0.1	0.2	0.6	0.4	0.4	0.4
35	0.1	0.1	0.1	0.1	0.5	0.5	0.4	0.4
37	0.1	0.2	0.1	0.1	0.8	0.4	0.3	0.5
39	0.2	0.3	0.2	0.2	0.6	0.6	0.4	0.5
42	0.2	0.3	0.2	0.2	0.4	0.8	0.5	0.6
44	0.1	0.2	0.1	0.2	0.4	0.6	0.3	0.4
46	0.2	0.3	0.2	0.2	0.3	0.6	0.3	0.4
49	0.2	0.3	0.2	0.2	0.4	0.6	0.4	0.5
51	0.2	0.2	0.2	0.2	0.6	0.5	0.5	0.5
53	0.3	0.3	0.2	0.3	0.4	0.6	0.5	0.5
56	0.2	0.4	0.3	0.3	0.5	0.6	0.6	0.6
58	0.2	0.6	0.2	0.4	0.4	0.8	0.5	0.6
60	0.3	0.6	0.5	0.4	1.2	1.0	0.8	1.0
63	0.3	0.6	0.3	0.4	0.6	1.0	0.4	0.7
65	0.3	0.6	0.3	0.4	0.6	0.7	0.7	0.7
67	0.3	0.5	0.3	0.4	0.6	0.8	0.6	0.7
70	0.8	1.0	0.7	0.9	1.1	1.5	1.0	1.2
72	0.4	0.7	0.3	0.5	0.7	0.8	0.7	0.8
74	0.4	0.7	0.3	0.5	0.8	1.1	0.8	0.9
77	0.5	0.9	0.4	0.6	0.0	1.5	1.8	1.1
79	0.5	0.5	0.6	0.5	1.6	1.1	1.0	1.3
81	0.5	1.0	0.3	0.6	0.4	1.5	0.6	0.8

	sCODin				tCODin			
	[g/L]	[g/L]	[g/L]		[g/L]	[g/L]	[g/L]	
Day	R1	R2	R3	ave	R1	R2	R3	ave
84	0.5	0.9	0.5	0.6	1.4	1.5	1.1	1.3
86	5.3	5.1	5.1	5.2	5.6	5.9	5.6	5.7
88	4.7	4.6	4.7	4.7	4.8	4.8	4.9	4.8
91	5.9	6.0	5.8	5.9	9.1	7.3	7.9	8.1
93	5.7	5.6	5.4	5.6	5.9	6.0	5.8	5.9
95	5.2	5.3	5.3	5.3	5.5	5.5	5.5	5.5
98	5.6	5.8	5.8	5.7	6.7	6.5	6.6	6.6
100	3.9	3.9	3.9	3.9	4.7	4.2	4.3	4.4
102	5.6	5.7	5.7	5.7	8.4	7.9	8.1	8.2
105	3.2	3.0	3.0	3.1	4.8	3.5	3.3	3.9
107	4.2	4.1	4.0	4.1	5.0	4.4	5.1	4.8
109	4.8	4.9	4.5	4.8	6.8	6.9	6.2	6.6
112	4.6	4.7	4.6	4.6	5.5	4.7	5.6	5.3
114	5.8	5.1	5.3	5.4	8.6	7.0	6.7	7.4
116	5.0	5.2	5.2	5.1	9.7	6.0	8.1	8.0
119	4.6	4.9	4.9	4.8	7.7	5.5	6.4	6.5
121	4.5	4.5	5.1	4.7	9.0	9.1	9.5	9.2
123	5.6	5.4	5.7	5.6	11.1	10.1	10.5	10.6
126	5.2	5.3	5.6	5.4	9.1	5.3	5.8	6.7
128	3.8	3.4	3.6	3.6	10.1	4.7	5.9	6.9
130	3.5	3.1	2.9	3.2	4.9	3.7	4.3	4.3
133	3.7	3.4	3.6	3.6	4.7	4.0	3.9	4.2
135	3.7	3.6	3.4	3.6	3.6	3.6	3.6	3.6
137	3.6	3.1	3.3	3.4	4.1	3.8	4.1	4.0
140	3.5	3.5	3.4	3.5	4.0	4.2	4.2	4.1
142	3.8	3.5	4.0	3.8	6.3	5.1	5.6	5.7
144	3.8	3.3	3.7	3.6	15.8	15.7	15.1	15.5
147	3.8	3.5	3.6	3.7	5.3	4.8	5.5	5.2
149	3.3	3.4	3.6	3.4	4.8	4.2	4.3	4.4
151	3.8	3.5	3.9	3.7	5.2	6.2	5.9	5.8
154	4.0	3.6	3.5	3.7	5.0	4.2	4.1	4.4
156	4.2	3.6	3.9	3.9	7.8	4.8	5.1	5.9
158	3.8	3.6	3.9	3.8	4.4	4.8	4.5	4.6
161	3.7	3.4	3.4	3.5	4.1	3.6	3.8	3.8

6.2.1.5 Alkalintiy and pH

	Alkalinity [g/L]			pH Effluent			pH Influent		
Day	R1	R2	R3	R1	R2	R3	F1	F2	F3
0	1.7	1.4	1.4	6.8	6.8	6.8	6.4	6.4	6.4
2	1.6	1.4	1.4	6.9	6.9	7.0	6.6	6.6	6.6
4	1.6	1.5	1.5	7.4	7.3	7.3	6.8	6.8	6.8
7	1.5	1.5	1.5	7.4	7.2	7.2	6.9	6.9	6.9
9	1.5	1.6	1.6	7.1	7.0	7.1	6.7	6.7	6.8
11	1.7	1.7	1.6	7.1	7.1	7.1	6.6	6.7	6.6
14	1.6	1.6	1.6	7.0	7.1	7.1	6.6	6.7	6.7
16	1.6	1.5	1.6	7.1	7.1	7.1	6.5	6.6	6.6
18	1.6	1.7	1.7	7.1	7.1	7.1	6.7	6.7	6.7
21	1.6	1.7	1.7	7.1	7.1	7.1	6.7	6.7	6.7
23	1.7	1.5	1.6	7.1	7.1	7.1	6.9	6.9	6.8
25	1.7	1.6	1.6	7.1	7.2	7.2	6.6	6.6	6.6
28	1.5	1.5	1.6	7.1	7.0	7.1	6.8	6.8	6.8
30	1.5	1.6	1.6	7.1	7.1	7.2	6.7	6.7	6.7
32	1.6	1.6	1.6	7.2	7.2	7.3	6.6	6.6	6.7
35	1.5	1.6	1.5	7.3	7.1	7.2	6.7	6.6	6.7
37	1.6	1.6	1.6	7.4	7.3	7.3	6.6	6.6	6.6
39	1.6	1.5	1.6	7.2	6.9	7.1	6.5	6.5	6.5
42	1.5	1.5	1.6	7.0	7.0	7.1	6.7	6.6	6.7
44	1.5	1.6	1.6	7.2	7.2	7.2	6.8	6.8	6.8
46	1.6	1.5	1.6	7.2	7.2	7.2	6.8	6.8	6.9
49	1.6	1.5	1.6	7.2	7.1	7.2	6.9	6.9	6.9
51	1.7	1.6	1.6	7.3	7.2	7.2	6.9	6.8	6.8
53	1.5	1.6	1.6	7.2	7.1	7.2	6.9	6.9	6.9
56	1.6	1.6	1.7	7.2	7.2	7.2	6.8	6.8	6.9
58	1.6	1.5	1.6	7.2	7.0	7.2	6.8	6.8	6.9
60	1.5	1.4	1.5	7.1	6.9	7.0	6.9	7.0	7.0
63	1.6	1.5	1.6	7.2	7.0	7.1	6.7	6.8	6.8
65	1.5	1.5	1.6	7.1	7.0	7.1	6.8	6.8	6.9
67	1.5	1.5	1.6	7.2	7.1	7.2	7.2	7.0	7.0
70	1.6	1.6	1.6	7.1	6.9	7.1	6.7	6.7	6.6
72	1.6	1.5	1.6	7.0	6.9	7.0	6.8	6.8	6.8
74	1.6	1.5	1.6	7.0	6.8	7.0	6.8	6.8	6.8
77	1.6	1.5	1.7	7.2	6.9	7.1	6.7	6.7	6.8
79	1.5	1.4	1.6	7.2	6.9	7.1	6.6	6.7	6.7
81	1.6	1.5	1.6	7.4	7.0	7.3	6.8	6.9	6.8
84	1.6	1.4	1.5	7.2	7.1	7.2	6.7	6.8	6.8
86	1.6	1.4	1.6	7.1	6.9	7.2	6.8	6.8	6.8

	Alkalinity [g/L]			pH Effluent			pH Influent		
Day	R1	R2	R3	R1	R2	R3	F1	F2	F3
88	1.6	1.5	1.6	7.2	7.1	7.3	7.0	7.1	7.1
91	1.7	1.5	1.7	7.1	6.9	7.2	6.7	6.8	6.8
93	1.7	1.5	1.7	7.3	7.0	7.2	6.6	6.7	6.7
95	1.6	1.5	1.7	7.2	6.9	7.2	6.7	6.7	6.7
98	1.6	1.4	1.7	7.0	6.8	7.0	6.6	6.6	6.7
100	1.6	1.5	1.6	6.9	6.8	6.9	6.5	6.6	6.6
102	1.6	1.5	1.7	7.0	6.9	7.0	6.8	6.8	6.8
105	1.5	1.4	1.5	6.9	6.8	7.0	6.5	6.6	6.6
107	1.5	1.4	1.5	7.1	6.9	7.0	6.7	6.8	6.8
109	1.4	1.3	1.4	7.0	6.7	6.8	6.8	6.8	6.9
112	1.5	1.3	1.4	6.9	6.7	6.9	6.6	6.7	6.7
114	1.4	1.3	1.4	6.8	6.6	6.8	6.8	6.8	6.8
116	1.5	1.3	1.5	7.0	6.7	6.8	6.7	6.7	6.7
119	1.4	1.4	1.4	6.8	6.7	6.8	6.6	6.7	6.7
121	1.4	1.4	1.4	6.9	6.7	6.8	6.7	6.7	6.7
123	1.4	1.4	1.5	6.8	6.5	6.7	6.6	6.7	6.7
126	1.4	1.4	1.5	7.1	6.7	6.9	6.7	6.7	6.7
128	1.9	1.9	2.0	7.1	6.8	6.9	6.9	7.0	6.9
130	2.0	1.9	2.1	7.2	7.0	7.1	6.9	6.9	6.9
133	2.0	2.0	1.9	7.1	6.8	6.8	6.6	6.7	6.7
135	2.0	1.9	1.8	6.9	6.9	6.9	6.9	6.9	6.9
137	1.7	1.7	1.7	7.1	7.0	7.0	6.6	6.7	6.7
140	1.9	2.0	1.8	7.1	6.8	7.0	6.7	6.7	6.7
142	1.9	1.8	1.9	7.0	6.8	6.8	6.8	6.7	6.8
144	1.9	1.8	2.0	6.9	6.8	7.0	6.7	6.8	6.7
147	1.7	1.8	1.7	7.1	7.0	7.0	6.6	6.6	6.7
149	1.6	1.8	1.8	7.0	6.8	7.0	6.7	6.8	6.7
151	1.4	1.8	1.8	7.3	6.9	7.0	6.8	6.7	6.7
154	1.8	1.8	1.8	6.9	6.8	6.9	6.8	6.9	6.9
156	1.7	1.9	1.9	7.0	6.9	7.0	6.8	6.8	6.8
158	1.5	1.7	1.8	7.1	7.0	7.0	6.8	6.9	7.0
161	1.3	1.4	1.5	7.2	7.1	7.1	6.9	6.9	7.0

6.2.2 Total and soluble COD removal, OLR and SSAA

6.2.2.1 Alkalintiy and pH

	tCOD rem			sCOD removal		
	[%]			[%]		
Day	R1	R2	R3	R1	R2	R3
0	84.5	90.5	89.7	94.4	91.5	94.0
2	91.1	91.8	92.5	94.3	92.3	94.9
4	86.6	87.6	86.4	94.3	92.7	94.0
7	86.4	86.4	84.9	93.7	90.1	94.2
9	80.7	75.8	61.6	91.9	89.6	91.4
11	91.0	87.0	79.7	91.9	88.9	91.4
14	88.3	82.5	87.2	93.8	92.7	93.9
16	90.8	82.2	87.6	94.3	92.1	94.8
18	91.6	80.3	88.7	94.8	91.5	94.6
21	87.4	85.3	87.0	94.5	93.1	94.6
23	89.6	82.9	79.8	93.6	88.1	92.5
25	94.0	89.7	86.6	94.8	94.0	93.6
28	78.3	76.6	84.9	94.2	92.6	94.8
30	84.4	83.2	73.5	95.5	94.5	95.5
32	83.8	87.9	88.1	95.1	92.7	95.7
35	83.6	83.3	87.1	94.9	94.7	95.2
37	72.3	84.8	89.1	95.1	94.0	95.8
39	86.5	91.0	92.8	93.4	90.4	94.2
42	96.4	87.1	91.5	94.2	89.4	93.4
44	88.9	81.9	90.6	95.1	91.8	95.2
46	92.0	80.6	90.8	94.6	89.5	94.8
49	86.5	79.2	86.0	94.0	88.9	94.3
51	86.8	89.9	89.1	94.6	94.7	93.4
53	91.7	83.4	85.8	90.0	86.6	92.1
56	81.7	82.4	80.1	92.0	86.3	87.6
58	86.6	75.1	86.3	92.8	79.5	90.4
60	83.5	87.9	90.1	91.7	82.7	85.3
63	83.8	73.2	90.9	91.2	82.8	92.2
65	84.4	80.6	82.6	92.0	81.0	92.3
67	84.6	78.1	82.5	91.9	85.8	90.5
70	84.5	67.9	83.0	79.8	71.4	82.9
72	84.4	78.8	82.7	89.4	81.0	93.9
74	82.3	71.4	80.0	89.6	78.9	92.3
77	80.8	68.8	79.1	86.6	75.3	90.3
79	64.2	85.5	79.8	86.1	88.9	76.3

	tCOD rem			sCOD removal		
	[%]			[%]		
Day	R1	R2	R3	R1	R2	R3
81	92.7	75.1	89.4	91.6	81.2	93.9
84	82.2	73.7	81.4	90.4	83.3	89.9
86	81.0	70.6	82.0	90.2	80.6	91.3
88	79.9	66.2	82.9	88.3	78.4	90.1
91	82.8	72.4	83.3	85.7	83.4	86.2
93	82.1	76.2	85.4	90.4	82.6	89.4
95	76.4	74.3	79.2	89.0	80.9	92.0
98	76.7	71.5	75.4	86.6	82.6	87.4
100	78.6	70.1	80.4	84.8	72.8	87.5
102	88.5	82.2	86.5	87.8	79.7	90.0
105	75.7	66.8	67.5	83.7	74.1	72.1
107	70.7	77.6	72.6	89.6	83.7	85.8
109	70.6	70.9	68.3	86.2	79.6	86.5
112	78.1	67.5	80.7	86.9	78.2	83.7
114	69.5	60.9	66.6	81.7	76.6	84.2
116	86.9	70.3	77.7	86.6	76.6	86.2
119	84.2	68.9	73.7	84.4	75.5	85.8
121	89.7	88.5	91.4	89.0	83.9	90.3
123	80.4	78.5	79.9	77.5	76.9	77.7
126	81.0	64.4	81.7	84.9	75.0	87.6
128	77.6	46.9	52.9	70.6	53.6	62.8
130	77.4	50.1	56.9	79.3	64.9	71.2
133	68.4	58.3	61.8	72.5	62.9	63.3
135	60.4	46.4	56.2	78.7	63.6	69.0
137	70.8	54.7	64.8	74.3	55.7	66.6
140	70.7	63.4	56.6	73.8	61.1	67.5
142	74.7	60.3	61.8	70.4	56.5	67.4
144	90.9	89.5	88.8	68.3	57.4	65.1
147	72.5	60.8	66.0	72.7	56.1	61.8
149	39.7	42.7	49.4	42.7	48.2	53.1
151	68.0	73.0	68.0	68.2	44.9	62.8
154	64.8	50.6	48.2	70.0	55.0	61.0
156	70.3	41.9	49.6	66.4	41.7	58.3
158	60.6	54.5	44.5	68.7	54.9	62.1
161	63.1	46.0	51.9	71.0	48.4	61.4

6.2.2.2 Organic loading rates and SSAA

	t OLR			sOLR			tSSAA			sSSAA		
	[g tCOD/L d]			[g sCOD/L d]			[g sCOD/m^2 d]			[g tCOD/m^2 d]		
Day	R1	R2	R3	R1	R2	R3	R1	R2	R3	R1	R2	R3
0	5.7	5.7	5.7	3.0	3.0	3.0	28.3	16.3	16.7	16.8	30.3	30.1
2	5.7	5.7	5.7	3.0	3.0	3.0	30.5	16.4	16.9	16.8	30.8	31.0
4	3.9	3.9	3.9	3.2	3.2	3.2	20.0	17.7	17.9	18.0	20.2	20.0
7	3.1	3.1	3.1	2.9	2.9	2.9	15.9	15.4	16.1	16.0	15.9	15.7
9	5.8	4.8	5.1	3.5	3.5	3.4	27.4	18.4	18.4	18.9	21.2	18.4
11	7.3	6.5	5.2	3.8	3.7	3.7	39.1	19.4	20.1	20.7	33.5	24.4
14	4.0	3.8	4.0	3.6	3.6	3.7	20.9	19.8	20.3	19.9	18.5	20.5
16	4.5	3.7	3.7	3.3	3.4	3.4	24.2	18.3	19.0	18.1	17.8	19.3
18	3.9	3.5	3.5	3.3	3.4	3.3	20.8	18.1	18.3	18.6	16.5	18.5
21	3.3	3.3	3.3	3.0	3.1	3.1	16.7	17.1	17.3	16.9	16.7	17.0
23	3.0	3.3	3.2	3.0	2.9	3.0	16.0	15.2	16.1	16.4	15.9	15.1
25	2.8	2.8	2.7	2.6	2.6	2.6	15.7	14.6	14.3	14.7	14.5	13.9
28	3.3	4.1	3.0	3.0	2.8	2.8	15.0	15.4	15.8	16.4	18.2	14.9
30	3.1	3.1	3.5	2.9	2.8	2.8	15.4	15.8	15.9	16.4	15.4	15.2
32	4.7	4.1	4.1	3.9	3.8	3.8	17.5	21.0	21.6	21.6	21.1	16.0
35	3.9	3.8	3.8	3.8	3.7	3.7	14.4	20.8	20.8	21.4	18.7	14.6
37	3.7	3.8	3.8	3.5	3.6	3.6	11.9	19.9	20.5	19.6	18.9	15.0
39	4.5	8.9	7.6	3.9	4.1	3.9	21.7	21.9	21.6	21.4	47.9	31.0
42	9.3	8.6	8.6	4.0	3.9	3.9	59.2	20.4	21.3	21.9	44.0	34.7
44	3.0	4.2	4.4	4.0	4.0	4.0	20.5	21.6	22.2	22.3	20.4	17.8
46	2.6	4.3	4.3	4.2	4.1	4.0	21.3	21.4	22.5	23.2	20.4	17.2
49	4.3	3.7	4.0	3.8	3.7	4.0	16.5	17.4	15.1	21.2	19.4	22.1
51	5.8	6.6	5.8	4.1	4.4	4.5	22.2	35.1	22.7	23.1	24.3	24.6
53	6.3	3.9	4.8	3.5	2.7	3.4	25.5	23.7	18.3	18.8	17.0	18.7
56	5.9	4.2	3.7	4.8	3.5	3.6	14.4	20.2	13.1	26.2	17.6	18.4
58	6.5	4.2	4.4	6.3	3.8	3.5	16.7	18.5	16.7	34.4	17.7	18.4
60	14.8	10.6	10.7	6.8	4.4	4.4	36.6	54.7	42.5	36.9	21.4	21.9
63	7.6	4.8	6.4	6.7	4.4	4.7	19.0	20.6	25.8	36.0	21.6	25.6
65	7.1	7.2	7.8	6.9	6.5	6.7	17.8	34.2	19.1	37.3	30.9	36.2
67	7.1	7.4	6.8	6.4	6.7	6.7	17.9	34.1	16.6	34.5	34.1	35.5
70	14.1	9.5	12.0	8.1	7.1	8.4	35.3	37.9	29.7	37.9	29.9	41.0
72	9.0	7.8	8.6	7.3	7.7	8.5	22.6	36.0	21.0	38.2	36.6	46.7
74	8.6	7.5	7.7	6.9	7.0	6.8	20.9	31.4	18.4	36.6	32.3	37.1
77	20.5	9.3	17.0	7.4	7.4	7.4	49.3	37.6	39.8	37.9	32.7	39.4
79	8.8	15.7	10.1	7.4	9.4	4.8	22.8	54.3	32.5	22.8	33.5	14.9

	tOLR			sOLR			tSSAA			sSSAA		
	[g tCOD/L d]			[g sCOD/L d]			[g sCOD/m^2 d]			[g tCOD/m^2 d]		
Day	**R1**	**R2**	**R3**	**R1**	**R2**	**R3**	**R1**	**R2**	**R3**	**R1**	**R2**	**R3**
81	12.3	11.8	11.7	11.2	10.4	10.5	45.8	35.6	42.1	45.8	34.1	39.8
84	15.9	11.1	11.2	10.3	10.6	10.5	52.7	33.0	36.8	52.7	35.5	38.0
86	13.4	13.7	12.9	12.6	11.8	11.8	43.8	38.9	42.7	43.8	38.3	43.6
88	11.4	11.0	11.4	11.1	10.7	10.8	36.6	29.4	38.1	36.6	33.9	39.2
91	21.6	16.7	18.3	14.0	13.8	13.3	72.1	48.8	61.5	72.1	46.6	46.2
93	14.1	13.8	13.4	13.6	12.8	12.5	46.8	42.4	46.2	46.8	42.7	45.3
95	13.0	12.6	12.8	12.5	12.1	12.2	40.0	37.8	40.8	40.0	39.5	45.4
98	16.0	14.9	15.1	13.4	13.3	13.4	49.3	43.0	45.9	49.3	44.2	47.2
100	11.2	9.8	10.0	9.4	9.0	8.9	35.7	27.6	32.4	35.7	26.4	31.5
102	20.0	18.3	18.7	13.3	13.1	13.2	71.4	60.6	65.1	71.4	42.1	47.8
105	11.4	8.0	7.6	7.6	6.9	6.9	35.0	21.6	20.8	35.0	20.7	20.1
107	15.0	13.3	14.9	12.6	12.3	11.9	42.9	41.5	43.6	42.9	41.4	41.1
109	20.5	20.7	18.3	14.6	14.8	13.2	58.4	59.1	50.5	58.4	47.3	46.2
112	16.6	14.0	16.7	13.9	13.9	13.5	52.4	38.2	54.2	52.4	44.0	45.5
114	25.9	20.9	19.7	17.5	15.2	15.7	72.6	51.4	52.9	72.6	47.0	53.3
116	29.4	18.1	23.8	15.1	15.6	15.5	103.1	51.2	74.7	103.1	48.3	53.7
119	23.2	16.4	18.8	14.0	14.5	14.6	78.7	45.5	55.8	78.7	44.3	50.4
121	27.2	27.2	27.9	13.7	13.4	15.1	98.3	97.0	102.9	98.3	45.2	55.1
123	33.6	30.3	31.1	16.8	16.2	16.8	108.9	95.8	100.2	108.9	50.3	52.6
126	27.5	15.8	17.0	15.8	15.7	16.6	89.7	41.1	56.0	54.2	47.6	58.5
128	30.6	13.9	17.4	11.5	10.0	10.7	95.8	26.3	37.1	32.7	21.6	27.1
130	19.4	14.5	16.4	14.0	12.2	11.0	60.4	29.4	37.6	44.8	32.0	31.5
133	18.7	15.7	14.7	14.8	13.5	13.4	51.6	37.0	36.6	43.3	34.1	34.3
135	14.4	14.1	13.5	14.7	14.1	13.0	35.1	26.4	30.6	46.5	36.1	36.1
137	16.3	15.2	15.5	14.4	12.4	12.5	46.6	33.4	40.4	43.1	27.9	33.6
140	15.8	16.6	15.7	13.8	13.9	13.0	44.9	42.6	35.9	41.0	34.1	35.5
142	25.0	20.3	21.1	15.1	14.0	15.0	75.3	49.3	52.7	42.8	31.8	40.8
144	62.6	62.3	56.9	15.0	13.1	13.9	-	-	-	41.4	30.3	36.5
147	21.1	19.1	20.8	15.0	14.0	13.8	61.8	46.9	55.4	44.1	31.8	34.3
149	18.9	16.5	16.3	13.0	13.6	13.6	30.3	28.4	32.6	22.4	26.4	29.1
151	20.8	24.6	22.3	15.1	13.9	14.8	56.9	72.5	61.0	41.6	25.2	37.5
154	19.6	16.6	15.4	15.7	14.4	13.4	51.3	33.9	30.0	44.4	32.0	33.0
156	30.8	19.2	19.5	16.8	14.3	14.7	87.2	32.4	39.0	44.9	24.1	34.5
158	17.5	19.1	16.9	15.1	14.3	14.7	42.7	41.9	30.4	41.8	31.7	36.8
161	16.0	14.1	14.2	14.5	13.3	12.8	40.9	26.1	29.8	41.5	25.9	31.7

6.2.3 Kinetic tests

Kinetic Tests at different HRT

	24 h		18 h		12 h		10 h		8 h		6 h
Time	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**	**g/L**
hours	average	st.dev.	average	st.dev.	average	st.dev.	average	st.dev.	average	st.dev.	average
0	5.2	0.2	5.7	0.0	9.3	0.6	8.8	0.1	7.4	0.7	6.3
1	5.2	0.2	5.7	0.1	8.9	0.6	9.6	0.0	7.3	0.7	6.4
2	5.5	0.2	5.6	0.2	8.8	0.5	8.6	0.2	7.5	1.2	6.9
3	5.1	0.3	5.7	0.1	8.6	0.7	8.4	0.2	7.2	0.4	6.3
4	5.4	0.1	5.6	0.1	8.5	0.8	8.4	0.4	7.0	0.8	6.5
24	5.2	0.3	5.1	0.3	7.9	0.9	6.6	0.6	5.8	1.2	6.3

Biomass removal contribution			
	Suspended	Attached	Total
HRT	g/d	g/d	g/d
24	0.1	11.7	13.1
18	0.6	14.5	21.6
12	1.0	24.7	33.6
10	2.4	40.0	44.6
8	1.7	46.7	74.7
6	0.2	-	45.2

6.2.4 Total biogas and methane production, and biogas composition

6.2.4.1 Total biogas and methane production

	Total Gas Production as N_2		Total Biogas	Methane
Day	ml/min	L/d	L/d	L/d
12	5.9	8.4	6.9	4.5
14	5.4	7.8	6.3	4.1
15	5.6	8.0	6.5	4.3
16	5.5	7.9	6.4	4.2
17	5.3	7.6	6.2	4.0
18	5.1	7.4	6.0	3.9
19	4.9	7.0	5.7	3.7
20	4.6	6.7	5.4	3.6
21	4.7	6.8	5.5	3.6
22	4.4	6.3	5.1	3.4
23	4.3	6.2	5.1	3.3
24	4.1	5.9	4.8	3.1
26	5.0	7.2	5.9	3.9
27	4.9	7.1	5.7	3.8
28	5.7	8.1	6.6	4.3
29	4.9	7.0	5.7	3.7
30	5.0	7.2	5.9	3.8
31	6.3	9.1	7.4	4.9
32	6.2	9.0	7.3	4.8
33	6.3	9.1	7.4	4.9
34	5.9	8.5	6.9	4.6
35	6.0	8.6	7.0	4.6
36	5.9	8.5	6.9	4.6
37	5.8	8.4	6.8	4.5
38	6.2	8.9	7.2	4.8
40	7.2	10.4	8.4	5.6
41	7.6	10.9	8.9	5.8
42	7.1	10.2	8.3	5.5
43	7.1	10.2	8.3	5.5
44	7.0	10.0	8.2	5.4
45	6.9	9.9	8.1	5.3
50	5.8	8.4	6.8	4.5
51	6.5	9.4	7.6	5.0
52	7.1	10.2	8.3	5.5
54	8.0	11.5	9.3	6.2
56	7.9	11.4	9.3	6.3
57	10.6	15.3	12.4	8.5

	Total Gas Production as N_2		Total Biogas	Methane
Day	ml/min	L/d	L/d	L/d
58	12.0	1.0	17.3	1.5
59	10.5	0.5	15.1	0.7
60	10.4	0.5	15.0	0.7
61	11.2	0.5	16.1	0.7
62	11.3	0.4	16.3	0.6
63	10.3	0.7	14.8	0.9
64	10.0	0.4	14.4	0.6
65	10.5	0.3	15.1	0.5
66	9.9	0.4	14.2	0.6
71	12.3	1.4	17.7	2.0
72	11.9	0.6	17.2	0.9
73	11.3	0.9	16.3	1.3
74	11.1	0.5	16.0	0.7
75	11.3	0.5	16.2	0.8
76	11.2	0.8	16.1	1.1
78	13.3	0.8	19.2	1.1
79	12.5	1.3	18.0	1.9
81	11.9	0.6	17.1	0.8
82	11.0	1.0	15.8	1.4
83	12.8	1.3	18.4	1.8
87	13.3	0.6	19.2	0.9
88	12.7	0.6	18.3	0.8
89	12.2	1.1	17.6	1.6
92	15.9	0.4	22.9	0.6
93	15.0	0.7	21.7	0.9
94	13.9	0.7	20.1	0.9
95	13.5	0.6	19.4	0.8
96	14.3	0.8	20.7	1.2
100	14.8	1.8	21.3	2.5
101	14.7	1.8	21.2	2.6
102	14.4	2.0	20.8	2.8
103	13.9	0.6	20.1	0.9
106	13.2	0.9	19.0	1.3
107	13.1	0.6	18.9	0.8
108	13.0	0.6	18.7	0.8
109	14.4	0.9	20.7	1.3
110	16.1	0.7	23.3	1.0

	Total Gas Production as N_2		Total Biogas	Methane
Day	ml/min	L/d	L/d	L/d
113	15.4	0.8	22.2	1.2
114	15.2	0.7	21.8	1.0
115	18.2	0.7	26.2	1.1
116	19.1	0.7	27.5	1.0
117	19.1	0.8	27.5	1.1
120	17.3	0.9	25.0	1.3
121	15.1	0.8	21.8	1.2
122	18.4	0.9	26.5	1.3
123	20.4	0.8	29.4	1.2
124	20.8	0.7	30.0	1.1
127	18.0	0.9	26.0	1.3
129	20.9	1.0	30.1	1.4
130	18.3	1.0	26.3	1.4
131	20.1	0.8	29.0	1.2
134	19.1	0.7	27.6	1.0
135	18.2	0.7	26.2	0.9
136	18.5	0.6	26.6	0.9
137	18.6	0.6	26.7	0.9
138	18.5	0.7	26.6	1.0
141	20.0	0.8	28.7	1.2
143	19.8	0.8	28.5	1.1
144	20.0	1.3	28.8	1.9
147	15.4	2.2	22.1	3.2
151	18.0	1.3	25.9	1.9
152	19.5	0.5	28.1	0.8
154	18.9	0.7	27.2	1.0
155	17.9	0.8	25.8	1.1
156	19.6	0.8	28.2	1.1
157	19.8	0.7	28.5	0.9
158	16.8	0.7	24.2	1.0
160	19.8	4.8	28.6	6.9
161	19.1	4.7	27.4	6.7

6.2.4.2 Biogas composition

Gas Composition					
HRT	**Day**	**Methane (%)**	**St.Dev**	**CO_2 (%)**	**St.Dev**
24	11	61.9	1.4	25.2	0.7
	18	63.1	2.2	21.8	0.7
	25	70.2	2.7	21.1	1.4
18	39	64.1	2.2	21.1	0.8
	46	68.1	3.4	22.0	1.8
	53	64.1	2.3	23.1	0.6
12	60	66.9	2.5	24.3	1.0
	67	70.1	3.0	22.1	0.9
	74	68.3	1.4	25.1	1.1
10	92	68.8	3.5	27.3	1.5
	95	70.8	2.2	22.7	0.6
	102	66.3	2.6	21.7	0.8
8	109	65.7	1.9	29.9	1.4
	116	65.8	3.5	30.0	0.9
	126	65.6	4.2	25.9	2.1
	130	66.8	3.8	27.2	1.3
6	137	63.5	2.0	32.5	0.7
	144	65.4	0.7	30.0	0.5
	154	58.0	1.2	30.6	2.7
	158	61.6	1.3	34.1	1.4

6.2.5 COD conversion into methane

	TCOD into methane	sCOD into methane
Day	LCH_4/g TCOD	LCH_4/g sCOD
15	0.315	0.319
17	0.291	0.304
20	0.313	0.289
22	0.295	0.285
24	0.294	0.298
27	0.344	0.327
29	0.341	0.325
31	0.466	0.316
34	0.352	0.304
36	0.352	0.319
38	0.384	0.301
41	0.153	0.374
43	0.143	0.345
45	0.330	0.325
50	0.333	0.260
52	0.255	0.426
57	0.443	0.471
59	0.396	0.403
62	0.336	0.429
64	0.297	0.301
66	0.325	0.299
71	0.260	0.315
73	0.328	0.329
76	0.362	0.302
78	0.199	0.371
83	0.253	0.245
87	0.257	0.245
92	0.212	0.246
94	0.249	0.227
101	0.372	0.223
106	0.414	0.216
108	0.246	0.183
113	0.258	0.191
115	0.250	0.237
120	0.234	0.247
122	0.150	0.222
127	0.235	0.337
129	0.320	0.338

	TCOD into methane	sCOD into methane
Day	LCH_4/g TCOD	LCH_4/g sCOD
134	0.371	0.276
136	0.486	0.283
138	0.372	-
141	0.393	0.272
143	0.271	0.283
155	0.378	0.235
157	0.304	0.270
160	-	0.294

6.3 Engineering significance

6.3.1 OLR and SALR and sCOD removal at different HRT

HRT	OLR	sCOD	SALR
[h]	[kg sCOD/m^3 d]	[%]	[g/m^2 d]
24	3.5	91	12.8
	3.8	91	13.8
	3.6	93	13.4
	3.3	94	12.3
	3.3	94	12.2
	3.1	94	11.4
	3.0	91	10.9
	2.6	94	9.6
	2.9	94	10.6
	2.9	95	10.5
18	3.8	94	14.1
	4.0	93	14.6
	3.9	92	14.4
	4.0	94	14.6
	4.1	93	15.1
	3.8	92	14.1
	4.3	94	15.9
	3.2	90	11.8
12	4.0	89	14.6
	5.3	89	19.5
	6.7	88	24.6
	6.6	89	24.3
	7.8	88	28.7
	6.9	87	25.4
	7.4	84	27.2
	7.2	84	26.5
	10.7	89	39.3
10	10.5	88	38.4
	13.0	87	47.8
	12.3	87	45.1
	13.4	86	49.1
	9.1	82	33.4
	13.2	86	48.5

HRT	OLR	sCOD	SALR
[h]	[kg sCOD/m^3 d]	[%]	[g/m^2 d]
8	12.2	86	45.0
	15.4	83	56.6
	14.4	82	52.8
	14.1	88	51.7
	16.6	77	61.1
	16.0	82	59.0
6	12.4	72	45.6
	13.6	67	49.8
	14.7	65	54.0
	14.0	64	51.5
	14.3	64	52.5
	13.4	48	49.2
	14.6	59	53.7
	14.5	62	53.4
	15.3	55	56.1
	14.7	62	54.0
	13.5	60	49.7

6.3.2 sCOD loadings, reactor and media volume requirement at different fillings and percentage removal

6.3.2.1 80% Removal

Loadings & required media volumes					
>80%	Wastewater per beer ratio	Mass sCOD	Media Volume Required (m^3)		
	L/L	kg sCOD/d	AC450	AC515	AC920
Small	2 : 1	30	1.3	1.1	0.8
	4 : 1	40	1.8	1.5	1.1
	10 : 1	49	2.2	1.8	1.3
Medium	2 : 1	1507	67	56	40
	4 : 1	2000	89	74	53
	10 : 1	2466	110	91	65
Large	2 : 1	4521	202	168	120
	4 : 1	6000	268	223	159
	10 : 1	7397	331	274	196

		Reactor Volume per media filling								
		AC450 (402)			AC515 (485)			AC920 (680)		
		30%	50%	70%	30%	50%	70%	30%	50%	70%
Small	2 : 1	4.5	2.7	1.9	3.7	2.2	1.6	2.7	1.6	1.1
	4 : 1	6.0	3.6	2.6	4.9	3.0	2.1	3.5	2.1	1.5
	10 : 1	7.4	4.4	3.2	6.1	3.7	2.6	4.3	2.6	1.9
Medium	2 : 1	225	135	96	186	112	80	133	80	57
	4 : 1	298	179	128	247	148	106	176	106	76
	10 : 1	368	221	158	305	183	131	217	130	93
Large	2 : 1	674	405	289	559	335	239	399	239	171
	4 : 1	895	537	383	742	445	318	529	317	227
	10 : 1	1103	662	473	914	549	392	652	391	280

6.3.2.2 90% Removal

Loadings & required media volumes					
>90%	Wastewater per beer ratio	Mass sCOD	Media Volume Required (m^3)		
	L/L	kg sCOD/d	AC450	AC515	AC920
Small	2 : 1	30	3.1	2.6	1.8
	4 : 1	40	4.1	3.4	2.4
	10 : 1	49	5.1	4.2	3.0
Medium	2 : 1	1507	155	128	92
	4 : 1	2000	206	170	122
	10 : 1	2466	253	210	150
Large	2 : 1	4521	465	385	275
	4 : 1	6000	617	511	365
	10 : 1	7397	760	630	450

		Reactor Volume per media filling								
		AC450 (402)			AC515 (485)			AC920 (680)		
		30%	50%	70%	30%	50%	70%	30%	50%	70%
Small	2 : 1	10.3	6.2	4.4	8.6	5.1	3.7	6.1	3.7	2.6
	4 : 1	13.7	8.2	5.9	11.4	6.8	4.9	8.1	4.9	3.5
	10 : 1	16.9	10.1	7.2	14.0	8.4	6.0	10.0	6.0	4.3
Medium	2 : 1	516	310	221	428	257	183	305	183	131
	4 : 1	685	411	294	568	341	243	405	243	174
	10 : 1	845	507	362	700	420	300	499	300	214
Large	2 : 1	1549	929	664	1284	770	550	916	549	392
	4 : 1	2056	1234	881	1704	1022	730	1215	729	521
	10 : 1	2535	1521	1086	2101	1261	900	1498	899	642

6.3.3 Fort Garry design parameters using AC920

AC920		
Value	**Unit**	**Parameter**
46	$g/(m^2\ d)$	SALR
85	%	Removal achieved at SALR
19565	m^2	SA required
680	m^2/m^3	SPSA
29	m^3 media	Media required
70	%	Filling fraction
41	m^3 tank	Reactor volume
0.9	kg/m^3	Stage 1 effluent
46	$g/(m^2\ d)$	SALR
85	%	removal achieved at SALR
2935	m^2	SA required
680	m^2/m^3	SPSA
4	m^3 media	Media required
70	%	Filling fraction
6	m^3 tank	Reactor volume
0.135	kg/m^3	Final effluent
47	**m^3**	**Total volume**
101.25	**mg BOD_5/L**	**BOD_5**