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Anaerobic Treatment Of Industrial Wastewater With a Pilot Scale UASB Reactor

MASTER'S THESIS

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Environmental Engineering

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Abstract

The focus of this study was to investigate the performance of anaerobic treatment with a UASB pilot reactor for removal of COD in industrial wastewater at ambient temperatures at Grødaland wastewater treatment plant in Rogaland, Norway. The background for the study was the high-energy cost for aeration of industrial wastewater with aerobic treatment in SBR, so pre-treatment with a UASB reactor would reduce the organic load on the SBR and thus save energy and produce biogas. The wastewater had about 1200 mg COD/l and the temperature was between 15℃ to 23 ℃ during the experimental period. The hydraulic retention time (HRT) was between 26 h and 18 h during the four months of testing. The organic loading rate (OLR) varied in the range of 0.5 to 2 g COD/l. d.

The results demonstrate a significant COD removal efficiency of above 50 % on average during the experimental period. In general, by increasing OLR and decreasing HRT, COD removal efficiency increased as well under steady state condition. This indicates the loading was below the maximum capacity and that the increased flowrate may have improved the hydraulic conditions in the reactor. However, the COD removal also increased at constant OLR probably due to the poor characteristics of first granules which were not in the steady state condition in the reactor during two first months of study. At HRT (18 h) with 0.8 g COD/l. d, and 23 ℃, COD removal efficiency reached the maximum of above 70% with second types of granules which had more regular shaped granules. Furthermore, the color of effluent and TSS removal clearly reveal that the granular sludge in UASB had good settling properties and that the sludge was retained efficiently.

An anaerobic treatment systems using UASB reactor for treating industrial wastewater represents an applicable and feasible alternative as pre-treatment for SBR units at IVAR Grødaland by reducing the COD load on the SBR, saving energy for aeration and converting the organic load into economically a valuable product as methane.

Keywords: *anaerobic treatment, industrial wastewater, UASB reactor, COD removal efficiency, methane production, TSS removal*

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Abbreviations

ADM1	Anaerobic Digestion Model No.1
AMB	Acetoclastic Methanogenic Bacteria
COD	Chemical Oxygen Demand
DAF	Dissolved Air Flotation
DO	Dissolved Oxygen
F/M	Food Mass Ratio
HAc	Acetic Acid
HMB	Hydrogenotrophic Methanogenic Bacteria
HRT	Hydraulic Retention Time
IVAR	Interkommunalt Vann Avløp og Renovasjon
OHPB	Obligate Hydrogen Producing Bacteria
OLR	Organic Loading Rate
SBR	Sequencing Batch Reactor
SRT	Solid/Sludge Retention Time
STP	Standard Temperature and Pressure
UASB	Up-flow Anaerobic Sludge Blanket
VFA	Volatile Fatty Acid
VSS	Volatile Suspended Solid
WWTP	Wastewater Treatment Plant

Chapter 1

Introduction

Nowadays, one of the great challenges in the water field is treatment of wastewater for the protection of public health and environment, and conservation of resources. Wastewater is defined as a combination of domestic and industrial wastewater, and surface water. The function of wastewater treatment is to accelerate the natural processes for purifying the water. Several technologies are applied today in the wastewater treatment plant, including aerobic treatment, anaerobic treatment, and a combination of anaerobic and aerobic treatment [5,7].

Anaerobic processes generate energy in terms of biogas, and produce lower sludge compared to aerobic processes. In contrast, aerobic processes require energy for aeration, and produce excessive sludge. Therefore, operation and maintenance cost of aerobic system including high sludge production led to the development of economically more high-rate anaerobic reactors. [50]. Consequently, anaerobic treatment combined with other proper methods is an example of advanced sustainable technology society needs [39]. Making transition to sustainable energy could achieve many privileges for all over the world, such as reduced air pollution, fuel supply diversity, and abatement global warming are substantial benefits of green energy [18]. The demand for energy sources has been grown due to a rise in the cost of available fuels, hence anaerobic treatment is also profitable from an energy point of view [7]. Anaerobic treatments have been developed and utilized in the recent decades with considerable attention towards the high rate anaerobic reactors [8, 12].

High rate anaerobic reactors are used for high strength industrial wastewater which is characterized by the COD concentration higher than 1000 mg/l. However, the treatment efficiencies of these reactors are sensitive to parameters such as wastewater compositions, pH, and temperature [12, 8]. All high rate reactors are based on the concept of retention of active biomass by bacterial sludge immobilization. Upflow anaerobic sludge blanket (UASB) reactors are the most successfully types of high-rate anaerobic systems for the treatment of wastewaters [8]. UASB reactors are operated frequently for pre-treatment of industrial wastewaters, and they are also feasible for treatment of municipal wastewater with low temperature [10]. The great success of UASB process is based on the capability to retain active sludge in the reactor due to granule formation, facilitating the economical wastewater treatment plants [15].

Anaerobic treatment processes have been commonly performed under mesophilic condition at optimum methanogenic growth rate of 35 – 37 ℃, however different types of wastewater fractions might be warmer or cooler. Moreover, many researchers have investigated the anaerobic processes performance under various temperatures, due to the advantages of this system for capable of treating high organic loading [7,37].

1.1 Scope of work

This master thesis was a part of project in IVAR, by taking consideration into the COD removal efficiency as well as production of biogas. IVAR *(Interkommunalt Vann Avløp og Renovasjon)* is a Norwegian public company that constructs and operates municipal facilities for water, wastewater and solid waste. In this study, UASB pilot scale was tested for treating wastewater effluent from dissolved air flotation unit (DAF) at Grødaland wastewater treatment plant (WWTP). According to the reports from IVAR, the plant receives wastewater from several sources, mainly industrial which are presented in Table 1.1.

The wastewater treatment process consists of a pre-treatment with screens, sand, and fat removal followed by a combined chemical-biological treatment. Figure 1.2 shows the block flow diagram (BFD) of the IVAR Grødaland plant.

Table 1.1 *Grødaland WWTP wastewater sources [1]*

Wastewater Sources	Average Loading rate (m^3/d)
Animal destruction in Norsk Protein	167
Municipal wastewater of 3000 houses in Varhaug and food industry	1680
Dairy and chicken slaughtering in Kviamarka	3284
Total Loading	**5131**

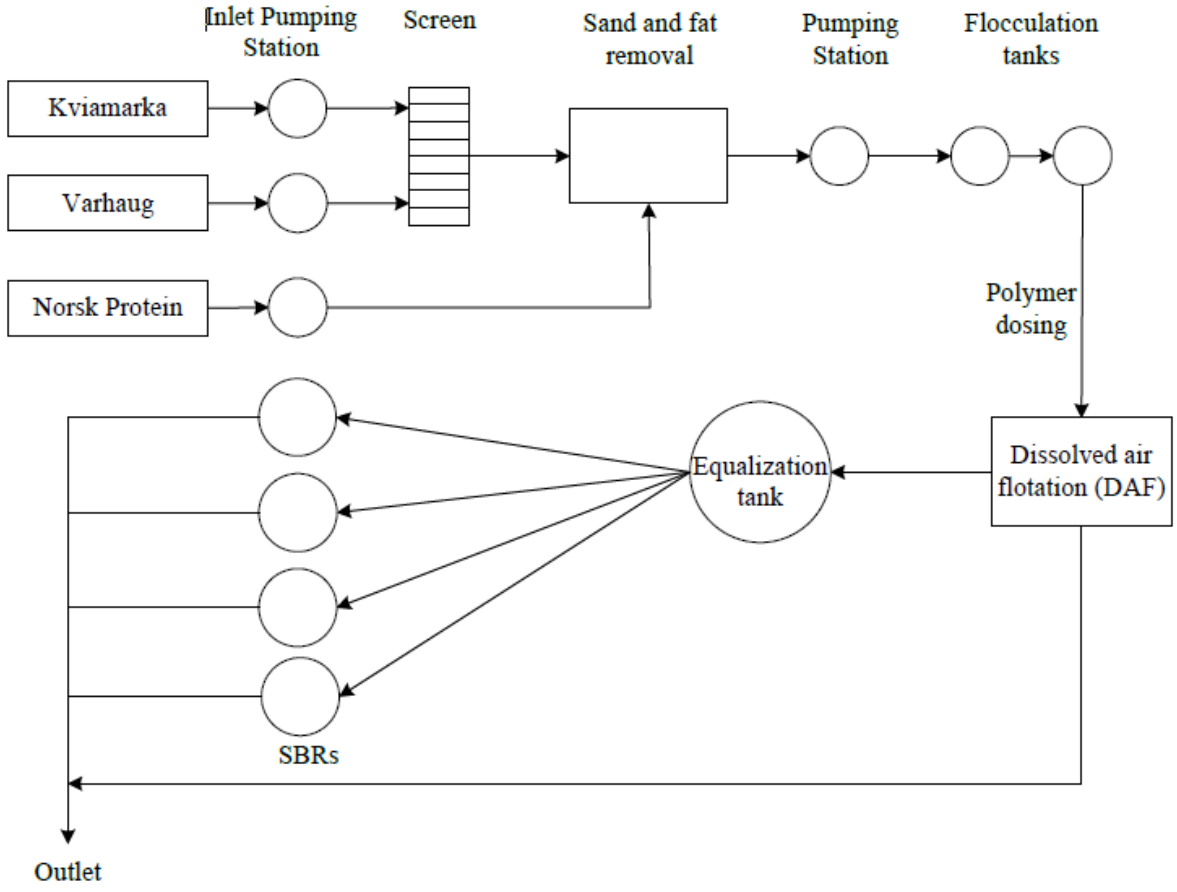


Figure 1.1 *Block Flow Diagram in IVAR Grødaland [1]*

The pre-treatments at Grødaland plant includes bar screen (3 mm opening), sand and fat removal, flocculation tanks, and dissolved air flotation. $FeCl_3$ was previously used as a coagulant for flocculation stage in the startup of the plant. Due to large pH variations in the influent wastewater, $FeCl_3$ was not a suitable coagulant. In addition, $FeCl_3$ removed all phosphorus from the wastewater, so the biological process in the SBR downstream did not work due to lack of phosphorus. Two flotation tanks units with a surface area of 48 m^2 each were designed for 7.5 m/h surface loading with maximum load of 200 l/s. Each DAF unit comprises six pressure pumps, and the number of pumps in operation depends on flow rate and temperature. In the DAF units, 30% of the effluent wastewater is recycled as dispersed water with 4 – 6 bar of back-pressure. The DAF unit removes approximately 60 % of suspended solids with polymer dosing around 2 ml/l. This polymer is not affected by pH variations and it does not remove phosphate, so then biological process downstream also works well. The SS removal is estimable from table 1.2 that presents the average flow rate and wastewater composition of inlet and outlet for the DAF unit in 2016. All pre-treatment processes are followed to reduce the concentration of suspended solids in addition to fat, and grease. Hence, several tests have done by focusing on removal efficiency of suspended solids. The results of performed tests indicated that the addition of more polymer dosing can increase the suspended solids removal to 80%.

On the contrary, the DAF units do not perform so well in COD removal. According to the table 1.2, this stage can remove approximately 20% of dissolved COD and 30% of total COD. High concentration of dissolved COD affects the performance of the biological treatment stage, sequencing batch reactor (SBR) unit. There are four batch reactors that performs both as bioreactor and settling tanks with 750 m^3 volume each. At high COD loading, the SBR cannot perform effectively because of reduction in dissolved oxygen for microorganisms, and poor sludge sedimentation. Thus, the organic loading is controlled by oxygen supply capacity.

Table 1.2 *Dissolved air flotation data [1]*

	Inlet Flotation unit	**Outlet Flotation unit**
SS (mg/L)	368	160
Dissolved COD (mg/L)	787	647
Total COD (mg/L)	1408	912
BOD (mg/L)	1059	646
Total Phosphorus (mg/L)	26,1	22,2
Conductivity (µS/cm)	1938	2006
PH	7,73	7,48
Flow (m^3/d)	5453	5453

A solution to resolve this challenge is to remove COD before the SBR units by using an anaerobic high reactor such as UASB as pre-treatment with high solid retention time. The UASB reactor is supposed to retain high concentration of biomass in granules, resulting high COD removal efficiency, and generate methane by converting organic matter to biogas as a valuable product. Based on this, IVAR has defined the project to treat wastewater anaerobically by using the pilot scale UASB reactor for removing COD and thus reducing the organic load on the SBR, with an aim to achieve 50-70% COD removal, based on pervious tests at UIS (University of Stavanger) [6].

1.2 Objectives

The main objective of this master thesis was to investigate the performance of anaerobic treatment for removing the high COD concentration of industrial wastewater. This study was conducted to set-up pilot scale UASB reactor for treatment of high strength wastewater under ambient temperature in such a cold country, Norway.

1.3 Thesis Outline

This master thesis is entitled: **"Anaerobic treatment of industrial wastewater with a pilot scale UASB reactor"** and divided into six chapters.

1.Introduction;

2. Literature Review and Theoretical Background;

3. Materials and Methods;

4. Results;

5. Discussions; and

6. Conclusions;

Appendixes are attached to present supporting materials of the whole study.

Chapter 2

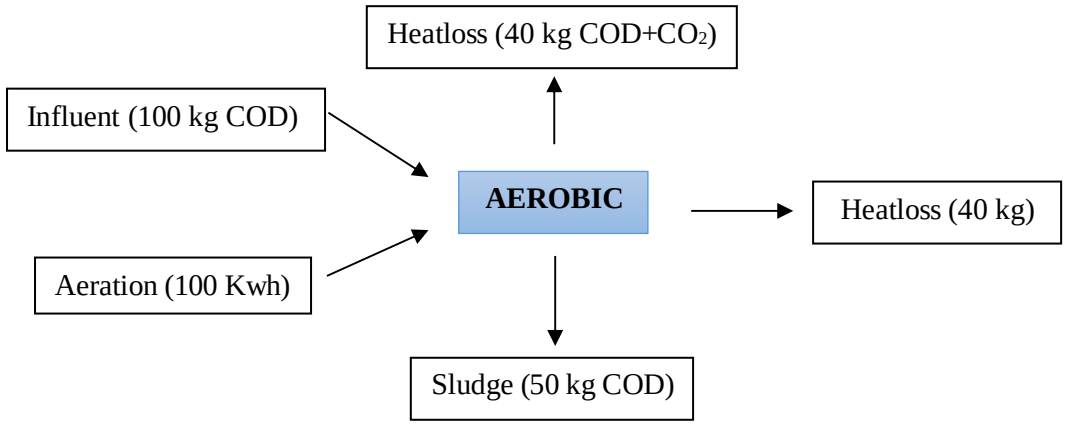
Literature Review and Theoretical Background

This chapter describes the theoretical explanations of anaerobic processes as well as defines anaerobic stoichiometry. Explanations follow a brief review of the factors that are affected the anaerobic treatment processes of UASB reactor, descriptions about UASB reactor and the application UASB reactor in the wastewater industry.

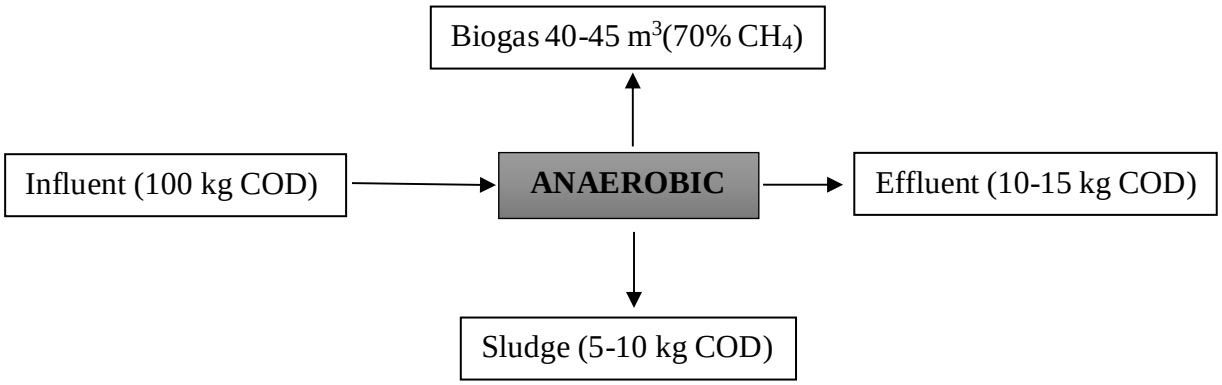
2.1 Anaerobic Treatment

2.1.1 Introduction

Anaerobic treatment is an effective method for the treatment of waste sludge and high strength organic wastes. This process is mediated by anaerobic microorganisms, converting the biodegradable material into products in the absence of oxygen [32]. Anaerobic process is advantageous mainly because of less biomass produced per unit of substrate, even anaerobic sludge which is produced in the bioreactor, is valuable [12, 32]. Moreover, energy in the form of biogas, are generated from the biological conversion of organic materials. For treating high-strength industrial wastewaters, anaerobic process is a suitable method because of savings in energy, nutrient addition, and reactor volume compared to aerobic process [25]. Anaerobic treatment is commonly used as a pretreatment step which is followed by an aerobic process to achieve good quality for effluent, mostly for removing the organic pollutants, and reducing the chemical oxygen demand (COD) level [33]. Figure 2.1 presents the carbon and energy flow in both aerobic and anaerobic wastewater treatment. In figure 2.1, The oxidation of 1 kg COD is assumed to require 1 kWh of aeration energy [12].



(a)



(b)

Figure 2.1 *The fate of carbon and energy in aerobic and anaerobic wastewater treatment, (a) Aerobic process, (b) Anaerobic process [12]*

Large fraction of influent COD in aerobic processes are converted to sludge usually about 50% as well as loosing heat, and head production. In anaerobic processes a large fraction of influent COD is converted to biogas, and the biomass yield [12].

2.1.2 Process Description

Anaerobic treatment is a series of biological processes where organic material is degraded by microorganisms without oxygen (low redox potential). The microbial consortia convert organic matter into mainly methane (CH_4), carbon dioxide (CO2), ammonium (NH_3), hydrogen sulphide (H_2S), and water (H_2O) [12]. The multi-step nature of anaerobic digestion follows four phases: hydrolysis, acidogenesis (also known as fermentation), acetogenesis and

methanogenesis, as presented in figure 2.2 [2]. In the acetogenesis step, the VFAs generated in acidogenesis [25]. The four basic processes will be described in the subsequent section.

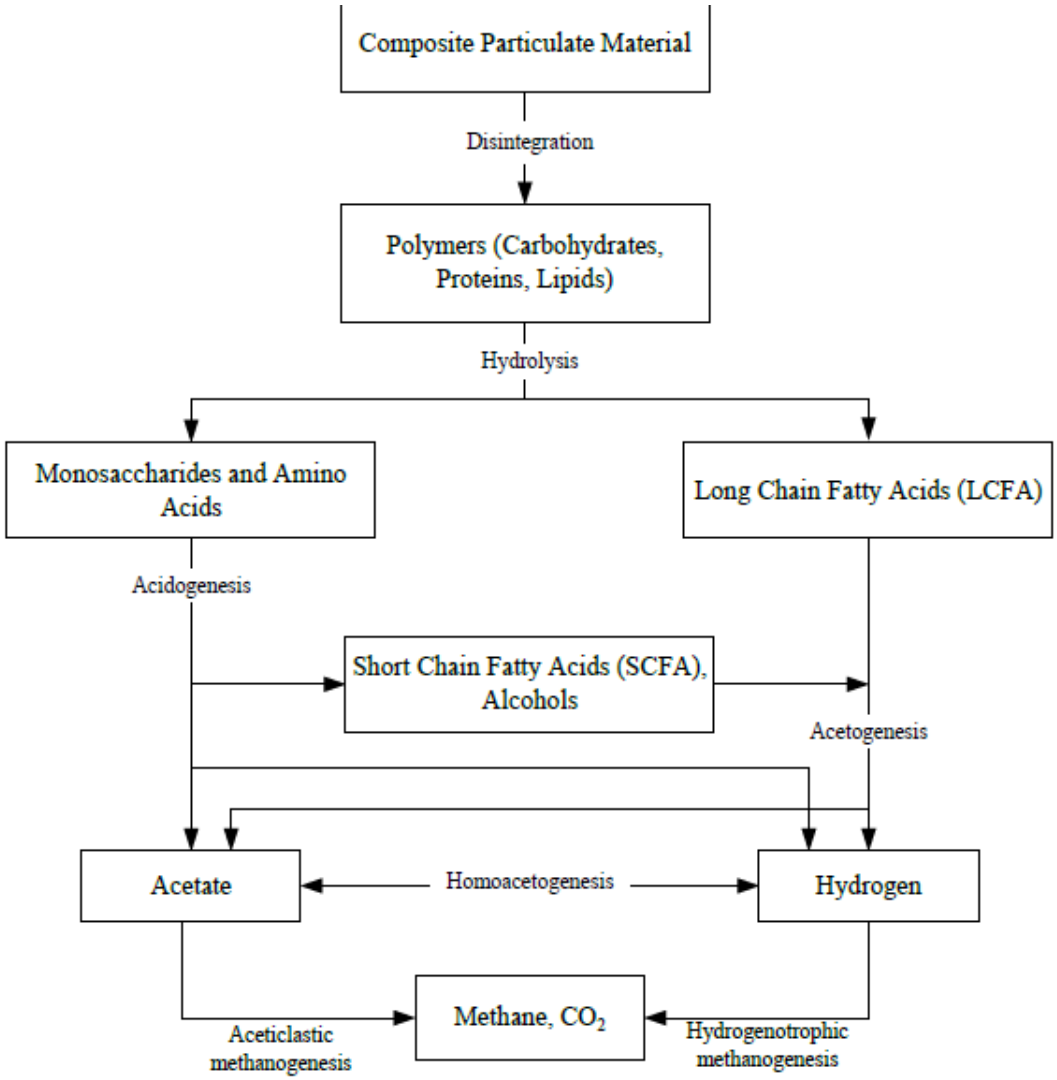


Figure 2.2 *Multistep of anaerobic processes [3,12,23]*

2.1.2.1 Hydrolysis

The first step of anaerobic degradation, in which particulate material is reduced in size to smaller products that can then simplify transport across the cell barrier, is termed hydrolysis. During the hydrolysis step, complex organic material such as carbohydrates, proteins and lipids are broken down to soluble compounds such as monosaccharide ($C_{12}H_{22}O_{11}$), amino acids (R-

CH (NH2)-COOH), and volatile fatty acids (VFAs), respectively. The process is carried out by extracellular enzymes in the vicinity of the particle which are produced by microorganism and gains from the soluble products [12]. The products of hydrolysis serve as substrates for the next step, acidogenesis. Solubilization and size reduction are catalyzed by extracellular enzymes that are produced by the facultative bacteria that carries out acidogenesis [12,24]. The main enzyme groups are proteases (acting on proteins), cellulases, amylases, glucanases (acting on polysaccharides), and lipases (acting on fats and oil; lipids) [2,24]. Hydrolysis is regarded as rate- limiting step for the digestion process, with substrates and wastewater with a high suspended solid content. Furthermore, the rate of hydrolysis is a function of temperature and temperature fluctuations [12,26].

2.1.2.2 Acidogenesis

Acidogenesis is the anaerobic oxidation of the hydrolysis products to simpler compounds, mainly volatile fatty acids (VFAs). The acidogenesis products include a variety of small organic materials i.e. acetate and higher organic acids such as propionate, and butyrate as well as, ethanol, lactic acid, ammonia, and CO_2 [12]. The composition of end products depends on various parameters such as substrate composition, environmental factors (pH, temperature, etc.) and operational factors in the reactor medium [12,2]. Table 2.1 lists some acidogenic reactions with sucrose from thermodynamically aspect. According to the table, the ΔG (free energy) of the reactions depends on the dissolved H_2 concentration. If H_2 accumulates, more reduced products such as propionate will appear, and more alcohols and lactate. However, if more H_2 is taken up by other organisms such as methanogens, acetate will be main product [12]. Acetic acid and H_2 are used directly for Methanogenesis as substrates and energy source, whereas the other fermentation products must be converted to acetic acid, and H_2, in acetogenesis process, in order to be utilized in methanogenesis [12,15].

Table2.1 *acidogenic reactions with sucrose at 25 °C [12]*

Reactions	ΔG°' (kJ/mol)
$C_{12}H_{22}O_{11} + 9H_2O \rightarrow 4CH_3COO^- + 4HCO_3^- + 8H^+ + 8H_2$	- 457.5
$C_{12}H_{22}O_{11} + 5H_2O \rightarrow 2CH_3CH_2CH_2COO^- + 4HCO_3^- + 6H^+ + 4H_2$	- 554.1
$C_{12}H_{22}O_{11} + 3H_2O \rightarrow 2CH_3COO^- + 2CH_3CH_2COO^- + 2HCO_3^- + 6H^+ + 2H_2$	- 610.5

The fermentation reaction is rapid and is a common microbial conversion. The growth rate of acidogenic bacteria is comparable to aerobic rates with μm of 2 - 7 d^{-1} [3]. In the fermentation

phase, organic compounds serve as both electron donors and acceptors. Free energy of acidogenic reactions are the highest of all anaerobic operations. The large fraction of energy which is obtained from the oxidation of H_2, causes to limit the biomass growth yield, thus the growth yield is low for fermentation step in about Y~ 0.1 - 0.2 gVSS/g COD [12,25,28]. However, the growth rate and conversion rate are higher compared to methanogens (Table 2.2) [12].

Table2.2 *Average kinetic properties of acidifies and methanogens [12]*

Process	**Conversion rate** (gCOD/g VSS.d)	**Y** (gVSS/ gCOD)	**Ks** (mg COD/l)	**µm** (1/d)
Acidogenesis	13	0.15	200	2.0
Methanogenesis	3	0.03	30	0.12
Overall	2	0.03-0.18	--	0.12

2.1.2.3 Acetogenesis

Acetogenesis refer to further fermentation phase by bacteria to convert acidogenesis products to acetate, hydrogen and carbon dioxide. However, the methanogens and the acidogens can also form syntrophic relationships, which provide cultures to change formate, hydrogen, and acetate to methane and carbon dioxide directly [12, 25]. Propionate and butyrate are the most significant acetogenic substrates in the anaerobic operation process. The acetogenic bacteria are obligate hydrogen producers (OHPAs), and their metabolism is dependent on the hydrogen concentration [12]. Studies from thermodynamically aspects in the acetogenesis phase carry out the relationship between ΔG and H_2, the lower H_2 concentration, more negative formation energy, hence the reactions can occur spontaneously. (Table 2.3) Stoichiometric conversion reactions follow propionate conversion as an example:

$$\Delta G = \Delta G + RT \ln * [CO2] * \frac{[Acerate] * [CO2] * [H2]3}{[Propionate]} \quad (2.1)$$

From Table 2.3 follows that the reaction for ethanol, butyrate, propionate, and LCSAs paltimate will not happen, as the ΔG is positive, thus the bacterial energy yield is negative [12].

Table 2.3 *Stoichiometry and change of free energy ΔG for some acetogenic reactions [12]*

Compound	Reaction	ΔG°' (kJ/mol)
Lactate	$CH_3CHOHCOO^- + 2H_2O \rightarrow CH_3COO^- + HCO_3^- + H^+ + 2H_2$	-4.2
Ethanol	$CH_3CH_2OH + H_2O \rightarrow CH_3COO^- + H^+ + 2H_2$	+9.6
Butyrate	$CH_3CH_2CH_2COO^- + 2H_2O \rightarrow 2CH_3COO^- + H^+ + 2H_2$	+48.1
Propionate	$CH_3CH_2COO^- + 3\,H_2O \rightarrow CH_3COO^- + HCO_3^- + H^+ + 3H_2$	+76.1
Methanol	$4\,CH_3OH + 2\,CO_2 \rightarrow 3CH_3COOH + 2H_2O$	-2.9
Hydrogen-CO_2	$2\,HCO_3^- + 4\,H_2 + H^+ \rightarrow CH_3COO^- + 4\,H_2O$	-70.3
Palmitate	$CH_3\text{-}(CH_2)_{14}\text{-}COO^- + 14H_2O \rightarrow 8CH_3COO^- + 7H^+ + 14H_2$	+ 345.6

In acetogenic conversions, there is an association between generation and consumption of hydrogen, that is called interspecies hydrogen transfer. Figure 2.3 is indicated that how the ΔG is related to hydrogen concentration for the for the anaerobic oxidation of propionate, butyrate, and palmitate.

Methanogenic niche is the limitation range between the partial pressures of 10^{-4} to 10^{-6} which is presented in the figure2.3 for the H_2 concentration to set the upper limit for acetogens, and a lower limit for methanogens [12, 2,3].

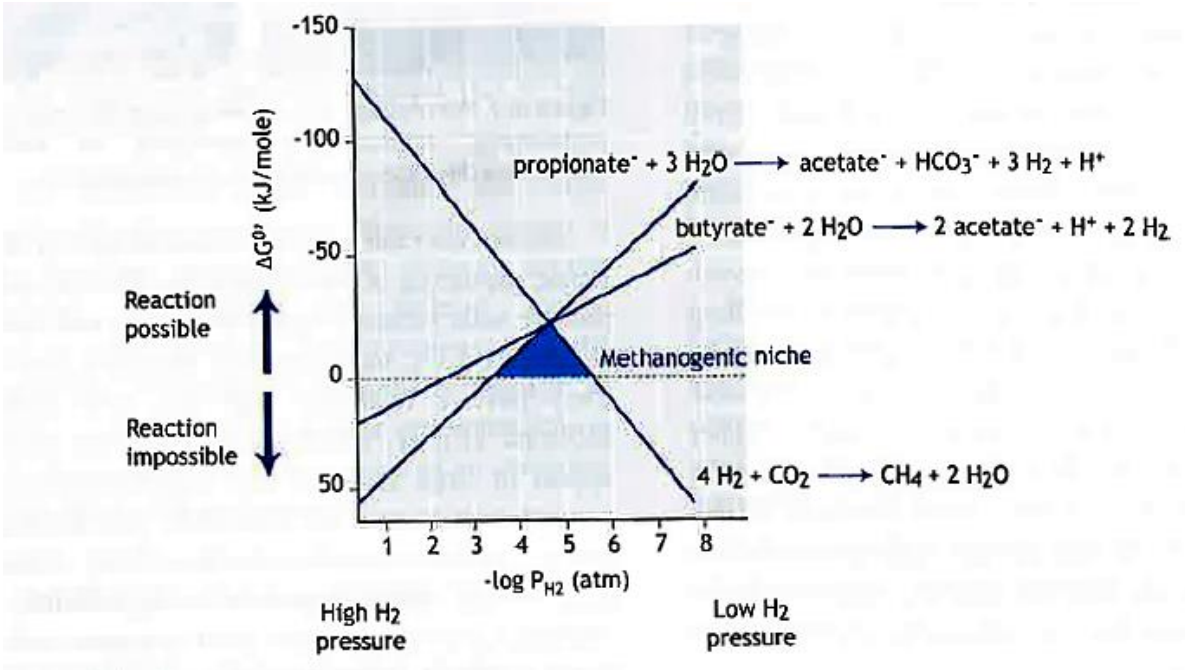


Figure2.3 *Free energy change as a function of H_2 partial pressure [12]*

2.1.2.4 Methanogenesis

Biogas is the ultimate product and the last step of the anaerobic conversion. The products of the acetogenic reactions are used as substrates by methanogens to generate the methane gas. Generation of methane involves two different types of methanogenic bacteria, as depicted in figure 2.4. Acetoclastic methanogenic bacteria (AMB) is the first group that split the acetic acid into methane and CO_2. The other group, hydrogenotrophic methanogenic bacteria (HMB) utilize H_2 as electron donor to reduce carbon dioxide to methane. The growth rate of Acetoclastic methanogens is very low 0.12 d^{-1}**,** resulting in doubling times of several days. The hydrogenotrophic methanogens, that have a much higher growth rate (2.85 d^{-1}). The low growth rate of these bacteria can explain why anaerobic reactors require long start-up time and long retention time. According to the figure 2.2, it is generally accepted that about two-thirds of methane generated is derived from acetic acid. Table 2.6 lists two types of methanogens bacteria with their kinetic characteristics [12].

Table 2.4 *Most important methanogenic reactions with some kinetic properties [12]*

Functional Step	**Reaction**	**ΔG** (kj/mol)	**μ_{max}** (1/d)	**T_d** (d)	**K_s** (mg COD/l)
Acetotrophic methanogenesis	$CH_3^- + COO^- + H_2O \rightarrow CH_4 + HCO_3^-$	-31	0.12	5.8	30
Hydrogenotrophic methanogenesis	$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$	-131	2.85	0.2	0.06

The composition of gas produced in the methanogenesis stage are typically 65 percent methane and 35 percent carbon dioxide. The optimum pH for methanogenesis step is around 6.8-7.6.

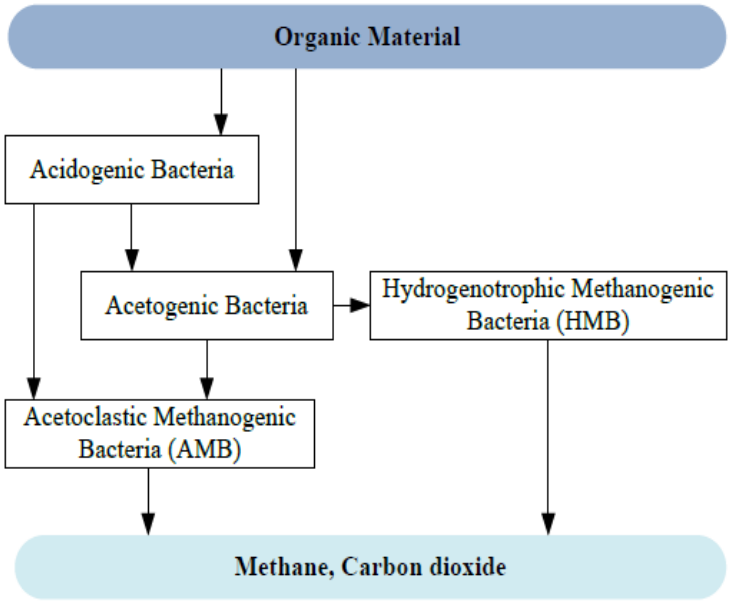


Figure 2.4 *Group of microorganism in anaerobic digestion [3,12,24]*

2.2 Anaerobic Stoichiometry

2.2.1 Chemical Oxygen Demand (COD)

Chemical oxygen demand (COD) is a measure of the electrons available in an organic material and it can be expressed in terms of the amount of oxygen required to oxidize organic compounds completely to CO_2 and water [2]. The number of electrons donated by the oxidant is expressed as oxygen equivalents in mgO_2/l. 1 mole of O_2 weight 32 g and accepts 4 electron equivalents. Hence, 1 electron equivalent (eeq) corresponds to 8 of oxygen or COD [12]. Equation 2.2 shows the relationship between COD and electron equivalents.

$$\frac{1}{2}H_2O \rightarrow H^+ + \frac{1}{4}O_2 + e^- \rightarrow \frac{1}{4}mol\ O_2 \times 32\ \frac{g}{mol} = 8\ gram\ O_2$$

$$(1\ eeq = 8\ gram\ COD) \qquad (2.2)$$

The theoretical COD can be calculated based on the chemical oxidation reaction as shown in equation 2.3. This equation expresses 1 mole of organic material demands *¼(4n+a-2b)* mole of O_2 or *8(4n+1-2b)* gO_2[12].

$$C_nH_aO_b + \frac{1}{4}(4n + a - 2b)\ O_2 \rightarrow n\ CO_2 + \frac{a}{2}H_2O \qquad (2.3)$$

For organic compounds containing nitrogen compounds equation 2.4 apply which includes the numbers of N and weight of N in the compound [12].

$$C_nH_aO_bN_d + \frac{1}{2}(2n + 0.5a - b)\ O_2 \rightarrow n\ CO_2 + d\ NH_3 + \frac{a-3}{2}\ d\ H_2O \qquad (2.4)$$

The theoretical COD per unit mass varies for different organic compounds. In case of methane, one mole of methane requires two mole of oxygen to oxidize it to carbon dioxide and water [2]. The calculation of COD equivalent of methane is shown in equation 2.5. Here the oxidation number for carbon varies from -4 (as found in CH_4) to +4 (as found in CO_2) [12].

$$CH_4 + O_2 \rightarrow CO_2 + H_2O$$

$$\frac{COD}{CH_4} = 2\times\frac{32}{16} = 4\ \frac{g\ COD}{g\ CH_4} \qquad (2.5)$$

The compound ($C_nH_aO_bN_d$) is converted by anaerobic microorganism into CH_4, CO_2 and NH_3, and the theoretical amount of methane gas can be found by Buswell Equation 2.6 [12].

$$C_nH_aO_bN_d + \left(n - \frac{a}{4} - \frac{b}{2} + \frac{3d}{4}\right)H_2O \rightarrow \left(\frac{n}{2} + \frac{a}{8} - \frac{b}{4} - \frac{3d}{8}\right)CH_4 + \left(\frac{n}{2} - \frac{a}{8} + \frac{b}{4}\frac{3d}{4}\right)CO_2 + d\ NH_3 \qquad (2.6)$$

2.2.2 COD Fraction

The total COD in wastewater are classified in four fractions based on the biodegradability and size of the compounds, as it shown in figure 2.5. The COD is divided into particulate or slowly biodegradable COD and soluble or readily biodegradable COD. A rough differentiation is by the TSS analysis, larger and smaller than 1 μm.

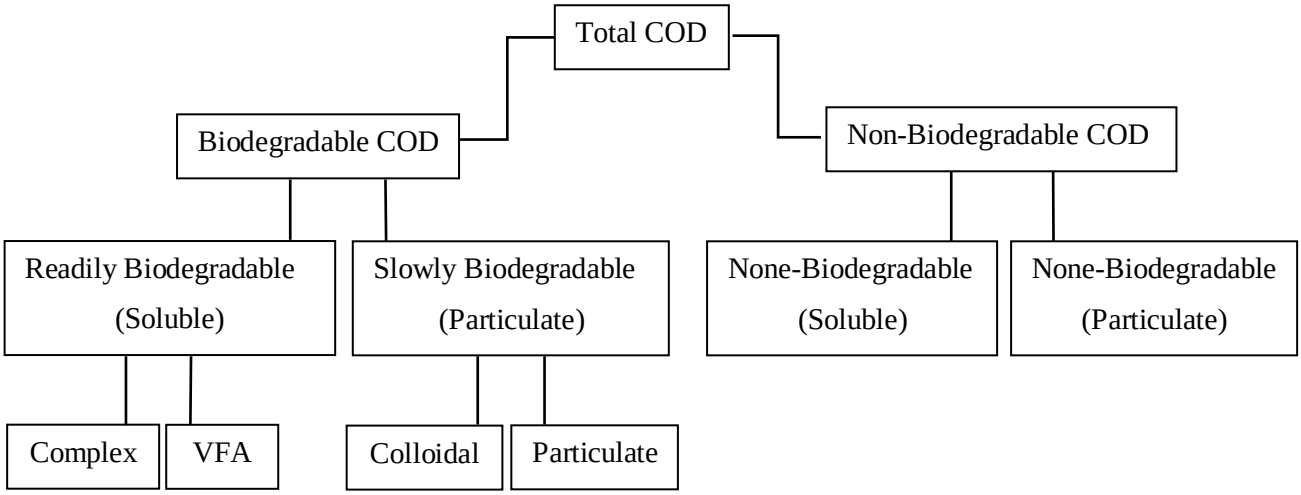


***Figure 2.5**: Fraction of COD and wastewater [25]*

The particulate and soluble fractions are further divided into biodegradable and non-biodegradable fractions. Dissolved biodegradable COD is being directly utilized by the microorganisms and it consists of small molecules and polymers. The particulate COD consists of colloidal and suspended solids and they must be hydrolyzed by extracellular enzymes to soluble compounds before utilization by microorganisms. The non-biodegradable particulate fraction will adsorb to the sludge, because it cannot use by microorganism and will contribute to the total sludge production. The dissolved non-biodegradable COD will be pass through to the effluent, unaffected by any biological or physical process [12,2, 25]. Equation 2.7 presents the calculation of total COD.

$$TCOD = COD_{soluble\ bio} + COD_{particulate\ bio} + COD_{non\ bio\ soluble} + COD_{non\ bio\ particulate} \quad (2.7)$$

2.2.3 COD Balance

Biological system must be monitored by relevant parameters and measurements which be evaluated the development of the processes by them. COD is the control parameter for the anaerobic processes. All COD that enters to the anaerobic system, will end up to the biogas mainly methane, and COD of new bacterial mass. The mass balance for the COD, as control tool, is made by equation 2.8, and is shown by figure 2.6.A COD balance is used to consider the changes of the COD during oxidation [25].

$$COD_{in} = COD_{out}$$

$$COD_{influent} = COD_{effluent} + COD_{gas} + COD_{sludge} \quad (2.8)$$

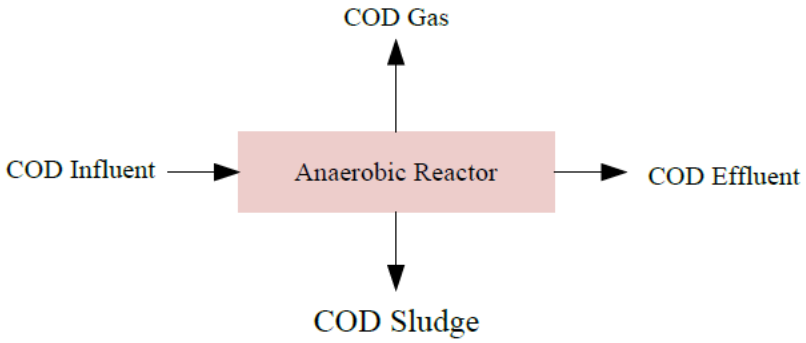


Figure 2.6 *COD balance in the anaerobic process [12]*

2.2.4 Methane Production

The total amount of CH_4 produced in the anaerobic process is determined by organic matter removal in the system. CH_4 is equivalent to a certain amount of COD. The stoichiometry of CH_4 produced is followed equation 2.9, and the production rate of methane is depended on the influent characteristics, i.e. flow rate, COD concentrations, and the biodegradability of the COD [12].

$$CH_4 + O_2 \rightarrow CO_2 + H_2O \quad (2.9)$$

At STP (standard temperature and pressure) of 0 ℃ and one atmosphere,1 kg of COD is converted in 0.35 m^3 of methane (22.41 m^3/64 kgCOD = 0.35m3 CH_4/kg COD), Table 2.5 presents the various methane yields at different temperature, assuming 100% COD conversion [12]. According to the COD balance, the COD effluent and the COD in sludge production should be known to estimate how much COD inlet is converted into biogas. In fact, the COD loss in the anaerobic reactor is accounted as methane production [25]. The gas produced contains carbon dioxide ranges between about 30 and 50%, depending on the nature of the substrate [12].

Table 2.5 *Methane production rate of 100% COD conversion in function of temperature [12]*

Temperature (°C)	**Methane Yield** (l CH_4/ g COD)
0 (273 K)	0.35
20 (293K)	0.37
25 (298K)	0.38
35 (308K)	0.40

2.3 Process Kinetics

2.3.1 Bacterial Growth

Bacterial growth is divided in to four phases:(a) lag phase is the time require for the bacteria to acclimate to the environment before growth; (b) growth phase that is the period of cell division. The biomass growth increase exponentially and the sensitivity of bacteria is high in this period; (c) stationary phase which is the biomass concentration remains constant in function of time; and (d) death phase is when there is not any more growth and the death rate exceeds the growth rate [25]. The growth of bacteria increases exponentially and the reaction for growth is the first order reaction based on the biomass concentration, as shown by equation 2.10.

$$\frac{dX_B}{dt} = \frac{\ln 2}{t_g} X_B \qquad \frac{dX_B}{dt} = \mu \,.\, X_B \tag{2.10}$$

Where dX_B/dt is biomass growth rate; t_g is generation time; μ is specific growth rate (gVSS/gVSS.d); and X_B is biomass (gVSS/l). Monod equation is used as mathematical formula for growth (equation 2.10), and Monod kinetics is illustrated in figure 2.7, where K_S is half saturation constant which is defined as the substrate concentration where μ is half of $\mu_{max;}$ Cs is growth limiting substrate concentration (g/l); μ_{max} is maximum specific growth rate (gVSS/gVSS.d).

$$\mu = \frac{\mu_{max}.C_s}{K_s + C_s} \qquad \frac{dX_B}{dt} = \mu . X_B = \frac{\mu_{max}.C_s}{K_s + C_s} . X_B \tag{2.11}$$

The ratio of amount of biomass produced to the amount of substrate consumption is related to each other by biomass yield (Y), as written in equation 2.12.

$$\frac{dX}{dt} = Y\frac{dC}{dt} \qquad \frac{dC}{dt} = \frac{\mu . X_B}{Y} \qquad \frac{dC}{dt} = k_m.X_B \tag{2.12}$$

Where dc /dt is substrate consumption rate (gCOD/l. d) and μ/Y (k_m) is specific substrate consumption rate (gCOD/gVSS.d).

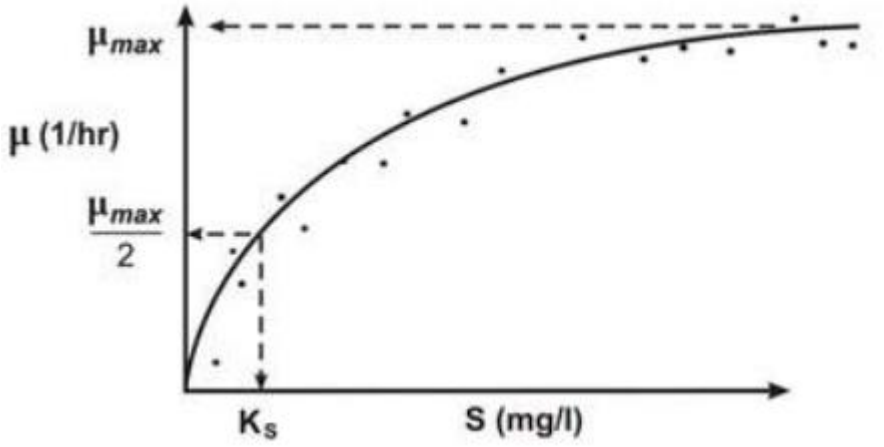


Figure 2.7 *Monod kinetic [25]*

2.4 Factors Affecting the Anaerobic treatment processes

The purpose of this section is that to introduce important factors in the process design and operation of the anaerobic treatment processes. nutritional requirement of microorganisms, operating condition and the environmental factors are factors that require to consider them for the anaerobic processes due to their effects on the microbial growth.

2.4.1 Sludge retention time (SRT)

The solids retention time is the fundamental control parameter to determine the performance of all anaerobic processes [2]. The SRT is defined as the ratio between the sludge mass in the system and sludge wasted (produced) from the system.

$$SRT = \frac{(Mass\ of\ sludge\ in\ system)}{(Mass\ of\ sludge\ wasted\ from\ system)} \quad (2.13)$$

In biological terms the SRT is linked to the growth rate of the biomass:

$$\frac{1}{SRT} = \mu - k_d \Rightarrow SRT = \frac{1}{\mu - k_d} \quad (2.14)$$

Where μ is specific growth rate and K_d is Endogenous decay coefficient. Equation 2.14 shows the relation between SRT and growth rate, indicating that microorganisms with low growth rate (such as methanogens) requires long SRT in order to grow in the system. In practical terms the SRT is controlled by the sludge wasting rate shown in equation 2.13.

SRT affects to the performance of microorganisms in the reactor. Higher SRT provides a greater mass of methanogenesis bacteria. This can become limiting factor to maintain stability of system because of the balance between volatile fatty acid production and consumption. In figure 2.8 presents SRT values for various anaerobic microorganism processes at 35 °C [2].

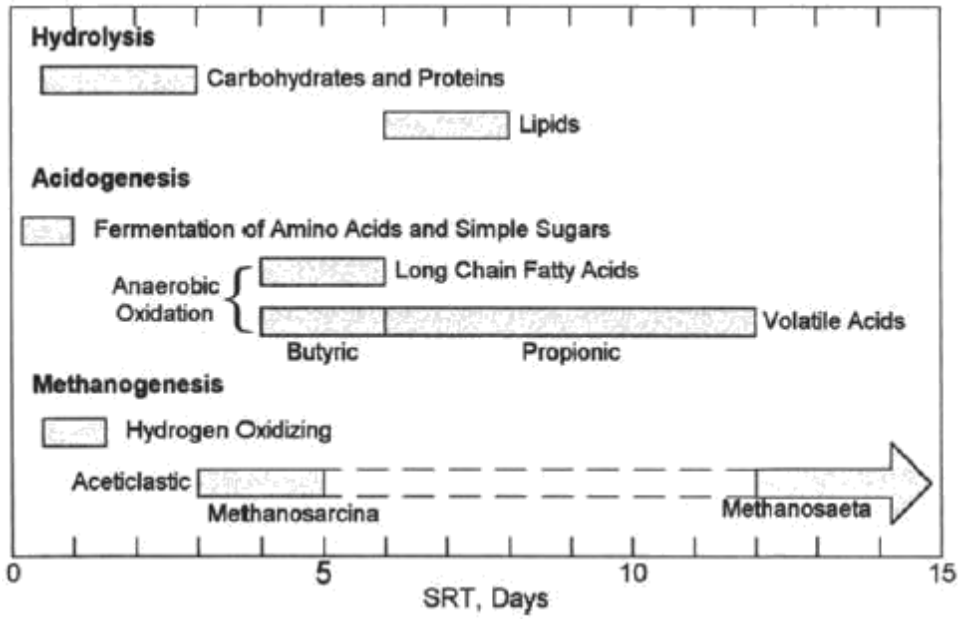


Figure 2.8 *Typical SRT ranges for various anaerobic microorganism processes at 35 °C [14]*

Temperature has significant effect in the SRT. Lower temperatures result in lower growth rate and requires longer SRT. Recommended SRT values as a function of temperature for domestic

wastewater by UASB reactors are provided in table 2.10. Prediction of SRT and biomass concentration in granular sludge in the UASB reactor is not possible to do accurately.

Table 2.6 *Recommended UASB SRTs for domestic wastewater treatment plant [25]*

Temperature (°C)	SRT,d
35	25
30	30
25	60
20	100
15	140

2.4.2 Hydraulic Retention time(HRT)

The volume of process per unit of flow rate of influent is known as the hydraulic retention time. (Equation 2.4) [12]. wastewater with higher substrate concentration will need longer hydraulic retention time for given OLR.

$$HRT = \frac{Volume\ of\ aeration\ tank}{influent\ flow\ rate} \tag{2.15}$$

SRT and HRT are both related to the time in the anaerobic treatment system .SRT gives the sludge age that means the length of time the material remains in the reactor. HRT is the nominal value, which expresses the length of time the liquid and dissolved material remains in the reactor. When there is not any sludge recycle or retention in the reactor, SRT and HRT are equal. The reactor with sludge recycle, the SRT does not have the same value as the HRT. The link between SRT and HRT is neither proportional nor linear. It depends more on the COD or BOD, and TSS [12].

2.4.3 Organic Loading Rate (OLR)

The volumetric organic loading rate (OLR) is the key design factor to determine the bioreactor volume for a particular process quantifies of anaerobic treatment process. The volumetric organic loading rate usually have units as kg COD/ m3. d and is shown in equation 2.16. From 1 to 50 kg COD/ m^3. d is the range of OLR for anaerobic treatment process, which is higher than OLR level (0.5 to 3.2 kg COD/ m^3. d) for aerobic treatment process. The type of anaerobic process used, type of wastewater, and temperature have influence on organic loading rates [25].

$$OLR = \frac{Q.C_{in}}{V} \tag{2.16}$$

Where Q is flow rate (l/d); C_{in} is COD for influent (gCOD/l); and V is reactor volume (l) [25]. This equation can be also made relationship between OLR and HRT by the equation 2.17. This equation shows that the OLR is inversely proportional to the HRT [2].

$$OLR = \frac{C_{in}}{HRT} \tag{2.17}$$

Organic loading rate and sludge retention time (SRT) are related inversely proportional to each other through active biomass concentration, shown in equation 2.18[2]. For a sufficient SRT in the anaerobic reactor requires larger volume and lower organic loading can be realized from equation 2.7[25].

$$SRT = \frac{X.V}{Y.Q.C_{in}} = \frac{X}{Y.OLR} \tag{2.18}$$

2.4.4 Temperature

Temperature has a significant effect on the anaerobic processes as to all other biological processes. As the temperature is increased, the growth rates of microorganisms will increase as well. Having a stable temperature is more important than higher and unstable temperatures, because sudden changes in temperature negatively affects the growth rates of the bacteria [9]. The temperature effect can be expressed as Equation 2.19.

$$\mu_{m(20)} = \mu_{m(T)}.\theta^{(T-20)} \tag{2.19}$$

Where $\mu_{m\,(20)}$ is maximum specific growth rate at 20 ºC; $\mu_{m(T)}$ is maximum specific growth rate at temperature, T ºC; and θ is temperature coefficient.

Microorganisms are categorized into various temperature levels based on the optimum temperature. Figure 2.9 presents the growth rates of methanogens bacteria in different temperature by defining classic groups of psychrophilic (0 – 20 ^{0} C), mesophilic (20 – 42 ^{0} C) and thermophilic (42 – 75 ^{0} C) [9,10]. The anaerobic reactors perform better in the mesophilic and thermophilic temperature ranges. It is also possible to operate the bioreactor at lower temperatures approaching 10 ºC, then the growth rates of methanogens will be limited and VFAs may accumulate in the bioreactor. Methane can be produced at temperatures down to 10 ^{0} C or lower, but for optimum production the temperature should be maintained above 20^0 C. Longer SRTs and more energy for heating are required to compensate for the lower temperatures [2,34,35]. However, some researchers indicate that anaerobic processes can operate successfully at low temperature, but most anaerobic operations are designed in the mesophilic temperature range [2]. In moderate climate conditions, dilute industrial and domestic wastewaters (below 1500 mg COD/l) are discharged at low ambient temperatures [36, 37]. Generally, it is recommended to designed and operated system with variations less than +/- 1 ºC each day [2].

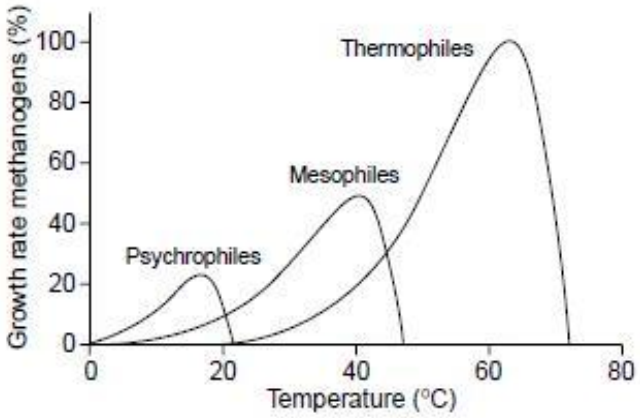


Figure 2.9 *growth rates of methanogens with different temperature ranges [14]*

2.4.5 pH

The pH of the medium culture exerts an effect on the bacteria activities in biological wastewater treatment processes. The pH range for adequate activity of microorganisms are between 6.0 and

8.0, whereas a pH range of 6.8 to 7.4 provides optimum conditions for methane microorganisms. Deviations of pH beyond 6.8 and 7.4 (from 6.0 to 8.0 range) may lead to a significant decrease in methanogenic activity [2,12]. Comment: Does the low pH result in VFA, or does the VFA cause drop in pH. This is not so clear here. A decrease in pH leads to produce the higher molecular weight VFAs, articulately propionic and butyric acid, and can result in an unstable anaerobic operation. If the VFAs production rate oversteps from the capacity of methanogenesis bacteria, excess VFAs will accumulate, reducing the pH. This condition is called "sour" or "stuck" anaerobic process. OLR reduction can resolve this phenomenon to reach the point which VFA production rate is less than VFA consumption rate. Thereby, it restores the neutral environment and methanogens activity will increase. In extreme cases, the chemical sources of alkalinity such as lime or sodium bicarbonate should be added to the system [2]. However, VFAs will react with bicarbonate alkalinity, and the production of carbon dioxide will become a part of the biogas [2, 38]. The influent alkalinity should be controlled to maintain pH levels of the bioreactor to stabilize the methanogenic activity [25].

2.4.6 Volatile Fatty Acids (VFAs)

The effects of VFA is closely linked to pH and alkalinity, as the VFA is the most common cause of affecting the pH in anaerobic reactors. Therefore, a balance between the acidogenic and methanogenic processes is needed. So normally long time before the high levels of VFA that become toxic in itself, the high VFA have caused the pH to drop so drastically that it is the low pH that inhibits the process and not the high VFA, unless there is extremely high alkalinity. Volatile fatty acids can affect the pH of the medium in the anaerobic reactors. "*When the pH is held constant near neutral pH, neither acetic nor butyric acids have any significant toxic effects upon hydrogen-utilizing methanogenic bacteria at concentrations up to 10000 mg/l* "."*Propionic acids, on the other hand, exhibits partial toxicity to methanogenic bacteria at a concentration of 1000 mg/l at neutral pH*" [9]. So, propionic acid can retard toxic effects with high concentration in the anaerobic processes, however, this statement does not prove for acetic and butyric acids. Consequently, VFA will have probably little inhibition effect at neutral pH [9].

In anaerobic treatment process, methanogenesis bacteria are sensitive to pH ranges. A decreased pH will lead to souring in the bioreactor, i.e. a sudden pH drop. As mentioned in pH sub-chapter, this will cause to accumulate the acetic acid or VFA s, and decreased subsequent pH. Acidifiers

are active even low pH (4) until the methanogenic capacity of system will be over [12]. This process is illustrated in figure 2.10.

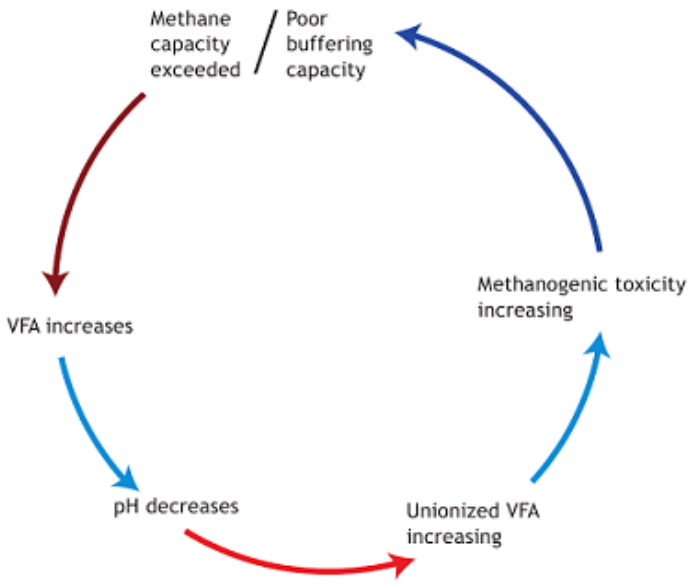


Figure 2.10 *Reactor pH drop as a result of methanogenic overloading and accumulating VFAs [12]*

2.5 Up-flow Anaerobic Sludge Blanket (UASB) Reactor

2.5.1 General Concept

The up-flow anaerobic sludge blanket (UASB) reactor is the most successfully anaerobic technology for treating the various types of wastewater. The UASB reactor consists of a circular or rectangular tank in which wastewater is introduced at the bottom of the reactor and flow in an upward at velocity which coordinates with the settling velocity of biomass. In this way, the Upflow reactor is formed with a bottom sludge bed, dense, and granular anaerobic biomass which occupies about half of the volume reactor. During the passage of the wastewater through the compact granules, the treatment process converts the organic matter to biogas and sludge. A gas-liquid-solid separator is at the top of reactor to separate granular solids from the effluent and collect biogas. The solid particles fall back to sludge blanket, while the produced gases are captured at the top of the reactor [12, 25]. Process loadings, upflow velocity, wastewater type, and settling characteristics of solids are the factors to affect the dimensions of UASB reactor. The UASB reactors are suitable for treating wastewater with low substrate concentration due to

the retention of high levels of active biomass in the reactor [2]. High retention biomass concentration causes to operate the anaerobic treatment at high organic loading rates [25]. The size of granular sludge particle is in the range of 1 to 2 mm, depending on the waste treated and hydraulic gas shear forces. Particle densities are in the range of 1 to 1.05 g/l with settling velocities of 5 to 50 m/h [12]. The granular sludge bed develops during some months, but this development is more rapid at higher temperatures (above 20 ºC) and with the presence of readily degradable soluble COD in the feed. Washing out unattached organism is common at high upflow velocities, and also the maintenance of granular sludge is affected by the wastewater characteristics. pH, divalent cations, and nutrient addition are also affected in the development of the granules.

2.5.2 Design Considerations

The main factors for UASB design which are required to consider are upflow velocity and organic loading rate (OLR). When the UASB reactor is used for domestic wastewater or wastewater with higher influent solids concentrations, lower upflow velocity is required to remain the solids in the reactor for providing the sufficient time for hydrolysis. The upflow velocity is determined by equation 2.20.

$$A=\frac{Q}{v} \tag{2.20}$$

Where v is maximum design upflow velocity (m/h), A is reactor cross-section rate (m^2), and Q is influent flowrate(m^3/h).

OLR is described in the section 2.4.3 and it depends on wastewater characteristics, reactor type, and temperature. The OLR which is described in equation 2.16, affects in the reactor volume [25]. Higher organic loading can be applied in UASB reactor compared to aerobic processes. Hence, less reactor volume is needed, as well as benefits of production of energy as biogas [4].Figure 2.11 is illustrated the schematic of UASB reactor.

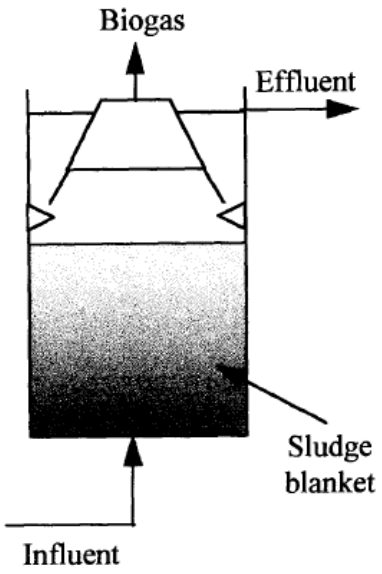


Figure 2.11 *Schematic of UASB [2]*

2.5.3 Application of UASB

The UASB reactor was developed for the treatment of high strength wastewaters. The success of UASB reactor attributes to its capacity for retaining a high concentration of sludge, and solid-liquid-gas phase separation [12]. The UASB reactor was tested for the first time for sewage treatment which was started during the early eighties in Cali, Colombia. The results achieved form the operation of the 64 m^3 UASB pilot scale reactor that presented the applicability of the system under the prevailing environmental and municipal wastewater characteristics. The initial trials were rapidly followed by full scale reactors in Colombia, Brazil and India [12]. Soon after some researches for UASB reactor was started to determine its potential with different types of wastewaters, under various types of temperatures condition, and in both full scale and pilot scale [39].

The application of UASB has been studied under psychrophilic and low mesophilic condition since 1976 [12, 2, 39, 38]. Seghezzo reported that a 6 m^3 UASB reactor with digested sewage sludge was operated at hydraulic retention times (HRT) of 14 – 17 h. His research achievements were showed 85 – 65% and 70 – 55% COD removal Efficiency at 20 °C and 13 – 17 °C, respectively [4]. He concluded that the UASB reactor was a compact, inexpensive, and suitable technology for sewage treatment, even at low temperatures [4]. Based on research carried out on different UASB reactors, the results concluded UASB reactors can treat low to high strength

domestic sewage (500 – 1500 mg COD/l) at 12 – 18 ℃ with HRT of 7 – 12 h with total COD removal efficiencies of 40 – 60%. The UASB reactor and the post-treatment step can be performed in more integrated set-up [12]. COD removal efficiency above 70 % in UASB reactors at HRT 1.4h with 15 g COD/ l.d organic loading rate were successful obtained in UASB reactor studies that was conducted at University of Stavanger [6].

Chapter 3

Materials & Methods

This chapter describes the pilot-scale experiment of the anaerobic treatment for removing COD of the industrial wastewater from DAF effluent at IVAR Grødaland plant, using UASB reactor. In this study, UASB pilot plant was considered for investigating the UASB reactor performance, and its potential to remove COD and to produce biogas in actuality for about four months from February till May. Third chapter also includes operational, maintenance, and control procedures, along with analytical methods for the pilot project. All experiments and laboratory works for this master's thesis project were conducted at the IVAR Grødaland wastewater plant (WWTP).

3.1 Characterization of wastewater

The wastewater of Grødaland plant receives mainly of industrial wastewater. The treatment process in this plant starts with a pre-treatment by sand and fat removal followed by a combined chemical-biological treatment. Chemical Treatment includes flocculation tanks and dissolved air flotation (DAF) unit with addition of polymer [1]. The treated wastewater from the DAF unit, it will enter to the anaerobic treatment stage which was tested at pilot scale in this study. Sequential batch reactors (SBR)covers the last step for treatment of the wastewater in the full-scale plant. The characteristics of influent to the UASB reactor was pretty much the same as outlet of the Flotation unit which is shown in table 3.1.

Table 3.1 *The DAF effluent unit average data 2017, and UASB inlet [1]*

	Outlet Flotation unit(Inlet UASB Unit)
SS (mg/L)	160
Dissolved COD (mg/L)	647
Total COD (mg/L)	912
BOD (mg/L)	646
Total Phosphorus (mg/L)	22,2
Conductivity (µS/cm)	2006
PH	7,48
Flow (m^3/d)	5453

3.2 Pilot Plant Configuration

The pilot plant was constructed and consist of an Upflow Anaerobic Sludge Blanket reactor(UASB), peristaltic pump for feeding the UASB, inlet storage tank, and main pump for wastewater supply to set up primarily the anaerobic treatment stage. Other instruments and elements were also used to operate the system more effectively such as gas detector.

UASB reactor was designed by Waterment company in total capacity of 18 L, and granules were constituted approximately 9 L of the total volume, as shown in figure 3.1. The UASB was seed with active granular sludge from a slurry factory two times during the period of experiment. Making the sludge bed for UASB was done by 5 L during the starting up the plant in February, and granules were also added for the second time in April. UASB reactor was fitted with three main ports to allow for feeding, withdrawal of effluent, and gases extraction which was air tight to avoid leakage or entrance the air to the UASB reactor. UASB had two other ports on top of the reactor with rubber stoppers, and manual valve, for addition of the granules to the sludge bed.

Figure 3.1 *Photo of the Upflow Anaerobic Sludge Blanket*

Figure 3.3 presents the process flow diagram (PFD) of the UASB pilot plant. Considering the PFD, the process of anaerobic treatment started up with the wastewater transportation from DAF unit to the storage tank by a drainage &wastewater submersible pump with constant flow rate. Two digital dosing pumps and bilge pump was previously used to feed the plant for two months, but they were not appropriate types of the pump for this stage. A peristaltic pump supplied the wastewater for the reactor with adjustable flow rate, and time. The operation of feed pump was set in such manner that for 10 minutes in an hour to pump water to the digester in about 700 ml flow rate during two first months, and 1000 ml for April till May. Reactor effluent flowed out to the outlet tank due to gravity, then treated wastewater went out to the drain. Figure 3.3 illustrates the photo of the pilot plant. The generated gas was monitored by variable methods and devices. At the initial set up, two gas counter were installed to record the biogas production. However, they did not work properly due to issues such as leakage or pressure. A liquid displacement system was another method that was used to monitor the volume of produced gas. For liquid displacement method, a small diameter rubber pipe was connected from gas hole on the top of the reactor to the graduated cylinder which was filled with water [10]. The graduated cylinder measured the gas volume, but however, analysis for

the gas composition was not possible by this method. A gas detector was used every day for a short period from mid of March to mid of April to assure the production of methane in the reactor. Table 3.1 describes the equipment used in the pilot plant.

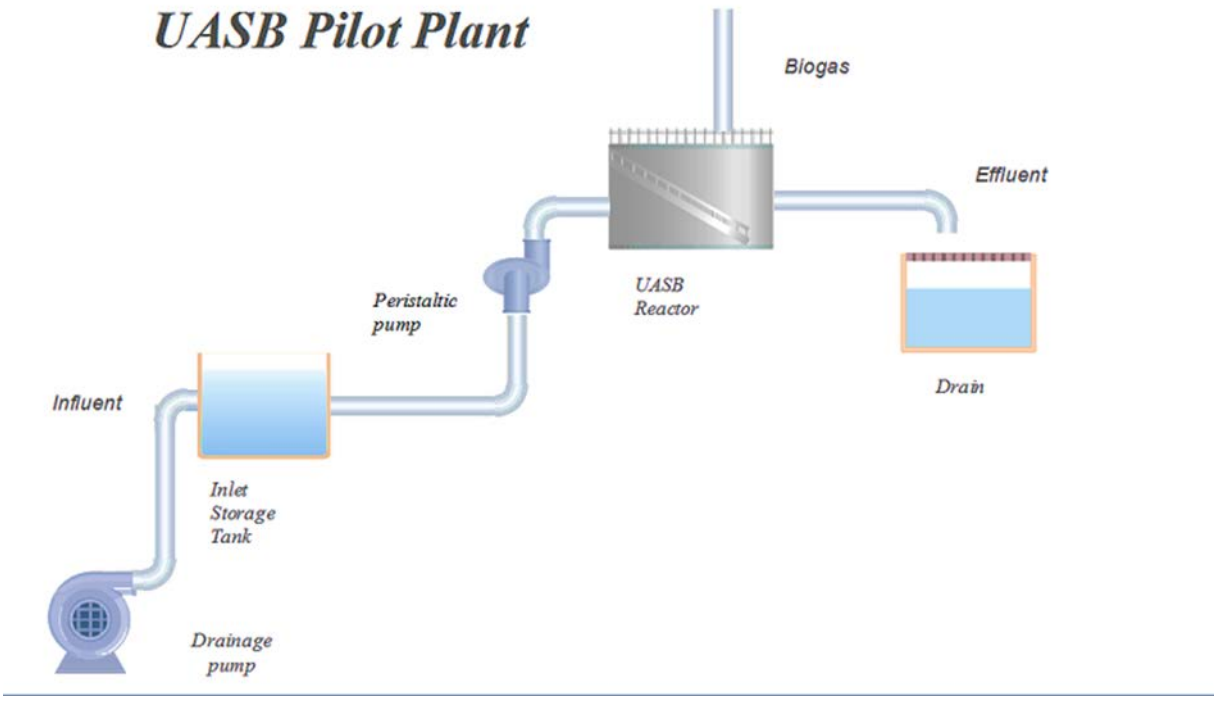


Figure 3.2 *Process Flow Diagram (PFD) of UASB Pilot Plant*

Table3.2 *Equipment characteristics for the UASB pilot plant*

Equipment	Manufacturer	Properties	
Feed and recirculation pump	Ismatec	Type	Peristaltic pump
		Model	Reglo ICC
		Flowrate	0 – 43 ml/min
	Biltema	Type	Bilge pump
		Model	259750

Bilge pump		Flowrate	60– 940 l/h
Digital dosing pump	Grundfos	Type	Digital dosing pump
		Model	DME
		Flowrate	0 – 43 ml/min
Drainage&wastewater submersible pump	KSB	Type	KSB Ama porter
		Model	500 NE
		Flowrate	0-40 m^3/h
		Motor Power	1.5 kW three phase
			1.1 Kw single phase
Gas Counters	Ritter	Model	MGC-1 V3.3 PMMA
		Gas flowrate	1 ml/h – 1 l/h
		Max. pressure	100 mbar
		Min. pressure	5 mbar
		Measuring accuracy	Less than ±1 %
Gas Detector	Dräger	Model	Dräger – x-am 7000

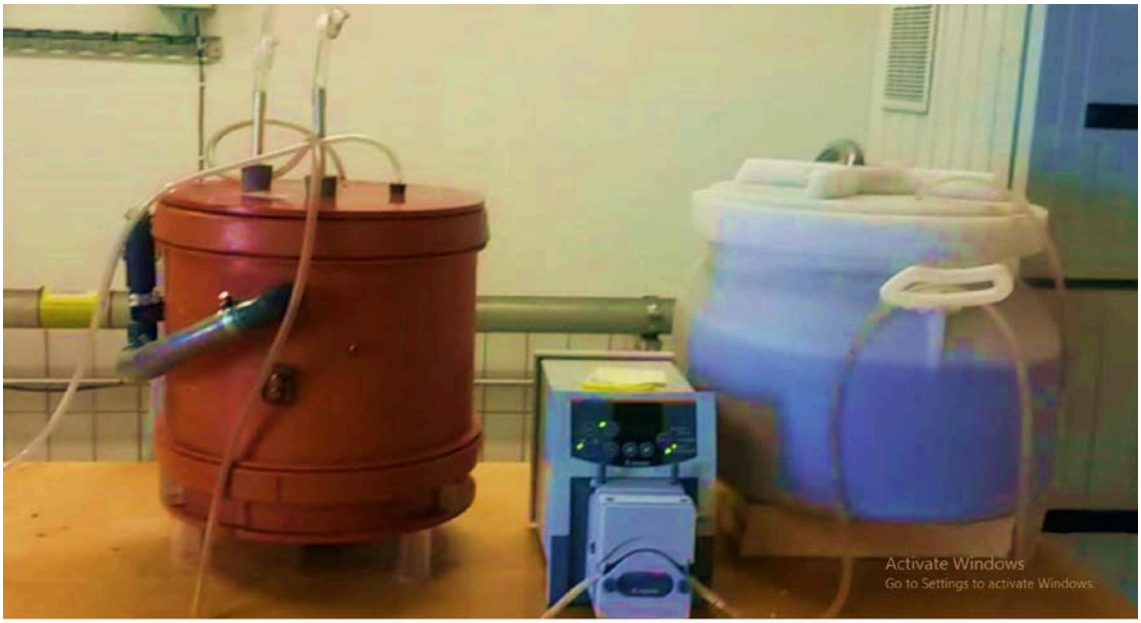


Figure 3.3 *Photo of the UASB pilot plant*

3.3 Operations management of Pilot Plant

In this master thesis, all operations management of UASB pilot plant was conducted to investigate UASB reactor performance for COD removal and analyze the gas production from 15th February to 30th May under ambient temperature between 15 and 23 °C. This section describes the starting-up process and operation conditions of UASB pilot unit.

Figure 3.4 presents the operation flow chart of the system. The first stage of starting up the pilot scale was hydraulically stable which was initially tested with tap water without sludge bed to make sure that all equipment, and instruments were installed and worked correctly. The success of UASB operation depended on granular sludge that is presented in the UASB reactor environment. The granules cultivated were acclimatized to the reactor in first five days, until then the system was running and it was assumed to be in steady-state condition. Required time for achieving the steady state condition in the reactor could be determined by OLR and SLR. steady state condition was achieved in the UASB reactor when the parameters, e.g. the effluent COD remained constant at the same OLR [9]. The operation of UASB system was started at an 0.5gCOD/l. d and a hydraulic retention time(HRT) of 26.4 h. Operating factors such as, OLR, pH, COD, flow rate were controlled to investigate the development of the anaerobic treatment process. The flow rate was adjusted to draw wastewater at two different rates, 700 ml/h, and 1000ml/h. Increasing the flow rate was done due to the reactor clogging which was temporarily

resolve by compressor Michelin 8 bar, wash out the sedimentations. The flow rate was also measured manually every time by graduate cylinder per unit to control the flow rate in the peristaltic pump display. The measurement of flow rate for influent and effluent was represented the Equal flow rate in the process.

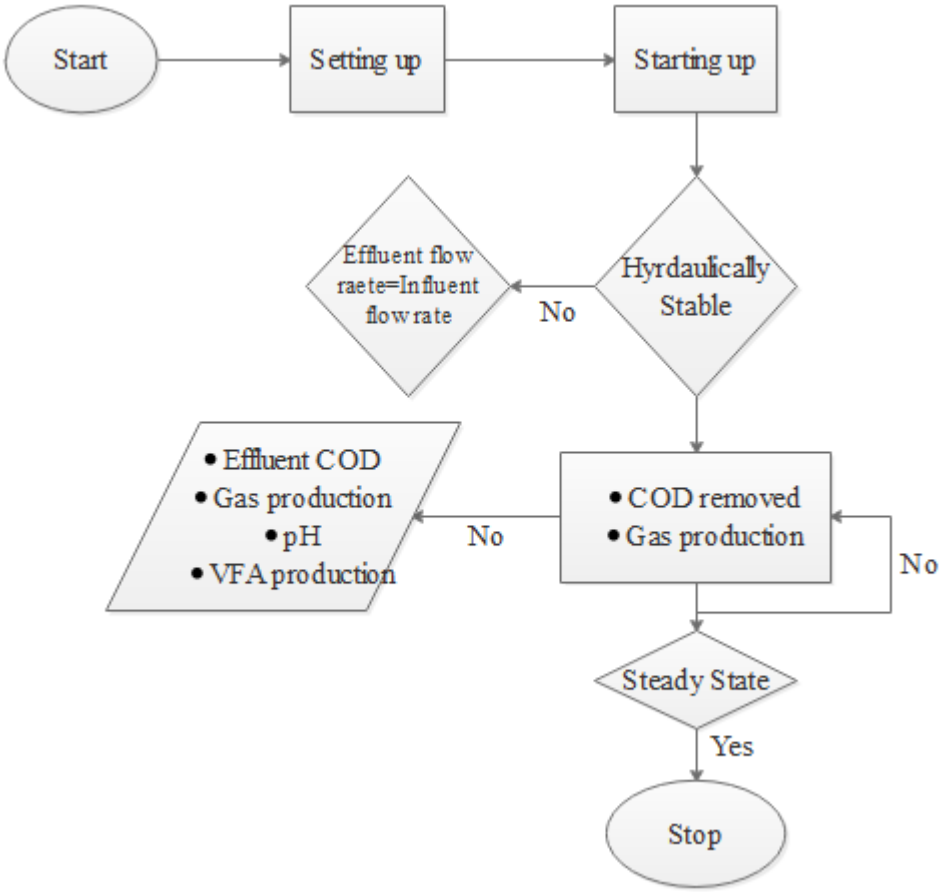


Figure 3.4 *Operations management flow chart of UASB pilot plant*

3.4 Analytical Methods

This part explains the analytical methods for evaluation of the anaerobic treatment process. For the analytical analysis of wastewater, Samples were collected from inlet storage tank, and UASB effluent on selected days from one to several samples a day. Samples were homogenized by shaking to make sure sufficient particles distribution was achieved. Diluting was also needed for some analyses, and it was done by deionized water. The following operation parameters were analyzed for performance evaluation of the UASB pilot plant:

- pH and conductivity
- Total volatile fatty acid and alkalinity
- Chemical Oxygen Demand (COD)
- Total suspended solids(TSS)& Volatile Suspended solids(VSS)

3.4.1 pH and Conductivity

pH and Conductivity was measured from inlet storage tank and effluent by using WTW Multi 340i multimeter which was attached to conductivity probe (WTW Tetra Con® 325) and pH probe (WTW pH-Electrode SenTix 41). The multimeter also was used to measure the pH in the titration for determination of VFA and alkalinity. Recording the temperature for influent and effluent also was done by pH meter.

3.4.2 Total Volatile Fatty Acids(VFA) and Alkalinity

Volatile fatty acids concentration and alkalinity were determined by 5-pH point titration method. The procedure was done on filtration 50 ml of samples by using a GF 6 Glass microfiber filter paper. Filtered samples were placed on a magnetic stirring on *IKA® Big Squid white* at low rotation in about 200 rpm to minimize CO_2 input or loss [40]. The initial pH, temperature, and conductivity were noted in this step. The samples were titrated to reach suitable pH values at about 6.7, 5.2, 5.9, and 4.3 +/- 0.1 with HCl 0.1 M, and the volume of acid consumption was recorded for each pH selected point. All data were entered in to the computer program (TITRA5) for the calculation of volatile fatty acids concentration as acetic acid(HAc) mg/l and alkalinity as mg/l $CaCO_3$.

3.4.3 Chemical Oxygen Demand (COD)

COD measurement was the most significant analysis in the UASB pilot plant. Total and dissolved COD was analyzed for the feed wastewater and UASB effluent. COD measurements were conducted three times a week from February till April to control the reactor performance. From April till June, total COD was measured almost every day to observe the progress of the reactor for COD removal, while dissolved COD was checked once a week.

The prepared COD cell kits from Merck Spectroquant® with Product Number 109773, were used to perform the analytical analysis of UASB reactor samples. The COD test kits facilitated measurements in the 25-1500mg/l of COD concentration range.

For preparation the dissolved COD, samples were filtered by using glass microfiber filters(GF/C) with 1 µm pore size. The samples for the total COD were homogenized by mechanical mixer (T 25 digital ULTRA-TURRAX®) in about 1 minute. The samples were heated at 148 °C for 2 h in a thermoreactor (TR620). First step for the COD measurement was mixing, and heating. COD cell tests were swirled gently because of the suspended sediment in the bottom. 3ml of sample were added to the cell tests, and strongly mixed the contents of the cells. Then, vials were heated in the thermoreactor. Second step was cooling. Cell tests were placed in a metal test tube rack and to cooled down to room temperature about 25 °C, and kits were swirled after 10 minutes of cooling. COD concentrations were read by spectrometer (Spectroquant Pharo 300, MERCK). For COD measurements, it is possible to store the vials in refrigerator with H_2SO_4 as about 1% of the sample volume until actual analysis.

3.4.4 TSS and VSS

The First step for solid analysis was weighting of the filter and porcelain with Sartorius laboratory balances. A volume of samples filtered through glass microfiber filters(GF/C), were recorded to calculate the concentration(mg/l). Filters were dried at temperature 105°C for at least 2 hours, then filter and porcelain were cooled in desiccator before weighting for the total suspended solid determination. The increase in weight of the filter represented the total suspended solids(TSS).

$$\text{Total Suspended Solids}\left(\frac{\text{mg}}{\text{l}}\right) = \frac{(\text{B} - \text{A})(\text{mg})}{\text{Sample Volume (l)}}$$

A=Weight of porcelain and filter before drying

B= Weight of porcelain and filter after drying

For volatile suspended solid(VSS), the residue retained on the filter was ignited in muffle furnace LT 5/12 oven at 550°C. The remaining solids represents the inorganic suspended solid while the weight lost on ignition is the volatile solids [42].

$$(\text{FSS})\text{Fixed Suspended Solids}\left(\frac{\text{mg}}{\text{l}}\right) = \frac{(\text{A}-\text{B})(\text{mg})}{\text{Sample Volume (l)}}$$

$$\text{TSS} - \text{FSS} = \text{VSS}$$

A = weight of residue + dish before ignition

B = weight of residue + dish or filter after ignition

Chapter 4

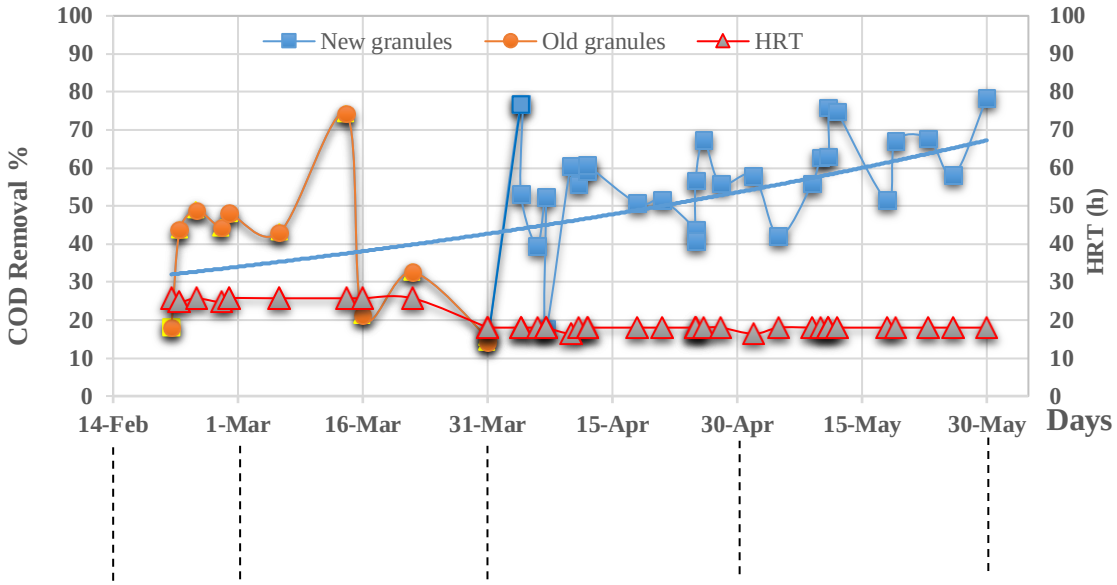
Results

This chapter presents the results obtained in the experiments. All the data and figures are summarized in this chapter while the collected raw data are attached in the Appendixes.

4.1 COD removal efficiency

The UASB reactor performance was monitored by COD removal efficiency and COD conversion. As gas flow monitoring appeared difficult with the equipment available, the COD conversion was used to calculate methane gas produced, assuming COD removed was converted to methane gas. Figure 4.1 and 4.2 presents COD removal efficiency and OLR versus HRT during the experimental period respectively. At the end of March, the influent tube was clogged and the wastewater pump was broken. As results of the technical problems, the lowest COD removal rate was observed at 14 % with 18(h) hydraulic retention time during this period. The trend for COD removal efficiency indicates a gradual increase to over 50 % after the addition of granules for the second time, and decreasing the HRT in the first week of April, in order to the leap giant, 78% efficiency in the last day of May. The COD removal efficiency was variable with the old granules. The COD removal efficiency had a progressive increase during the experimental period and seemed not to be significantly affected by temperature. (Table 4.1). The OLR was lowest in the start-up of the experiment (0.53 g COD/l. d), and then gradually increase, until the HRT was decreased from 1.1 to 0.8 days. The OLR of the system was remarkably increased from 0.53 to 1.9 g of COD liter^{-1} day^{-1} at an average temperature of 19 ℃ (Figure 4.1 b). The OLR was achieved the peak loads of 1.9 g COD/l. d when the COD removal efficiency reached 77% at fourth of April. The system maintained considerably

stability and high efficiency from mid of April, where HRT was constant at 0.8 day (18 h), and the OLR raised up to 1.5 g COD/l. d.



Month	Feb	Mar	Apr	May
Avg Temp	15℃	17℃	22℃	22.5℃

Figure 4.1 *COD removal efficiency of UASB reactor versus HRT over the experimental period*

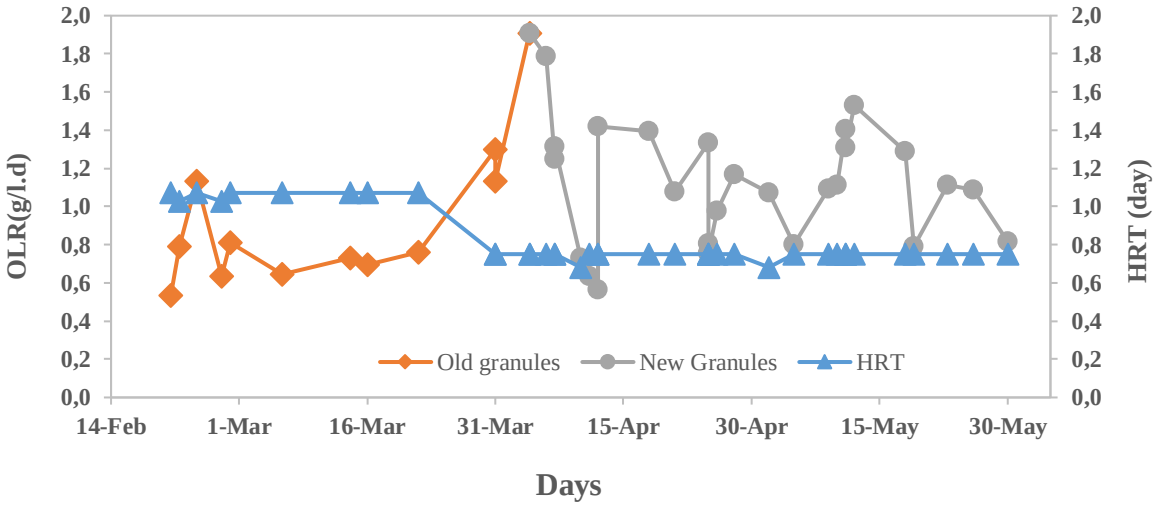


Figure 4.2 *Time course of OLR and HRT*

Figure 4.3 illustrates the relationship between COD removal and OLR in the UASB reactor. The effect of OLR on COD removed can be realized by constant OLR, i.e. 1 g COD/l. d to 1.2 g COD/l. d of OLR, COD removed increased slightly. The trend follows linearly the effect of OLR on COD removed at new granules. Table 4.1 presents the average results of each month based on the raw collected data. COD removal efficiency was calculated with two different methods: (a) the average COD_{in} and the average COD_{ef} of each month was used to obtain the COD removal efficiency of that month, (b) All COD removal efficiencies of each month were averaged from raw collected results.

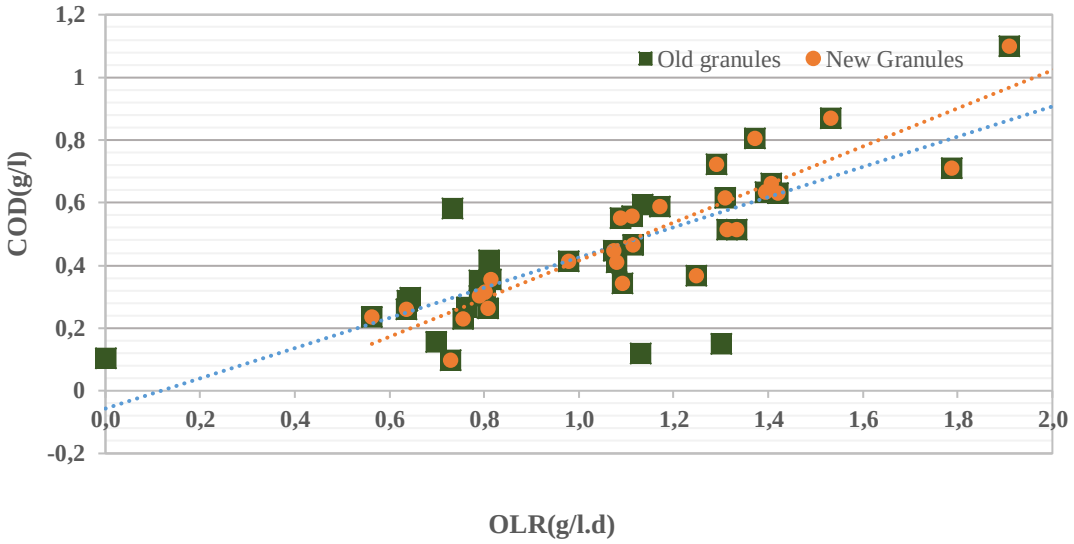


Figure 4.3 *Effect of OLR on COD removal*

4.2 TSS and VSS

Variations of effluent and influent suspended solids (SS) are shown the figure 4.4 from February until May. TSS concentration in the influent ranged from 120 mg/l to 383 mg/l. The highest value for TSS effluent was on the day 5, at the end of March, which can be explained by reactor clogging and technical problems for wastewater pump. The clogging of the reactor was resolved by high-pressure water, hence, some sludge particles were washed out by water to the effluent. The UASB always generated effluent samples with clear color, which shows that the granular sludge in UASB had good settling properties. All concentrations of TSS in effluent were lower than TSS influent concentrations as shown in figure 4.4, and TSS removal efficiency were over 40 % (figure 4.5). TSS and VSS removal followed a linear ratio with regression of

0.96(VSS/TSS-ratio**)** in figure 4.5 that means removal rates for both of organic and inorganic suspended solids were similar.

Table 4.1 *Average characteristics of Influent and effluent of UASB pilot Scale during four months*

Month	Temperature (°C)	HRT (h)	OLR (gCOD/l.d)	COD_{in} (mg/l)	COD_{eff} mg/l)	COD [b] Removal %	COD [a] Removal %
February	17.5	26	0.975	824	473	40	42
March	18	26	1.054	810	548	40	32
April	19	18	1.086	850	340	56	60
May	20	18	1.24	852	288	69	66

(a) Average value was calculated based on the average COD_{in} and COD_{eff} of the table 4.1.

(b) Values were averaged from the COD removal efficiencies of each month from raw collected data.

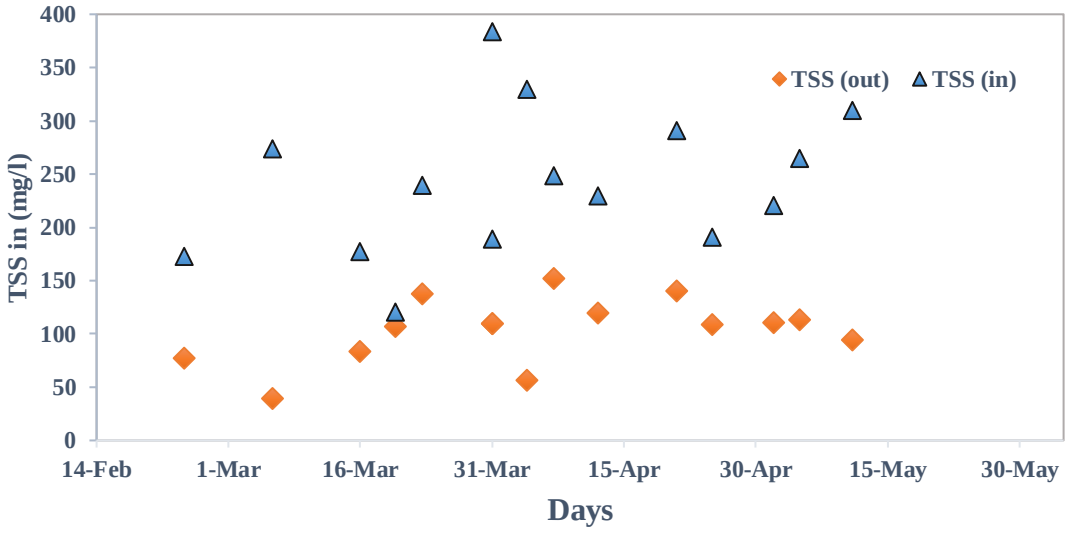


Figure 4.4 *TSS influent and TSS effluent over the experimental period*

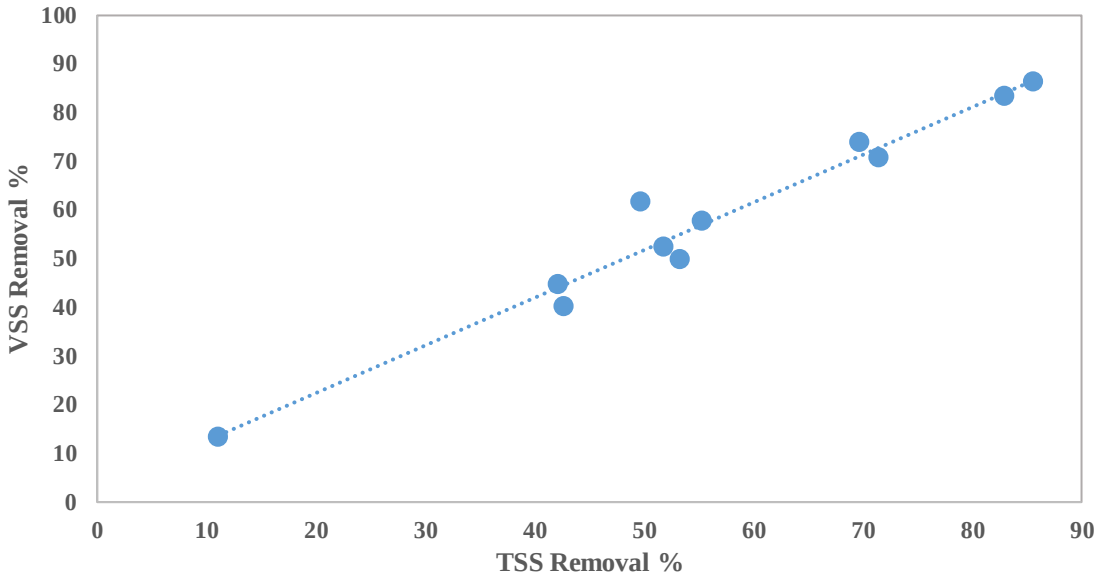


Figure 4.5 *Relationship between VSS and TSS removal*

4.3 Methane Production

Biogas composition was mainly 65% methane and 35 % CO_2. There were no accurate measurements for the gas volume produced, while gas composition was measured during the experiment. Two methods were applied to analyse the gas produced by UASB reactor, as mentioned in chapter 3.A Gas detector measured volume of methane in percentage and the results of this method are shown in the table 4.2. The second method was liquid displacement, which monitored just gas volume production, by a graduated cylinder. It showed 300 ml gas production during three hours of measurement, but was not an accurate and trustable way to analyze the results of this method. Table 4.3 also reports the gas composition of UASB plant in June.

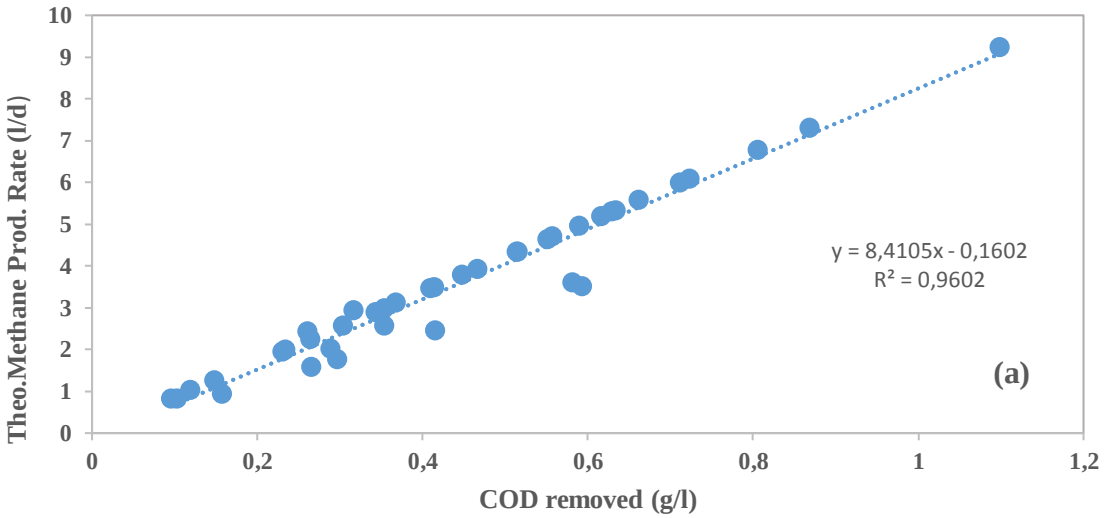
Table 4.2 *Composition of Methane by gas detector*

Date	Methane%	COD_{in}(mg/l)	COD_{eff}(mg/l)	OLR(g/l.d)
29.3.2017	30	1524	703	0.12
31.3.2017	43	975	826	0.149
7.4.2017	47	985	469	0.516
11.4.2017	60	422	187	0.235

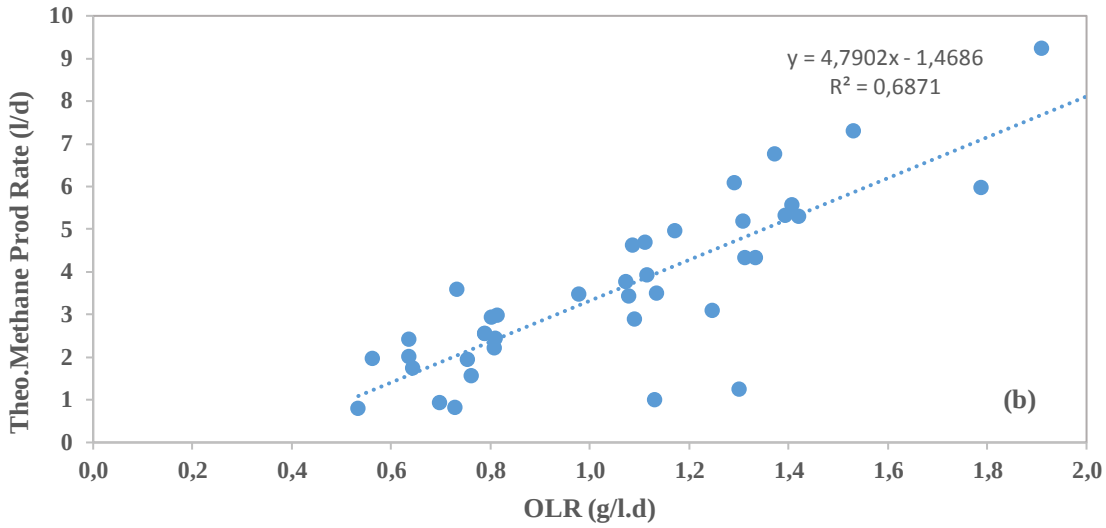
Table 4.3 *Composition of Methane gas in June*

Date	Methane%	Description
14.06.2017	80	pumping feed into reactor first time
	70	pumping feed into reactor next time 30 min later
	62	no pumping of feed into reactor about 1 h after previous measurement
21.06.2017	80	no pumping feed into reactor
	68	pumping feed into reactor first time 20 min later
	63	pumping of feed into reactor about 1 h after previous measurement

Figure 4.6 presents the theoretical methane production as a function of (a) COD removed and (b) OLR, using the 0.35 l CH_4/ g COD relation. The correlation between COD removal and methane production rate are y = 8.2339x - 0.0904 and the regression coefficient is (R^2=0.9514). The Relationship between OLR and methane production rate were observed as a direct link from figure 4.6(b), as the OLR increases, the theoretical methane production rate increases as well.



(a)



(b)

Figure 4.6 *effect of theoretical methane production on (a) COD removed, (b) OLR*

4.4 pH, alkalinity, and VFA variability

The pH of the influent and effluent are presented in figure 4.7. The pH values were within the range of 6.5 and 8.5. The pH variations for the effluent increased from 6.80 to 7.70 during the experiment, indicating a gradually increasing alkaline environment. which is provided the alkaline environment slightly. The pH level (7.5 ± 0.5) were within the optimum value for growth of methanogens bacteria. The Alkaline condition in the UASB reactor indicated that the VFA production did not affect the pH significantly. Therefore, the reactor content had sufficient buffer capacity to maintain a neutral pH. In figure 4.7, it can be seen that the pH values for effluent were higher than influent values, due to removal of VFA and production of alkalinity in methanogenesis. The effect of pH on VFA profiles for both influent and effluent are illustrated in figure 4.8. At 7.5 ± 0.5 pH range, volatile fatty acids concentration of effluent was at highest concentration in about 240 mg /l to zero mg/l at lowest concentration. Complete consumption of VFA by methanogens for production of biogas happened in optimum range of pH(6.8- 7.4).

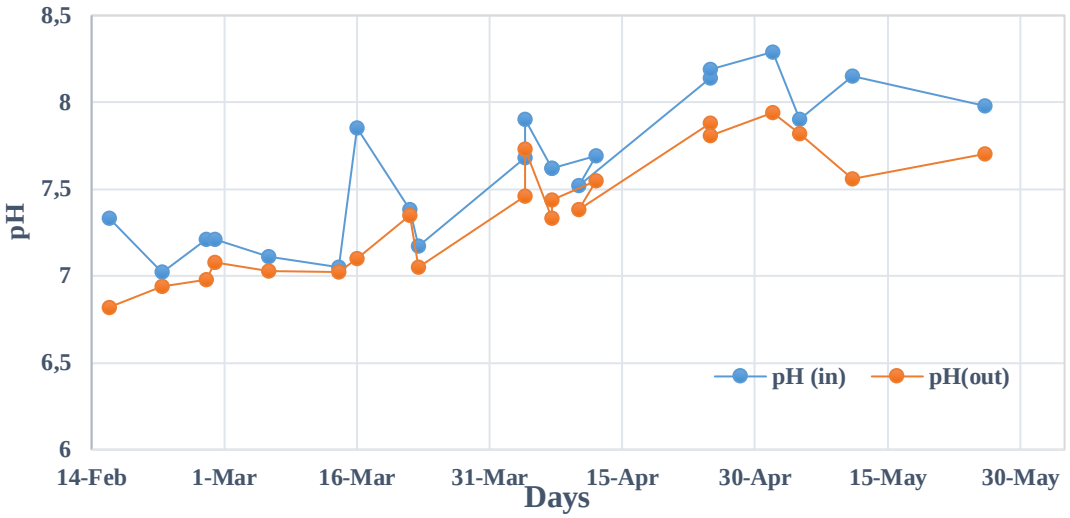


Figure 4.7 *pH variations of the UASB reactor during four months*

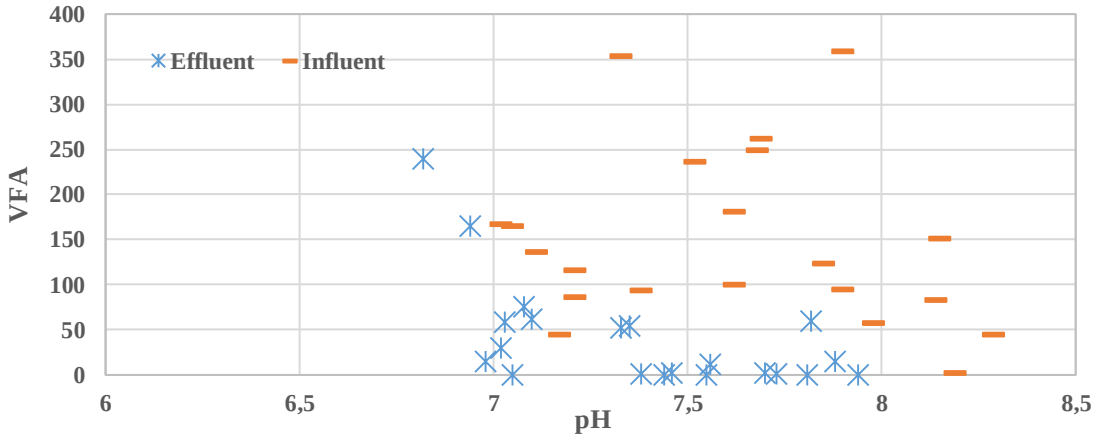


Figure 4.8 *Effect of pH on VFA*

Figure 4.9 shows VFA and alkalinity in the influent and effluent of the UASB reactor during the experimental period. The VFA and alkalinity were defined as mg acetic acid/l and mg $CaCO_3$ /l respectively. The accumulation of volatile fatty acids decreased during the experimental period. The reduction rate was high and reached 100%, from the third week of March, to zero effluent VFA. The reactor maintained high VFA removal from March until the end of May. Figure 4.9 also show alkalinity variations for influent and effluent during four

months. The concentration in the effluent was higher than influent due to consumption of acids. The reduction of VFA concentration was associated with increase in alkalinity values.

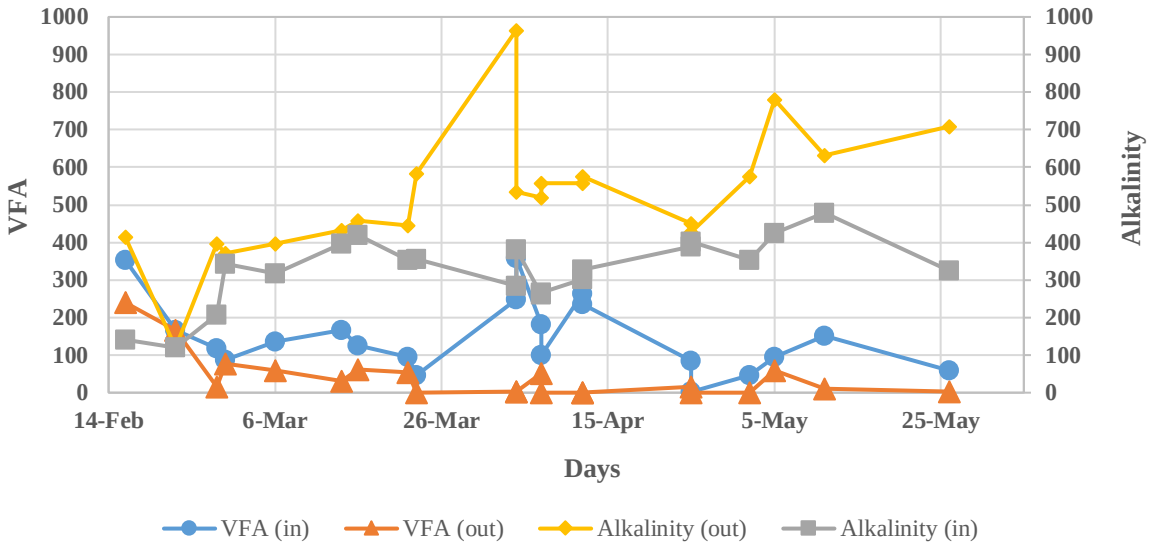


Figure 4.9 *VFA accumulation, and Alkalinity profiles in the UASB reactor*

Chapter 5

Discussion

Results obtained from experiment are discussed in this chapter. This chapter is divided into the four sub-chapters: (a) reactor acclimatization; (b) reactor performance based on COD removal efficiency; (c) operating factors including pH variability, VFA generation, and alkalinity; (d) TSS removal.

5.1 Reactor Acclimatization

The granules were added to the UASB reactor two times during the periods of experiment and there were two months' gap between them, from February till May. In the startup step of the system, the acclimatization of the granules was in the first week of experiment, and for the second time was even less than a week. The second granules were considered to be more active as they were more regular in shape and contained less suspended sludge compared to the first granules.

The food mass ratio (F/M) refers to the supply and demand balance of food for the bacteria in the bioreactor. There should be (F/M) balance for anaerobic bioreactor to operate satisfactorily [2]. The first granules seemed to have been exposed to higher food mass ratio due to a tendency of formation of flocs rather than granules. When the food mass ratio is high that means, there are a greater quantity of food relative to the quantity of microorganisms available to be consumed by bacteria. In this situation bacteria disperse in the environment, and will not form large, dense floc or granules [22]. Dispersion of the bacteria results in poor settling properties and granules deformation, with the result of losing biomass to the effluent. Therefore, careful control of F/M ratio is necessary in order to get dense flocs or granules.

As a high F/M ratio leads to enhanced metabolic activity rate and microbial growth, and high conversion rate of organic matter to methane. Excessive F/M ratio could result in poor flocs and granule formation and loss of biomass [23, 43]. On the other hand, a low F/M ratio means there is limitation for food and when the food supply is not sufficient, the bacteria begin to clump together to form dense flocs and granules that will be retained in the bioreactor and thus increasing the biomass and capacity of organic removal. The low F/M value thus increase the biomass retention, although such a so low value will lead to restricted cell growth [23, 43]. So far, any specific system there is a value of the F/M ratio which will be optimum or ideal. The acclimation of granules will be observed as reaching the steady state condition in some time, with a high COD and organic removal efficiency. This time will be system specific and affected by environmental conditions such as temperature etc. For this study the acclimation time was shortened by getting active granules from another plant, but adaption to new wastewater and environmental conditions and a relatively low temperature made the acclimation period to a few weeks.

5.2 COD Removal Efficiency

Figure 4.1 and figure 4.2 presents the variations of COD removal efficiency and organic loading rates (OLR) with respect to HRT, during the period of operation. In general, COD removal efficiency of UASB reactor increased and achieved the average of 50 % efficiency during four months. The UASB was started with active granules and showed some COD removal from the beginning. The First type of active granules had high F/M ratio, this estimation was just based on the flocs formation which were observed not dense and compact flocs. The COD removal was between 20 and 40 %, while the loading was relatively low for such systems, less than 1 kg COD/m^3. d. Otherwise the Chemical conditions in the system (pH, alkalinity, etc.) were within the requirements of anaerobic processes. The effect of OLR is mainly on the microbial ecology of granules [20]. Low OLR causes to limit the mass transfer by low up flow velocity, and also restrict the methanogenic activity of the reactor. Consequently, inert organic material accumulates at the center of granules, leading to disintegration of granules. The first granules had a "sludge-like" condition and it was uncertain about the hydraulic performance in the reactor and particularly the sludge bed. The influent was added in pulses in order to improve the hydraulic performance of the reactor. High F/M ratio of first granules and low OLR caused to have lower COD removal efficiency from start-up day till end of March. The factor that may have most affected the process performance was temperature, which was at ambient conditions.

During the first phase the temperature was between 15 ℃ and 17 °C, which would be considered as very low for anaerobic processes and thus may have affected the COD removal in that period. Due to the low COD removal and uncertainty about the characteristics of the granules, it was added new granules after about 2 months in April, which had a more regular sphere-like shape. These were expected to have higher methanogenic Activity and provide better hydraulic conditions, with an expectation of higher COD removal.

After the addition of new granules, the performance showed a gradually increasing COD removal, even the HRT was reduced from 1.1 to 0.8 days. This may, however, also contribute to the observations as higher flowrate and more regular shaped granules may have improved the hydraulic performance of the reactor and thus increased the active fraction and COD removal to above 50%. By decreasing the HRT, the contact time between wastewater and granules will decrease, while the flow distribution and contact between wastewater and biomass may have improved, hence the COD removal efficiency will enhance, as it seen in the figure 4.1. At a glance to the both figures (4.1, and 4.2) when the OLR increased the COD removal efficiency increased as well. Increasing the OLR provided more substrates concentrations in the UASB reactor that were consumed by bacteria and converted to biogas. However, for the first granules, COD removed increased with approximately constant OLR between 0.5 to 0.8 g/l. d probably due to the characteristics of first granules and not being in the steady state condition in the reactor. In addition, there was a gradual temperature increase from the initial 15-17 °C to about 22-23℃ during the experimental period, which also may have had a positive effect on the process. Increasing the temperature will enhance the growth rate, leading to promote the conversion capacity of bacteria and increasing the maximum growth rate.

Some instability in UASB reactor were observed at end of March with only 14% COD removal efficiency. This low value was associated with the clogging in the influent tube of UASB reactor due to the precipitation of solids in the elbow of the tube. Resolving of this problem was done by high pressure water. The wastewater pump also broken down in that period, causing some instability to the system.

Similar experiments for investigations of the UASB reactor have been reported by Rizvi *et.al*. (2013), presenting the COD removal efficiency of UASB reactor treating the municipal waste water was 50-70% at HRT of 3h [45]. Farajzadehha *et. al*. (2012) also reported optimum HRT for influent COD concentration of 1200 mg/l strengthen wastewater was shown to be 4 h (20 °C to 30 °C) with 70-85% of COD removal efficiency in the lab scale study [5]. Shankar Singh

et. al. (1998) was operated two UASB reactors and concluded the COD removal efficiency of 80-85% at 20 ℃ under stable condition in the treatment of municipal wastewater [7].

5.3 Operating Factors

The anaerobic treatment process is affected by the operating conditions such as VFAs, pH, and alkalinity, hence these factors were measured to evaluate the performance of the UASB reactor. As presented in figure 4.7, pH varied from 6.8 to 7.7 throughout the experimental period. In general pH of the influent were higher than the effluent. The maximum accumulation of VFA for influent and effluent were 380 and 240 mg/l acetic acid respectively. Generally, the increase in alkalinity contributed to maintain pH around neutrality and even some VFA accumulation in the UASB reactor. Based on the data from figure 4.8, the pH of the effluent did not drop because the alkalinity was sufficient to buffer the produced VFA. For the periods the VFA concentrations were zero as all produced VFA were converted to methane gas.

The UASB reactor did not need addition of buffer for pH control due to alkalinity of 520 mg $CaCO_3$/l in average. Some alkalinity is being recovered from anaerobic process by the release of ammonia NH_3, which reacts with H^+ to $NH_4^{+,}$ resulting in excess OH^- that reacts with CO_2 to form bicarbonate, HCO_3^- [46]. From April, the UASB reactor was in a relative stable condition and the COD removal efficiency was above 50% and almost zero VFAs concentration, and the TSS values indicated a high degree of sludge retention in the UASB reactor. Such low VFA also indicate a loading lower than maximum capacity, and further testing of the reactor will show the maximum capacity of the reactor.

5.4 TSS removal

Temporal variations of influent and effluent suspended solids are illustrated in figure 4.4. Higher TSS concentrations of the influent compared to effluent showed the ability of the UASB reactor to remove TSS and retain the sludge. TSS concentration in the effluent ranged from 39 to 137 mg/l from start-up day to end of March with TSS/ VSS ratio of about 1.04. The TSS concentration in the effluent decreased to about 56mg/l at the end of March. After the addition of second granules in the first week of April, the high concentration of TSS in the effluent 150 mg/l was observed at this period which reflected some washout of sludge particles, and SS concentration in the effluent was considerably decreased to 94 mg/l with increased VSS/TSS ratio of about 1.09.

Chapter 6

Conclusion

The results of present study clearly reveal the feasibility of the UASB reactor for the treatment of industrial wastewater at the tested conditions, with HRT between 18 and 26 h, under (semi)psychrophilic conditions at temperature between 15℃ and 23 ℃.

COD removal efficiencies of more than 50 % was achieved at organic loading rates up to 2 kg COD/ m^3·d. The highest recorded COD removal was above 70 %, but the maximum capacity of the pilot UASB was not seen due to time limitations of the study. The concentration of VFA in the effluent was considerably low (<50 mg/l) throughout the study indicating a stable process, and low loaded. The alkalinity of the wastewater was sufficient to balance the pH under the generation of VFA. This could, however, be more critical when the organic loading is increased on the system. The settle ability of sludge inside the reactor improved during the experimental period, as the first granules was not optimum and had some sludge with them while the second granules added about half way in the experiment showed better performance and sludge retention.

These results show that this technology is applicable for high strength wastewater in cold regions such as Norway. This study concludes that UASB reactor is a suitable alternative for the pre-treatment of SBR units at IVAR Grødaland. The UASB process merely provides advantages for the quality of final wastewater by reduction of COD concentration and conversion organic matter into profitable products as methane gas. Further studies should be conducted to evaluate the maximum capacity of the system with respect to hydraulic and organic loadings.

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Appendixes

Appendix 1: COD Analysis Data

Appendix 2: Raw Operating Data of UASB Reactor

Appendix 3: Solid Measurements Data

Appendixes

Appendix 1: COD Analysis Data

The COD measurement, OLR, HRT, and methane theoretical rate of UASB Reactor are calculated and summarized in A.1.

A.1 *Recapitulation of COD analysis data*

Date	Q_{in}(l/d)	Q_{eff}(l/d)	COD $_{in}$ (g/l)	COD_{eff} (g/l)	Total COD removal	COD Removal %	OLR (g/l.d)	Methane Theoretical (l/d)	HRT (d)	HRT (h)
21-Feb	16.8	15.6	0.572	0.469	0.103	18	0.0	0.80262	1.1	26
22-Feb	17.52	15.12	0.811	0.457	0.354	44	0.8	2.554608	1.0	25
24-Feb	16.8	16.8	1.216	0.622	0.594	49	1.1	3.49272	1.1	26
27-Feb	17.52	15.6	0.654	0.365	0.289	44	0.6	2.017428	1.0	25
28-Feb	16.8	16.8	0.868	0.452	0.416	48	0.8	2.44608	1.1	26
6-March	16.8	16.8	0.69	0.393	0.297	43	0.6	1.74636	1.1	26
14-March	16.8	14.4	0.784	0.202	0.582	74	0.7	3.59184	1.1	26
16-March	16.8	16.8	0.747	0.589	0.158	21	0.7	0.92904	1.1	26
22-March	16.8	16.8	0.817	0.551	0.266	33	0.8	1.56408	1.1	26
31-March	24	24	0.975	0.826	0.149	15	1.3	1.2516	0.8	18
31-March	24	24	0.848	0.728	0.12	14	1.1	1.008	0.8	18
4-April	24	24	1.432	0.333	1.099	77	1.9	9.2316	0.8	18
4-April	24	24	1.341	0.629	0.712	53	1.8	5.9808	0.8	18
7-April	24	24	0.935	0.566	0.369	39	1.2	3.0996	0.8	18
7-April	24	24	0.985	0.469	0.516	52	1.3	4.3344	0.8	18
10-April	24	24	0.547	0.45	0.097	18	0.7	0.8148	0.8	18
11-April	26.4	26.4	0.433	0.171	0.262	61	0.6	2.42088	0.7	16
12-April	24	24	0.422	0.187	0.235	56	0.6	1.974	0.8	18
18-April	24	24	1.065	0.435	0.63	59	1.4	5.292	0.8	18
21-April	24	24	1.045	0.411	0.634	61	1.4	5.3256	0.8	18
25-April	24	24	0.809	0.399	0.41	51	1.1	3.444	0.8	18
25-April	24	24	1	0.485	0.515	52	1.3	4.326	0.8	18

Date	Q_{in}(l/d)	Q_{eff}(l/d)	COD_{in} (g/l)	COD_{eff} (g/l)	Total COD removal	COD Removal %	OLR (g/l.d)	Methane Theoretical (l/d)	HRT (d)	HRT (h)
25-April	24	24	0.606	0.341	0.265	44	0.8	2.226	0.8	18
26-April	24	24	0.566	0.335	0.231	41	0.8	1.9404	0.8	18
28-April	24	24	0.734	0.319	0.415	57	1.0	3.486	0.8	18
2-May	24	24	0.878	0.288	0.59	67	1.2	4.956	0.8	18
5- May	24	24	0.805	0.356	0.449	56	1.1	3.7716	0.8	18
9-May	26.4	26.4	0.547	0.23	0.317	58	0.8	2.92908	0.7	16
10-May	24	24	0.818	0.474	0.344	42	1.1	2.8896	0.8	18
11-May	24	24	0.836	0.369	0.467	56	1.1	3.9228	0.8	18
11-May	24	24	1.055	0.393	0.662	63	1.4	5.5608	0.8	18
12-May	24	24	0.982	0.365	0.617	63	1.3	5.1828	0.8	18
18-May	24	24	1.148	0.279	0.869	76	1.5	7.2996	0.8	18
19- May	24	24	0.968	0.244	0.724	75	1.3	6.0816	0.8	18
23- May	24	24	0.592	0.287	0.305	52	0.8	2.562	0.8	18
26- May	24	24	0.834	0.276	0.558	67	1.1	4.6872	0.8	18
30- May	24	24	0.815	0.263	0.552	68	1.1	4.6368	0.8	18

Appendix 2: Raw Operating Data of UASB Reactor

The raw operating data of UASB reactor are summarized in Table A.2.

A.2 *Recapitulation of operating data of UASB reactor*

Date	pH_{in}	pH_{eff}	VFA_{in}	VFA_{eff}	$Alkalinity_{in}$	$Alkalinity_{eff}$	Temperature
16-Feb	7.33	6.82	353.9	239.9	140.4	414.3	21
22-Feb	7.02	6.94	167.1	165.5	119.90	120.00	19.1
27-Feb	7.21	6.98	116.9	15.40	206.90	395.90	18.4
28-Feb	7.21	7.08	86.4	76.4	343.9	372	20.3
6-March	7.11	7.03	136.2	58.6	318	396.9	18.8
14-March	7.05	7.02	165.8	29.8	397.5	433.4	19
16-March	7.85	7.1	124.4	62	419.9	456.7	18.9
22-March	7.38	7.35	94	55	353.3	446	18.4
23-March	7.17	7.05	45.6	0	355.6	584.1	23.4
4-April	7.68	7.46	249	2.5	283.4	965.1	23.3
4-April	7.9	7.73	359.1	2	379.9	533.4	21.6
7-April	7.62	7.33	181.7	52.5	260.5	518.5	22.4
7-April	7.62	7.44	100.9	0	266.5	558.7	23.1
12-April	7.69	7.55	262.7	0	301.9	556.4	20
12-April	7.52	7.38	236.7	1.5	327.7	575.1	20.8
25-April	8.14	7.88	83.5	15.3	389.5	451	20.9
25-April	8.19	7.81	2.1	0	402.5	427.8	21.4
2-May	8.29	7.94	45.2	0	352.7	575.9	24
5-May	7.9	7.82	94.7	59.6	425.7	780.2	23.5
11-May	8.15	7.56	151.1	11.7	478.6	631.8	23.2
26-May	7.98	7.7	57.9	2.5	325.9	707.9	23.7

Appendix 3: Solid Measurements Data

The total suspended solid(TSS) and volatile suspended solid(VSS) measurements data of UASB reactor are calculated and summarized in A.3.

A.3 *recapitulation of solid measurements data of UASB reactor*

Date	TSS_{in}(mg/l)	VSS_{in}(mg/l)	TSS_{eff}(mg/l)	VSS_{eff}(mg/l)	TSS Removal	VSS Removal
24-Feb	7.33	6.82	353.9	239.9	140.4	414.3
16-March	7.11	7.03	136.2	58.6	318	396.9
20-March	7.05	7.02	165.8	29.8	397.5	433.4
23-March	7.85	7.1	124.4	62	419.9	456.7
31-March	7.38	7.35	94	55	353.3	446
31-March	7.17	7.05	45.6	-*	355.6	584.1
4-April	7.68	7.46	249	2.5	283.4	965.1
7-April	7.9	7.73	359.1	2	379.9	533.4
12-April	7.62	7.33	181.7	52.5	260.5	518.5
21-April	7.62	7.44	100.9	-*	266.5	558.7
25-April	7.69	7.55	262.7	-*	301.9	556.4
12-April	7.52	7.38	236.7	1.5	327.7	575.1
2-May	8.29	7.94	45.2	-*	352.7	575.9
5-May	7.9	7.82	94.7	59.6	425.7	780.2
11-May	8.15	7.56	151.1	11.7	478.6	631.8

*- The errors were determined for VSS, due to losing of the filter weight.