APPLICATION OF UPFLOW ANAEROBIC SLUDGE BLANKET (UASB) REACTOR FOR FOOD WASTE AND POME TREATMENT WITH BIOGAS PRODUCTION

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SUPERVISOR’S DECLARATION

I hereby declare that I have checked this thesis and in my opinion this thesis is satisfactory in terms of scope and quality for the award of the degree of Masters of Engineering in Civil Engineering.

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I hereby declare that the work in this thesis is my own except for quotations and summaries which I have been duly acknowledged. The thesis has not been accepted for any degree and is not currently submitted for award of other degree.

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Specially dedicated to my lovely father “Muhammad Latif”, mother “Zubaida Begum” and my sweat wife “Noorshisyuhada Binti Sharipudin”

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ABSTRACT

The need for clean water is rising and anaerobic wastewater treatment can be used as a cost effective way out for biodegradation and energy production of organically polluted industrial waste streams. Indeed, the anaerobic digestion process can be applied to conduct various types of wastewaters in a more sustainable way than alternative processes. Applications comprise the treatment of municipal, industrial, agricultural and farming wastewaters. The upflow anaerobic sludge blanket (UASB) reactor is a system in which substrate passes first through an expanded sludge bed having a high concentration of biomass. Anaerobic treatment of liquidized food waste (LFW) was carried out in UASB reactor by stepwise increase in OLR and temperature. Thermophilic condition was achieved by increasing the temperature from 30-55 ^{o}C and pH was maintained at 7±0.5 throughout the experiment. Maximum COD removal efficiency was 93.67% (r=0.84) at an OLR of 12.5 g-COD/L.day and 4 days HRT. Maximum TOC removal efficiency was 79.14% (r=-0.94) at an influent TOC concentration of 3.59 g/L. Biogas and methane yield were recorded to a maximum of 1.364 L/g-COD$_{\text{removed}}$.day (r=0.81), 0.912 L/g-COD$_{\text{removed}}$.day (r=0.83) and average methane content of biogas was 63%. The reactor was fully acclimatized at 55^{o}C and achieved stability with high removal efficiency and biogas production. An OLR of 12.5 g-COD/L.day and HRT of 4 days were suitable for the treatment of LFW in UASBR. The treatment process can also be extendable for more than ten weeks without any measurable problem. Anaerobic bach treatment of palm oil mill effluent (POME) was carried out with cement kiln dust (CKD) as neutrilizing agent. The UASB reactor was operated continuously at 35 °C for 41 days, with varying OLR from 1.5 to 4 g-COD/L.day at HRT. The digester pH was improved steadily by increasing CKD concentrations and 23-40 g-CaO/L of CKD dosage was found suitlable to maintan pH above 7.5. The correlation (r=0.78) between CKD dosage and reactor pH showed moderate to strong relationship between CKD and reactor pH. The maximum COD removal of 95% was obtained at 4 days HRT at a highest OLR of 4 g-COD/L.day (r=0.95). The suspended solids profile was drawn along four sampling ports of reactor with increasing CKD dosage and 80% removal of suspended silids was found at 37 g-CaO/L CKD dose (r=0.88) and biogas yield was 0.90 L/g-COD$_{\text{removed}}$.day. Overall performance of CKD showed satisfactory results by replacing expensive buffer solutions thus giving a new technique of neutrilizing the acidic wastewaters.

ABSTRAK

Keperluan terhadap air yang bersih adalah meningkat pada masa kini dan rawatan untuk air sisa anaerobik boleh digunakan sebagai satu cara yang effektif dari segi kos. Ini adalah untuk pengeluaran biodegradasi dan penghasilan tenaga daripada aliran sisa industri organik yang tercemar. Malah, proses pencernaan anaerobik boleh digunakan untuk mengawal pelbagai jenis air sisa dalam cara yang lebih mampan berbanding dengan proses alternatif. Applikasinya terdiri daripada rawatan air sisa perbandaran, perindustrian, pertanian dan perladangan. Selimut aliran ke atas reaktor anaerobik enapcemar (UASB) adalah satu sistem di mana substrat akan melaluinya terlebih dahulu melalui katil enapcemar yang berkembang dan mempunyai kepekatan biomass yang tinggi. Rawatan ApplAnaerobic terhadap sisa makanan yg telah dikisarkan secara basah diletakkann dalam reaktor UASB dengan meningkatkan OLR dan suhu secara berperingkat . Kondisi Thermophilic dicapai dengan meningkatkan suhu daripada 30 °C kepada 55 °C dan nilai pH dikekalkan pada 7 ± 0.5 sepanjang percubaan dijalankan. Efisiensi bagi penyingkiran COD maksimum adalah 93,67% (r = 0,84) iaitu pada OLR 12.5 g-COD/L.day dan selama 4 hari HRT. Manakala efisiensi bagi penyingkiran TOC maksimum pula adalah 79,14% (r = -0,94) iaitu pada kepekatan influen TOC 3,59 g/L. Biogas dan hasil metana direkodkan kepada satu nilai maksimum iaitu 1,364 L/g-CODremoved.day (r = 0,81), 0,912 L/g-CODremoved.day metana (r = 0.83) dan purata kandungan biogas adalah 63%. Reaktor adalah sepenuhnya diaklimatisasikan pada suhu 55 °C dan mencapai kestabilan dengan efiensi penyingkiran yang tinggi dan pengeluaran biogas. OLR 12.5 g-COD/L.day dan HRT 4 hari adalah sesuai untuk rawatan LFW dalam UASBR. Proses rawatan juga boleh dilanjutkan selama lebih daripada sepuluh minggu tanpa sebarang masalah untuk proses pengukuran. Rawatan bach anaerobik yang digunakan untuk kilang minyak sawit efluen (POME) dijalankan dengan menggunakan debu tanur simen (CKD) sebagai ejen peneutralan. Reaktor UASB beroperasi berterusan pada suhu 35 °C untuk 41 hari, dengan pelbagai nilai OLR 1,5-4 g-COD/L.day pada HRT. PH pencerna bertambah baik secara berterusan dengan meningkatkan kepekatan CKD dan 23-40 g-CaO/L dos CKD yang ditemui dan ini sesuai untuk mengekalkan pH yang berada pada atas 7.5. Korelasi (r = 0,78) antara dos CKD dan pH reaktor menunjukkan kesederhanaan kepada ikatan yang kukuh antara CKD dan pH reaktor. Penyingkiran COD maksimum sebanyak 95% diperolehi pada 4 hari HRT pada OLR tertinggi 4 g-COD/L.day (r = 0,95). Profil pepejal terampai telah diambil sepanjang empat kawasan bagi tujuaan pensampelan reaktor dengan peningkatan dos CKD dan 80% untuk penyingkiran silids yang akan terampai yang dapat dilihat pada 37 CaO-g/L CKD dos (r = 0,88) dan hasil biogas 0,90 L/g-CODremoved. hari. Prestasi keseluruhan daripada CKD menunjukkan keputusan yang memuaskan dengan menggantikan penyelesaian yang memerlukan penampan yang mahal.Malah, ini juga merupakan teknik baru bagi peneutralan air sisa yang berasid.

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A	Cross-sectional area
AD	Anaerobic digestion
AFBR	Anaerobic fluidized bed reactor
AHMPR	Anaerobic hydrogen and methane production reactor
ASBR	Anaerobic sequencing batch reactor
ASCD	Anaerobic semi continuous digester
atm	Atmospheric
BOD	Biochemical oxygen demand
BOD_5	Biochemical oxygen demand (5 days)
CER	Certified emission reduction
CKD	Cement kiln dust
COD	Chemical oxygen demand
COD_{eff}	Effluent chemical oxygen demand
COD_{in}	Influent chemical oxygen demand
CPO	Crude palm oil
CSTR	Continuous stirrer tank reactor
EFB	Empty fruit bunch
EIA	Environmental impact assessment
F/M	Food to micro-organism ratio
FAFBR	Flocculant-anaerobic fluidized bed bioreactor
FFB	Fresh fruit bunch
FFR	Fix film reactor

FSAD	Full scale anaerobic digester
Gg	Giga gram (10^9)
GHG	Greenhouse gas
GSL	Gas-solid-liquid
HRT	Hydraulic retention time
IBR	Integrative biological reactor
KSLH	Kilang Sawit Lepar Hiller
LFW	Liquidized food waste
MFBR	Mesh filter bioreactor
MSW	Municipal solid waste
Mwh	Mega watt hour
OLR	Organic loading rate
POME	Palm oil mill effluent
PPF	Palm pressed fiber
PVC	Polyvinyl chloride
Q	Flow rate
RE	Renewable energy
SAMR	Submerged anaerobic membrane reactor
SCOD	Soluble chemical oxygen demand
SCR	Semi continuous reactor
SD	Standard deviation
SE	Standard error
SMAD	Self mixing anaerobic reactor

SRT	Sludge retention time
SS	Suspended solids
TCOD	Total chemical oxygen demand
Tg	Tera gram (10^{12})
TKN	Total Kjeldahl nitrogen
TN	Total nitrogen
TOC	Total organic carbon
TP	Total phosphorous
TS	Total solids
TSS	Total suspended solids
UASB	Upflow anaerobic sludge blanket
UASSR	Upflow anaerobic solid state reactor
VFA	Volatile fatty acid
VLR	Volumetric loading rate
VS	Volatile solids
VSS	Volatile suspended solids
$^{\circ}$C	Degree Celsius
Km	Kilometer
MPa	Mega Pascal
min	Minute
cm	Centimeter
cm^2	Square centimeter
d	Diameter

g	Gram
H	Height
h	Hour
K	Kelvin
kg	Kilogram
L	Liter
m	Meter
M	Morality
m^3	Cubic meter
mm	Millimeter
S	Siemens
V_{up}	Upflow velocity
V_r	Working volume
V_w	Total volume
µ	Micro (10^{-6})
$ZnCl_2$	Zinc chloride
SO_3	Sulphure Trioxide
SiO_2	Silicon Dioxide
P	Phosphorous
PO_4^{3-}	Phosphate
N	Normality
N_2	Nitrogen
NaOH	Sodium Hydroxide

NH_4Cl	Ammonium chloride
NO	Nitrogen oxide
$MgCl_2$	Magnesium chloride
MgO	Magnesium oxide
H_2S	Hydrogen sulfide
H_2SO_4	Sulfuric acid
H_3BO_3	Boric acid
K_2O	Potassium oxide
K_2PO_4	Potassium phosphate
KCl	Potassium chloride
KOH	Potassium hydroxide
Fe_2O_3	Ferric oxide
$FeCl_2.6H_2O$	Ferric chloride hexahydrate
CO	Carbon monoxide
CO_2	Carbon dioxide
$CuCl_2.2H_2O$	Copper chloride dihydrate
Al_2O_3	Aluminum oxide
$Ca(OH)_2$	Calcium hydroxide
CaO	Calcium oxide
CH_4	Methane
$(NH_4)_2SO_4$	Ammonium sulphate

CHAPTER 1

INTRODUCTION

1.1 INTRODUCTION

This chapter describes the background of food waste and palm oil mill effluent (POME). The first part of chapter explains about food waste; food waste generation in municipal solid waste (MSW) and current waste handling techniques. Meanwhile the second part explains a brief background of POME; its annual production and treatment methods. Finally for the last part of this chapter, problem statement, objectives and the scope of study were described.

1.2 THE FOOD WASTE

The demand for the reduction of food waste in Malaysia has been increased during recent years. The rapid development of economic growth is throwing impact on the work-of-art of the solid waste. In Malaysia, the municipal solid waste comprises of 62% of food waste (Hanssan et al., 2001). Because of growing population and increasing consumption, the amount of solid waste generated in Peninsular Malaysia was 17,000 ton/d in 2002 and 19,100 ton/d in 2005. Generation of solid waste is expected to reach 30,000 tons/d in 2020 (Nadzri, 2008, and Ngoc and Schnitzer 2009).

The present of methane concentration in atmosphere is due to increasing organic decay, particularly, as methane has proven effective at trapping infrared radiation and tends to persist long in the atmosphere (Black, 1993; Calander, 1995; Kamarudin, 2008). Methane gas that traps heat in the atmosphere is one of the most important GHGs such as CH_4, CO_2, NO, NO_2, NO_3 and Ozone because its global warming potential is 21 times higher than carbon dioxide (Ishigaki et al., 2005). Currently, landfills are the major source of methane gas emissions (53%) followed by palm oil mill effluent (38%), swine manure (6%), and industrial effluent (3%) in Malaysia (MEWC, 2004). The expected methane emissions from landfills are approximately 180 Gg in 2010 (Abushammala et al., 2010) in which 70% emissions are from anthropogenic sources and 30% from natural sources (El-Fadel and Masoud, 2001). Moreover, anaerobic digestion of organic waste in landfills releases the gases methane and carbon dioxide that escape into the atmosphere and pollute the environment (Zhu et al., 2009). The anaerobic decomposition of food waste in landfills produces gas, containing approximately 50–60% methane and 30–40% carbon dioxide, respectively (Abushammala et al., 2009).

Recently, the organic fraction of solid waste has been recognized as a valuable resource that can be converted into useful products via microbially mediated transformations such as hydrolysis, acidogenesis, acetogenesis and methanogenesis (Yu and Huang, 2009, and Lesteur et al., 2010). There are various methods available for the treatment of organic waste however anaerobic digestion appears to be a promising approach (Lee et al., 2009). Anaerobic digestion involves a series of metabolic reactions such as hydrolysis, acidogensis and methanogenesis (Themelis and Ulloa, 2007).

1.3 THE PALM OIL MILL EFFLUENT

The production of palm oil is increasing every year due to its application for biodiesel production. This leads to the increasing amount of palm oil mill effluent (POME); a by-products from the oil-palm extracting process (Poh and Chong, 2009). Malaysia is known as the world’s largest palm oil producer, with more than 40 million tons of POME produced every year (Yacob et al., 2006). In Malaysia, 20 million tons of palm oil was produced in year 2008 (Wu et al., 2010). For every ton of palm oil produced, 2.5 tons of POME is generated (Ahmad et al., 2006).

POME is a viscous brown liquid with fine suspended solids and the pH ranges is in between 3.5 and 5 with high content of COD (16–100 g/L) (Najafpour et al., 2006; Zhang et al., 2008 and Wu et al., 2009). Meanwhile the de-oiled POME is a thin brown liquid, obtained from a clarification process of POME where the floating fats and the settling organic particles are removed. At this clarification process, while most of particles and floating fats are removed, some parts of the organic matter in POME are hydrolyzed/fermented which resulting in volatile fatty acid (VFA) production as well as lower pH of effluent (Ahmad et al., 2006). The pH is one of the key factor that influence anaerobic digestion of POME because methane producing bacteria require a neutral to slightly alkaline environment in order to produce maximum methane from POME. Optimum pH for most of the microbial growth is between 6.8 and 7.2, while pH lower than 4 and higher than 9.5 is not tolerable (Gerardi, 2006; Edgar et al., 2006). Thong et al. (2010) reported methane potential of POME around 45 m^3 methane/m^3 of POME, corresponding to a biogas energy potential of 18 GWh per year for utilization of the Malaysian POME.

1.4 TREATMENT TECHNIQUES OF FOOD WASTE AND POME

Currently, various methods for reutilization and disposal of the food waste are available, which include landfills, incineration, aerobic composting and anaerobic digestion but due to environmental regulations some disposal methods are going to be prohibited and less desired (Oh et al., 2008). With capacity of landfill gradually filling up and fewer landfills being commission, it is critical to look out for alternative disposal methods. For instance, composting is an aerobic process and it produces humus that can be used as fertilizers or soil conditional (Cheng et al., 2007). Meanwhile, incineration is yet another approach however it is not feasible because the moisture contents of food waste are very high (Komemoto et al., 2009). The high moisture content in food waste is feasible for composting and anaerobic degradation (Hernandez-Berriel et al., 2008). Recently, a variety of new bioreactor designs have been developed, which facilitate a significantly higher rate of reaction for the treatment of food waste (Bouallagui et al., 2003; Mumme et al., 2010, and Xing et al., 2010).

It was reported that most of the palm oil mills employed anaerobic digestion as their primary treatment of POME (Tay, 1991). Malaysia for instances, more than 85% of its palm oil mills have adopted the ponding system for POME treatment (Ma et al., 1993) and the rest opted for open digesting tank (Yacob et al., 2005). However, treatment of POME in ponds/open tanks requires long retention time and large treatment areas, causing odour and environmental load due to gas emissions and leachate contamination to groundwater (Chin et al, 1996, and). In order to shorten the retention time as well as to reduce the treatment area, and to capture the biogas for energy utilization, the high-rate anaerobic digestion has already gained greatly attention (Fang et al., 2011). For example Borja and Banks (1994, 1994a, 1995) reported the application of an upflow anaerobic sludge blanket reactor, upflow anaerobic filtration and fluidized bed reactor. Najafpour et al. (2006) reported the application of upflow anaerobic sludge fix film reactor. Anaerobic contact digester (Ibrahim et al., 1984) and continuous stirred tank reactor have also been studied for treatment of POME (Chin, 1981). Other than anaerobic digestion, POME has also been treated using membrane technology (Ahmad et al., 2006a, 2007), aerobic activated sludge reactor (Vijayaraghavan et al., 2007), and evaporation method (Ma et al., 1997).

1.5 PROBLEM STATEMENT

Due to increasing demand for renewable energy and diversion of organic residuals from landfills to reduce greenhouse gas emissions among other environmental impacts, treatment of food waste using anaerobic digestion technologies has become a more attractive method as this method is considered as the best method capable to produce renewable energy (Chen et al., 2010). In Malaysia, the renewable energy potential from food waste has been reported to be 248 MW (Oh et al., 2010). Thus, the disposal of biodegradable materials is of particular importance in any waste management scheme. Besides composting, anaerobic digestion and bio-methanation of organic solid waste are also considered as promising treatment options for this particular waste fraction (de Baere, 2006; Jingura and Matengaifa, 2009, and Khalid et al., 2011). During anaerobic digestion, a bioreactor design has been reported to exert a strong influence on the performance of a digester (William and David, 1999). According to Ward and his co-workers, an anaerobic bioreactor should be designed in a way that allows a continuously high and sustainable organic load rate with a short hydraulic retention time and has the ability to produce methane at the maximum level (Ward et al., 2008).

During anaerobic treatment of POME, pH exhibits a prominent role followed by reduction in volatile fatty acid for further exchange to biogas. Volatile fatty acids degradation depends on the buffering capacity of substrate utilizing the biomass. The neutralization of wastewater during anaerobic treatment is still a challenge for researchers (Patel and Madamwar, 2002, and Parawira et al., 2006). Many buffering solutions like sodium hydroxide, bicarbonates and reagents of similar buffering capacity have extensively been used for wastewater treatment (Yu and Fang, 2003, and Borja et al. 2005) and some of them are quite expensive which are suitable for pilot or large scale treatment plants. The CKD is a byproduct and environment friendly waste from cement industry (PCA, 2007). Moreover, CKD is an easily available waste from cement industry and has much attraction for the anaerobic treatment of POME.

1.6 RESEARCH OBJECTIVES

i. To study the effect of temperature and organic loading rate parameters on upflow anaerobic sludge blanket reactor (UASBR) during liquidized food waste treatment.
ii. To study the feasibility of CKD as a potential neutralizing agent of a POME treatment on upflow anaerobic sludge blanket reactor (UASBR).
iii. To measure the biogas production from liquidized food waste and POME using upflow anaerobic sludge blanket reactor (UASBR).

1.7 SCOPE OF STUDY

Anaerobic treatment of liquidized food waste (LFW) was carried out in an UASB reactor by stepwise increase in organic loading rate (OLR) and temperature. Thermophilic condition was achieved by stepwise increase in the temperature from 30-37 °C, 37-50 °C and 50-55 °C and pH was maintained at 7±0.5 throughout the experiment. The applied OLR range was 2, 3, 4, 5, 6.25, 8.33, 10 and 12.5 g.COD/L.day. The chemical oxygen demand (COD), total organic carbon (TOC), biogas and methane production were measured at hydraulic retention time (HRT) of 10 to 4 days, and start up strategy of the reactor was monitored for 10 weeks.

Anaerobic degradation of palm oil mill effluent (POME) was carried out using cement kiln dust (CKD) as a neutrilising agent. Upfow anaerobic sludge blanket (UASB) reactor was operated continuously at 35 °C for six weeks, with varying OLRs (1.5, 2.5 and 4 g-COD/L.day) at a fix hydraulic retention time (HRT). The digester pH was improved steadily by increasing CKD concentrations and CKD dose of 20-40 g-CaO/L was found suitlable to maintan pH avobe 7.5. The results revealed good performance of CKD in terms of pH control, COD removal, effluent volatile fattly acids and alklinity, suspended solids concentration and removal. A correlation (r = 0.78) showed moderate to strong realtionship between CKD dose and reactor pH. The volatile farry acids (VFA), alkalinity, COD removal, SS removal and biogas production were measure during the experiment. A flow rate (Q) of 0.675 L/d and upflow velocity (V_{up}) of 0.2 m/h was maintained through the study while VFA/alkalinity ratio was remained below 0.40 which supported successful anaerobic degradation of POME.

1.8 OVERVIEW OF THE THESIS

This thesis consists of six chapters. A brief introduction about the food waste and POME, wastes generation and treatment is given in Chapter 1 (Introduction). This chapter also includes problem statements that provide some basis and rationale to identify the research directions to be followed in this study. Then, the objectives of the present study are elaborated in detail together with the scopes of the study to be covered. The organization of the contents of this thesis is also given in the last section of this chapter.

Chapter 2 (Literature Review) discusses about food waste, POME and CKD. The technical aspects of anaerobic digestion process, a comparison of aerobic and anaerobic treatment process, current treatment methods of food waste and POME that are related to the present study. Comparison of different bioreactors with UASB reactor is also included in this chapter. A review of UASB treatment potential with its key design parameters and performance has also given at the end of this chapter.

Chapter 3 (Materials and Methods) presents the detail of the materials and chemicals used in the present study. Detail of the experimental set-up is then elaborated in this chapter. This followed by the detail experimental procedures, which includes the mathematical equations used for anaerobic treatmement of food waste and POME, batch experiments for CKD slacking process and analytical techniques.

Chapter 4 (Results and discussion) which is the main part of this thesis is outlined by two main studies. In first section, the effect of temperature and OLR on the anaerobic treatment of food waste along with COD, TOC removal and biogas production has been presented . Then, study of POME digestion by using CKD as a neutralizing agent in the UASB reactor is discussed in the second section. In the following section, performance of CKD with POME is investigated.

Chapter 5 (Conclusions and recommendations) concludes the findings from the current studies and recommendations for future studies in the related field made from the understanding and information generated in the present study. These recommendations are given due to their significance and importance to be further investigated and explored by future research work in this area.

CHAPTER 2

LITERATURE REVIEW

2.1 INTRODUCTION

This chapter provides a brief review of organic fraction of MSW, POME, CKD and anaerobic treatment. It is also review on the general concepts of anaerobic wastewater treatment processes, the leading mechanism of anaerobic digestion, various anaerobic treatment processes as well as factors affecting anaerobic process. Medium and high rate anaerobic reactors have also been discussed in this chapter along with their characteristics. Finally this chapter also focuses on the upflow anaerobic sludge blanket (UASB) reactor treatment potential with its advantages and disadvantages along with the key design, its performance and operational parameters being used by previous researchers.

2.2 PROPERTIES OF FOOD WASTE

According to Ninth Malaysia Plan 2006-2010, the Malaysian food was the major constituents of MSW with the range of 45-49% while larger ranges (60-62%) has been reported by Hanssan et al. (2001) and Kathirvale et al. (2003). Figure 2.1 shows MSW components in Kuala Lampur from which we can justify the amount of food waste could be suitable for anaerobic treatment and energy production.

The physico-chemical properties of mixed food waste are shown in Table 2.1.

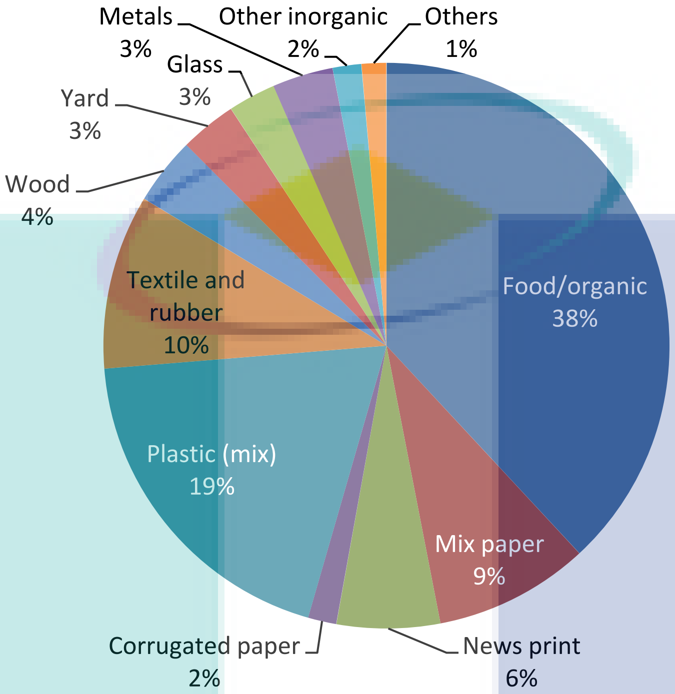

Figure 2.1: Average composition in weight percentages of MSW components generated by various sources in Kuala Lumpur (Adapted from Kathirvale et al., 2003)

Table 2.1: Physico-chemical properties of mixed food waste

Parameter*	Kubaská et al. (2010)	Kuo and Lai (2010)
pH	-	-
BOD	-	-
COD	252	379
VSS	142	154 as VS
TSS	155	170 as TS
TKN	-	3.2
TP	-	1.5

*All values are in g/L except pH

2.2.1 Food Waste Management Techniques

Landfilling

Landfilling of solid waste represents the most widespread method of solid waste disposal in the world. Currently, in Malaysia it has been reported that landfills are the major source of methane gas emissions (53%) followed by POME (38%), swine manure (6%) and industrial effluent (3%) (MEWC, 2004, and Kamarudin, 2008). Landfill gas contains roughly 50-60% CH_4 and 40-50% CO_2 and the decomposition of each metric ton of solid waste could potentially release 50-110 m^3 of carbon dioxide and 9-140 m^3 of CH_4 into the atmosphere (Ayalon et al., 2001, and Abushammala et al., 2009). The CO_2 released, although a greenhouse gas, does not have a net effect on global warming since the carbon in CO_2 is fixed by photosynthetic plants and returned to the carbon cycle (Canadell et al., 2007). It is believed that 18-20% of the global warming effect is due to CH_4 emission which traps 20-25 times more heat than CO_2 (Francese et al., 2000). Methane emissions from landfills are expected to reach 180 Gg in 2010 (Abushammala, 2010) in which 70% emissions are from anthropogenic sources and 30% from natural sources (El-Fadel and Masoud, 2001).

Incineration

Waste incineration, like any other combustion processes, releases CO, NO_x (NO, NO_2, NO_3) and volatile organic compounds, which contribute to environmental pollution (Ayalon et al., 2001). Generally, municipal waste is incinerated in order to reduce waste volume. Therefore it may reduce landfill costs and it may be economically beneficial due to the energy recovery; either for heating or electricity generation. In general, incineration suitable for residues containing less than 50% water otherwise oil or gas must be added to fuel the combustion process (Chynoweth and Legrand, 1988). Due to the environmental hazards create by incineration and landfilling, it makes both methods are environmentally unsuitable. Thus, anaerobic/anaerobic techniques are the most preferable methods among solid waste management.

Aerobic and Anaerobic Treatment

Aerobic Treatment/Composting

An aerobic treatment or composting involves the use of oxygen as an electron acceptor by microorganisms during the degradation of organic matter into CO_2, water, nitrates and sulfates. Of all biological waste treatment methods, aerobic treatment is the most widespread process used throughout the world (Malakahmad et al., 2009). The compost contains nutrients and it is used as a soil conditioner in agriculture. Composting, although good at stabilizing organic solid waste, it can be only applied to a structured solid with water contents between 50% and 60% (Insam and de Bertoldi, 2007).

Composting is an exothermic biodegradation process that involves a complex web of biochemical reactions in which facultative and aerobic micro-organisms catabolism substrates for their growth and metabolic needs (Paul and Clark, 1996; Alexander, 1997). A part of the carbonaceous substrate first hydrolyzes, and is then oxidized to produce carbon dioxide (CO_2), water and heat, and the rest transforms into humic substance. The heat released causes the temperature to rise and accelerates the degradation process up to a point, reflecting an increase in microbial growth rates, enzymatic production, and the reaction rate of the enzymes production (Richard, 1997). Temperatures over 65 °C and excessive heat generation can kill some of the microorganisms responsible for the degradation of the organic compounds (Chang et al., 2005).

Anaerobic Treatment

Anaerobic digestion with energy recovery is an attractive method for the treatment of solid waste and wastewater. In general, anaerobic digestion is a complex biochemical process that is carried out in a number of steps by several types of microorganisms in the absence of oxygen. Methane and carbon dioxide are the principal end products of the process along with minor quantities of nitrogen, hydrogen, ammonia and hydrogen sulphide (Garcia et al., 2000). Due to high organic and moisture contents, biodegradation using anaerobes is another potential treatment method. Anaerobic treatment presents the potential of producing biogas which can be used for cooking, heating and electricity generation. There are three distinct stages in an anaerobic process (Paola et al., 2007). Firstly, the hydrolysis of long chain hydrocarbon into smaller chain hydrocarbon. The second phase is the conversion of smaller chain hydrocarbon organic matter to acetic acid, fatty acid and hydrogen by acetogenic bacteria. At this stage, the pH will drop due formation of acid. The final stage of anaerobic process is the methanogenesis i.e. conversion of acetic acid into methane and carbon dioxide. This process is pH sensitive and temperature sensitive. The temperature should ranges from 30 to 60 degree C (Bernd, 2007). It is ideal process for tropical climate due to high daily temperature. The biodegradation of food waste also produces leachate. Factors that influence the anaerobic process are pH, temperature and nutrients. Methane producing bacteria require a neutral to slightly alkaline environment (pH 6.8 to 8.5) in order to produce methane. Acid forming bacteria grow much faster than methane forming bacteria. If acid-producing bacteria grow too fast, they may produce more acid than the methane forming bacteria can consume. Then, excess acid will build up in the system. When the pH drops, and the system may become unbalanced, inhibiting the activity of methane forming bacteria. Methane production may stop entirely. The major nutrients required by the bacteria are ammonium and phosphate (Cheng et al., 2007). There are normally plenty of proteins present from the decomposition of household waste so there is not usually a shortage of ammonium ions, but some additional phosphate may be required to ensure optimum process rates. The micro-nutrients which methane bacteria require in trace quantities include: iron, manganese, copper, nickel, zinc and cobalt, all of which are likely to be present in municipal solid waste. The production of biogas is dependent on the optimum biodegradation process. Generally, the composition of biogas is 40 – 60% methane, 50-40% carbon dioxide, and other gases such as hydrogen sulfide, ammonia and other tracer gases (Chua et al., 2008).

2.3 PALM OIL MILL EFFLUENT

Oil palm [*Elaeis guineensis*] is one of the most versatile crops in tropical countries (Zinatizadeh et al., 2007). In Malaysia, palm oil industry is one of the most important contributors to boost Malaysia economy. The total oil palm planted area in the country has increased by 4.3% to 4.48 million hectares, and fresh fruit bunch production has reached 90.5 million tons in 2008 (MPOB, 2009). In year 2009, 43.8 million m^3 (11,600 million gallon) of POME was generated from Malaysian palm oil mills base on the production of 17.56 million tons of total crude palm oil (MPOB, 2010). In fact, the palm oil industry was identified as one of the agricultural industry in Malaysia that generates the highest pollution load into rivers throughout the country. Moreover, POME is a non-toxic liquid waste with unpleasant smell; its COD and BOD value are high enough to cause serious pollution and environmental problem to the rivers (Wu et al., 2007). Nevertheless, many palm oil mills are still unable to adhere to the wastewater discharge limits imposed by authority and thus resulting to a dramatic increase in the number of polluted rivers (Ahmad and Chan, 2009).

During palm oil extraction, generation of palm oil mill effluent (POME) ranging from 0.6-1.5 ton per ton of fresh fruit bunch (FFB) processed by the oil palm mill (Ahmad et al., 2003; Wahid et al., 2006). Although palm oil extraction process has been recognized for its contribution towards economic growth, its rapid development has also contributed to environmental pollution correspondingly to the large quantity of waste products produced. For instance, POME generation was ranging in between 0.6-1.5 ton per ton of FFB produce by oil palm mill during palm oil extraction (Ahmad et al., 2003). Other than POME, the waste products also consist of fibrous material such as empty fruit bunches (EFB), palm pressed fibers (PPF) palm kernel shell as well as less fibrous material such as palm kernel cake (Cheah, 2003).

In general, POME is a thick brownish liquid which consists of 95-96% water, 0.6-0.7% oil and 4-5% total solids including 2-4% suspended solids originating from the mixture of a sterilized condensate, separator sludge and hydro-cyclone wastewater (Ahmad et al., 2003). The average values of biochemical oxygen demand (BOD) and chemical oxygen demand (COD) reported was 26.3 and 55.9 g/L respectively at pH ranges from 4-5 (Najafpour et al., 2006). The physico-chemical properties of POME reported by previous researchers is are given in Table 2.2.

Table 2.2: Physico-chemical characteristics (POME)

Parameter*	Ahmad et al. (2003)	Najafpour et al. (2006)
pH	4.7	3.8-4.4
BOD	25.0	23.0-26.0
COD	50.0	42.5-55.9
TS	40.5	-
SS	18.0[a]	16.5-19.5[a]
TKN	0.750	0.50-0.70[b]
Oil and Grease	4.0	4.9-5.7

* All parameters are in g/L except pH
[a] as TSS
[b] as total nitrogen

The physico-chemical properties of POME vary widely throughout the year because of mill operations and seasonal cropping (Yacob et al., 2005).

It was reported that the atmospheric methane concentrations has incredibly increased by 30% in the last 25 years (IPCC, 2006). It is believed that POME treatment process has contributed to the atmospheric methane concentration increment. Henson (2009) reported a net carbon emission from POME is approximately 1.4×10^6 tons per year while Yacob et al. (2005) reported 5.5 kg of CH_4 (or approximately 36% of biogas) is emitted from open digesting tanks. Assuming a mean annual increase of 29% as experienced from 1990 to 2004 and the estimated CH_4 gas emission may be 0.502×10^6 tons in the year 2020 (Wahid et al., 2005). Certified Emission Reduction (CER) can be obtained by using methane gas as a renewable energy (Tong and Jafar, 2006). To date, 85% of POME treatment in Malaysia is based on an anaerobic and facultative ponding system, which is followed by an open-tank digester coupled with extended aeration (Siang, 2006). At high organic loadings, the insoluble organic degradation of the effluent tends to accumulate within the granules or sludge density region of the ponds, leading to granule destabilization or inhibition of granule formation (Vijayaraghavan et al., 2007).

2.3.1 Environmental Regulations of POME Discharge

The environmental quality regulations for the crude palm oil industry were the first set of regulations promulgated under the Environmental Quality Act (EQA) 1977 in order to control industrial pollution sources (Thani et al., 1999). Specifically, all the activities including effluent discharge is regulated under Environmental Quality prescribed (prescribed premises crude palm oil) Regulations 1977. The effluent discharge standards ordinarily applicable to crude palm oil mills are presented in Table 2.3.

Table 2.3: The parameter limits for effluent discharge

Parameter*	**EIA Standards (2007)[a]**
pH	6-9
COD	100
BOD	50
TS	-
SS	0.10
Total Nitrogen	-
Oil and grease	0.01

* All parameters are in g/L except pH

[a] EIA is under guidelines of DOE Malaysia.

2.3.2 Renewable Energy from Oil Palm Industry in Malaysia

Due to increasing demand for energy, cost saving and the protection of the environment, anaerobic digestion technology has become a worldwide focus of research. Generally, Malaysia’s energy sources primarily comprise of oil, natural gas, hydropower and coal, however renewable energy (RE) sources such as solar power and biomass are now being exploited. It was reported that renewable energy potential from oil palm mill waste was 400 MW in the year 2010 (Oh et al., 2010). Figure 2.2 shows different types of biomass generated by an oil palm mill. For example EFB has the energy potential of 8.3 x 10^6 Mwh, energy potential from shell is 1.4 x 10^6 Mwh along with the POME energy potential of 10^4 Mwh.

Empty fruit bunch (EFB)
Energy potential 8.3x10^6 Mwh

Fiber
Energy potential 2.5x10^6 Mwh

Shell
Energy potential 1.4x10^6 Mwh

Palm oil mill effluent (POME)
Energy potential 10^4 Mwh

Figure 2.2: Palm oil wastes as renewable energy sources (Adapted from Hitam, 1999)

In Malaysia, the most extensive study on the use of biomass to produce energy has been on oil palm wastes (Energy Commission, 2002). As Malaysia has the potential to be one of the major contributors of renewable energy in palm oil biomass thus Malaysia may become a role model to other countries in the world that has huge biomass feedstock (Ong et al., 2011). Therefore, renewable energy has been identified by Malaysian government as the 5th fuel under ‘The New Five-Fuel Diversification Strategy’ (Kannan et al., 2005). From the four biomass sources, three of them (EFB, fiber and shell) can be directly burned as fuel while POME must first be anaerobically converted into biogas. Therefore, a high rate anaerobic bioreactor is essential to be applied as it can serves dual-function such as wastewater treatment as well as energy generation (organic conversion into biogas).

2.3.3 POME Treatment Systems

Palm oil industries are facing tremendous challenges to meet the increasingly stringent environmental regulations. Over the past decades, several cost-effective treatment technologies comprising anaerobic, aerobic and facultative processes have been developed for the treatment of POME. It has been reported that more than 85 % of palm oil mills use solely ponding systems due to their low costs and only a few mills are equipped with biogas recovery systems (Yeoh, 2004).

Recently, an application of an efficient, stable and economic of a high rate anaerobic treatment system is seriously being investigated by many researchers in Malaysia (Yu et al., 2004). This is due several drawbacks of conventional POME treatment methods. Such as long HRT, low treatment efficiency, high sludge production, extensive land area and emission of large amount of GHG (CO_2 and CH_4). Figure 2.3 shows a POME ponding treatment system at a palm oil mill in Kuantan, Pahang, which is a common practice in most palm oil mills in Malaysia.

Oil removal from POME

Continued oil removal from POME

Anaerobic pond

Aerobic lagoons

Figure 2.3: Working POME ponding system in oil palm mill

2.3.4 Gaseous Secretion from POME and Food Waste Treatment

There are two main sources of air pollution in the mills which is the boiler or through incinerator either which are caused by incomplete combustion of the solid waste materials; waste fiber, shell materials and EFB (Mannan and Ganapathy, 2004). Meanwhile, since POME contains high level of organic matters, adoption of anaerobic digestion in the first stage of the treatment process is important to convert a bulk of wastes into biogas before it is further subjected to an aerobic treatment in order to meet the required discharge standards. In Malaysia, the current practice of treating POME by either using ponding system and/or open digesting tank systems (Ma et al., 1999) has been reported in producing GHG (CH_4 and CO_2) which is then emitted to the atmosphere (Yacob et al., 2005). Methane gas is one of the most crucial GHGs among other gases because its global warming potential is 21 times more than carbon dioxide (Ishigaki, 2005). A life cycle assessment study on Malaysian palm oil mill processes revealed that the non-recovered biomethane emission from POME contributed the highest impact towards the environment (climate change category) and therefore makes the overall processes not environmentally friendly (Subramaniam et al., 2008).

Similarly, food waste contains large quantity of organic matter (Rao and Singh, 2004) and it is a precious biomass source for anaerobic digestion and biogas production (Han et al., 2005). Bio-technologies like landfills and composting are presently the most popular techniques for their disposal, receiving roughly 60-89% of domestic waste in different countries (Toms et al., 1995). Biogenic production of methane from landfills and composting are well understood (Wang et al., 1997; Akesson and Nilsson, 1998) and estimated to account for 4-15% of the total global budget (Tsujimoto et al., 1994). Most laboratory studies have measured the reduction of the organic fraction during composting (Inoko et al., 1979; Baca et al., 1992; Wang et al., 2003), while some studies have included a setup to capture and measure gases produced during composting.

2.4 ANAEROBIC DIGESTION

Anaerobic digestion is applicable for a wide range of material including municipal, agricultural and industrial wastes, and plant residues (Chen et al., 2008). This process has some advantages over aerobic process due to a low energy requirement for operation and a low biomass production (Kim et al., 2006), as well as simultaneous production of a renewable energy (Jingura and Matengaifa, 2009). Furthermore, the anaerobic digestion of organic waste is also an environmentally useful technology. It is capable in reducing the environmental pollution in two main ways; (i) the sealed environment during the process will prevents methane release into the atmosphere, (ii) by burning the methane, it will release carbon–neutral carbon dioxide (no net effect on atmospheric carbon dioxide and other greenhouse gases) (Ward et al., 2008). The comparison between anaerobic treatments is listed in Table 2.5.

Table 2.4: Comparison of aerobic and anaerobic biological waste (water) treatment

	Aerobic digestion	**Anaerobic digestion**
Start-up	• Short start-up period	• Long start-up period
Process	• Integrated nitrogen and phosphorus removal possible. • Production of high excess sludge quantities. • Large reactor volume necessary. • High nutrient requirements.	• No significant nitrogen or phosphorus removal, nutrients removal done via post treatment. • Production of very little excess sludge (5-20%). • Small reactor volume can be used. • Low nutrient requirements.
Carbon balance	• 50-60% incorporated into CO_2; 40-50% incorporated into biomass.	• 95% converted to biogas; 5% incorporated into microbial biomass.
Energy balance	• 60% of available energy is used in new biomass; 40% lost as process heat.	• 90% retained as CH_4, 3-5% is lost as heat, and 5-7% is used in new biomass formation.
Residuals	• Excess sludge production. • No need for post-treatment.	• Biogas, nitrogen mineralized to ammonia. • Post-treatment required for removal of remaining organic matter and malodorous compounds.
Costs	• Low investment costs. • High operating costs for aeration, additional nutrient and sludge removal, and maintenance.	• Often moderate investment costs. • Low operating costs due to low power consumption and additional nutrients hardly required.
State of development	• Established technology.	• Still under development for specific applications.

(Adapted from Lettinga *et al*., 1984; Lepisto and Rintala, 1997; Banerjee *et al*., 1999; Zoutberg and Eker, 1999; Gijzen 2001 and Lettinga, 2001).

With the introduction of both commercial and pilot anaerobic digestion plant designs during early 1990s, anaerobic digestion of organic waste has received worldwide attention (Karagiannidis and Perkoulidis, 2009). It is a process by which almost any organic waste can be biologically transformed into another form in the absence of oxygen. The diverse microbial populations degrade organic waste, which results in the production of biogas and other energy-rich organic compounds as end products (Lastella et al., and 2002. Lata et al., 2002).

At present, anaerobic digestion of organic matter and wastewater has meticulous attraction due to the economic advantages of energy production (Yu et al., 2002; Edgar et al. 2006; Bohn et al., 2007, and Sankaran et al., 2010). Anaerobic treatment of wastewater has been well thought-out to have number of advantages over conventional aerobic treatment processes.

Generally, there are four (4) basic steps which involve in an anaerobic digestion of organic matter namely hydrolysis, fermentation (acidogenesis and acetogenesis) and methanogenesis (Park et al., 2005, and Charles et al., 2009). These four steps are further illustrated schematically in Figure 2.4. The digestion process begins with bacterial hydrolysis of the input organic matter in order to break down insoluble organic polymers such as carbohydrates and make them available for bacteria. Acidogenic bacteria convert the sugars and amino acids into carbon dioxide, hydrogen, ammonia, and organic acids, where after acetogenic bacteria convert these resulting organic acids into acetic acid, along with additional ammonia, hydrogen, and carbon dioxide. Methanogenic bacteria finally are able to convert these products into biogas i.e. methane and carbon dioxide (Hartmann and Ahring, 2006).

This process operates at atmospheric pressure, producing quality biogas. The biogas composition is generally in the range of 50-85% methane and 15-50% carbon dioxide (Verstraete, 1981). However, biogas consists of approximately 65% CH_4 and other gases like hydrogen sulphide, nitrogen, hydrogen, oxygen, carbon dioxide, ammonia and some volatile organic compounds (Noyola et al., 2006). The exact gas composition is highly dependent on the type, concentration of substrate; and the process parameters like temperature, pH, and alkalinity. For the sake of this research, the temperature, pH, alkalinity, volatile fatty acids, volatile suspended solids in the digester are named as process control parameters.

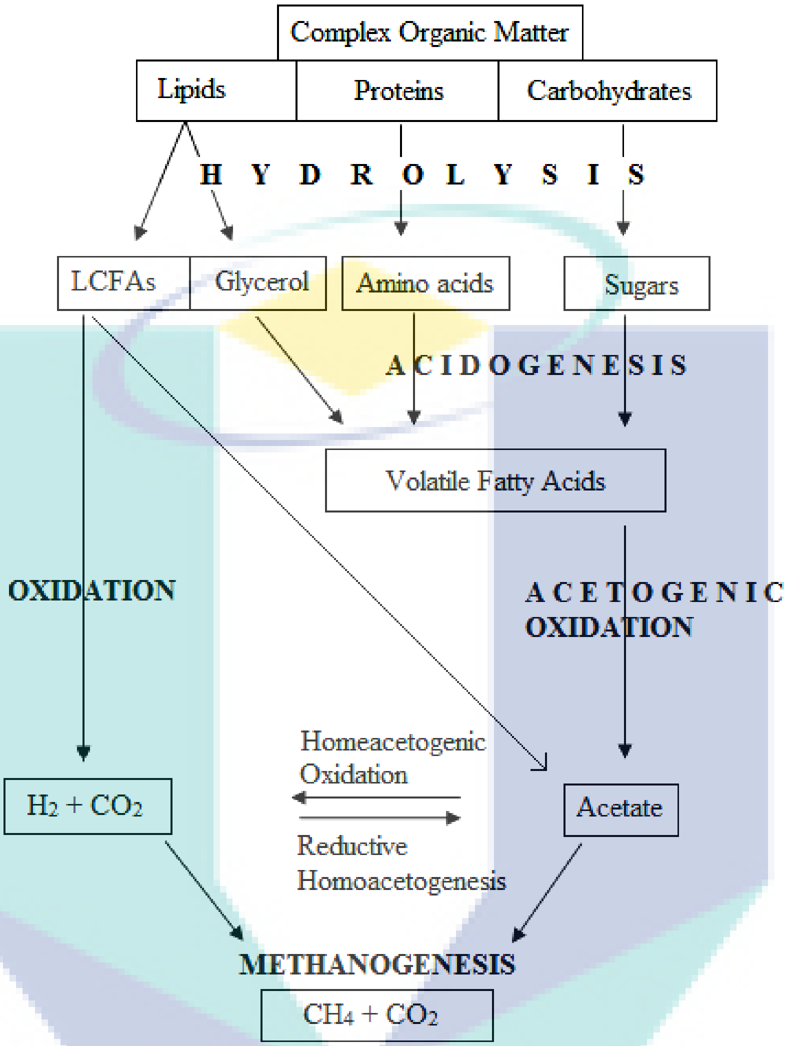

Figure 2.4: Anaerobic conversion of organic matter to biogas, (Adopted from Pavlostathis and Giraldo, 1991).

2.4.1 Hydrolysis

The first step for digestion process is hydrolysis during which, particulate organic matter is converted into soluble compounds that can be hydrolyzed further into simple monomers which can be subsequently utilized by fermentative bacteria. The group of microorganisms responsible for the fermentation process consists of facultative and obligate anaerobic bacteria (Metcalf and Eddy, 2003). Although most of biopolymers are readily degradable, the cellulose of highly lignified material (straw, wood, etc.) has been shown to be resistant to hydrolysis (Lynd et al., 2002). The rate of hydrolysis is a function of factors such as pH, temperature, composition of organic matter and particle size of the substrate (Veeken et al., 2000). Cellulose and hemicellulose are both polymers built up by long chains of sugar units, which after pretreatment and hydrolysis, can be converted into intermediate products which can be transformed into biofuels or other industrially important products (Öztürk et al., 2010).

In conventional methods, the raw materials are split by acid, alkaline, or enzymatic hydrolysis, whereas the addition of further chemicals is necessary for the two former methods. The use of immobilized enzymes in different reactor types leads to incomplete hydrolyses, even at residence times of 8–20 h (Lasch et al., 1987). The ionic product of sub-critical water is as much as three orders of magnitude higher than under ambient conditions. Under these conditions, there is a high H_3O^+ and OH^- ion concentration. As such, some acid-catalyzed organic reactions can be carried out without acid addition. Hence, acid hydrolyzed cleavages of peptide bonds can pass off without the addition of catalysts. However, the ionic product decreases greatly above the critical point. This fact makes sub-critical water an ideal reaction medium for the hydrolysis of organic compounds (Zhu et al., 2010) and for the recycling of different organic wastes, such as municipal solid wastes, refractory pollutants, sludge, different polymers (Chen et al., 2010; Esteban et al., 2008 and Zhang et al., 2010).

2.4.2 Acidogenesis

The second step of an anaerobic digestion is fermentation (also referred to as acidogenesis). In this process, amino acids, sugars, and some fatty acids are further degraded. The organic substrates serve as both the electron donors and acceptors. The principal products of fermentation are acetate, propionate, butyrate, hydrogen and carbon dioxide. The propionate and butyrate are fermented in order to produce hydrogen, carbon dioxide, and acetate. The final products of fermentation (acetate, hydrogen, and carbon dioxide) are the precursors of methane formation (methanogenesis). The free energy change associated with the conversion of propionate and butyrate to acetate and hydrogen requires that hydrogen should be at low concentrations in the system; otherwise the reaction will not proceed (Tchobanoglous et al., 2003).

$$C_6H_{12}O_6 + 2H_2O \longrightarrow 2CH_3COOH + 2CO_2 + 4H_2$$
(2.1)

$$C_6H_{12}O_6 + 2H_2 \longrightarrow 2CH_3CH_2COOH + 2H_2O \qquad (2.2)$$

$$C_6H_{12}O_6 \longrightarrow 2CH_3CH_2CH_2COOH + 2CO_2 + 2H_2 \qquad (2.3)$$

The first reaction (Equation 2.1) is most preferred which produces acetic acid as the major precursor of methane. The next two reactions (Equation 2.2 and 2.3) occur when there is an accumulation of hydrogen in the system. In Equation 2.2, there is a clear utilization of hydrogen while in Equation 2.3, there is hydrogen production in less quantity (two molecules against four in the first reaction).The increase in the acid load of the system is also lower (one mole butyric acid against two moles acetic acid in Equation 2.1).

2.4.3 Acetogenesis

Propionate and butyrate are converted into acetate only by syntrophic acetogens (propionate and butyrate react with water and produce heat) in performance with hydrogen-utilizing methanogens (Kosaka et al., 2006, and Tatara et al., 2008). Propionate-oxidizing bacteria have been identified in microbes in intimate association with methanogens (de Bok et al., 2004). These bacteria are responsible for converting organic products from fermentative bacterial activity such as alcohols, propionic acid and butyric acid into acetic acid, carbon dioxide and water (Rittmann and McCarty, 2001) as shown in Equation 2.4 to 2.7.

$$CH_3CH_2OH + H_2O \longrightarrow CH_3COO^- + H^+ + 2H_2 \qquad (2.4)$$

$$CH_3COO^- + H_2O \longrightarrow CH_3COO^- + HCO_3^- + H^+ + 3H_2 \qquad (2.5)$$

$$CH_3CH_2CH_2OO^- \longrightarrow 2CH_3COO^- + H^+ + 2H_2 \quad (2.6)$$

$$4H_2 + 2CO_2 \longrightarrow CH_3COOH + 2H_2O$$
(2.7)

Acetate is the major intermediate byproduct in the bioconversion of organic matter into methane and carbon dioxide. About 70% of the total methane produced in anaerobic digestion originates from acetate. Thus, the production of methane from acetate is an important step in the anaerobic digestion process (Rittmann and McCarty, 2001).

2.4.4 Methanogenesis

Methanogenesis is carried out by a group of micro-organisms collectively known as methanogens. There are two groups of methanogenic micro-organisms which are involve in methane production. One group is known as aceticlastic methanogens which break the acetate into methane and carbon dioxide. Meanwhile, the second group is known as hydrogen utilizing methanogens, which use hydrogen as the electron donor and carbon dioxide as the electron acceptor to produce methane. Besides that, bacteria within anaerobic processes, acetogens, are also able to use carbon dioxide in order to oxidize hydrogen and thus making acetic acid. However, the acetic acid will be eventually converted into methane. Therefore, the impact due to this this reaction is minor. The conversion of these compounds to CH_4 shown in Equation 2.8 to 2.10 (MetCalf and Eddy, 2003):

$$CH_3COO^- + H_2O \longrightarrow CH_4 + HCO_3^-$$
(2.8)

$$HCO^{3-} + 4H_2 + H^+ \longrightarrow CH_4 + 3H_2O \quad (2.9)$$

$$CH_3COOH \longrightarrow CH_4 + CO_2 \quad (2.10)$$

2.5 MEDIUM AND HIGH RATE ANAEROBIC REACTORS

All modern high rate biomethanation processes are based on the concept of retaining high viable biomass by mode of bacterial sludge immobilization. Several types of bioreactors that are commonly in use one such as batch reactors, a one stage continuously fed system and a two stages or multi-stage continuously fed system. Table 2.6 describes different types of bioreactors used for the anaerobic digestion of various organic wastes. From the table, it can be deduced that the batch reactor are the simplest, filled with the feedstock and left for a period (time depending upon substrate type and COD concentration) that can be considered to be the hydraulic retention time, after which they are emptied.

Anaerobic batch reactors are useful because they can perform quick digestion with simple and inexpensive equipment, and also are helpful in assessing the rate of digestion easily (Parawira et al., 2004, and Weiland, 2006). However, batch reactors have some limitations such as high fluctuation in gas production as well as gas quality (sometimes less amount of CH_4 in biogas), biogas losses during emptying the bioreactors and restricted bioreactor heights (Linke et al., 2006). The second type of bioreactors is known as ‘one-stage continuously fed systems’, where all the biochemical reactions take place in one bioreactor. This type of bioreactors cannot be used for large volumes of wastewater. Moreover bioreactor with only one vessel is not suitable for continuous process of wastewater treatment. The third type of bioreactors are ‘two-stage’ or ‘multi-stage continuously fed systems’, in which various biochemical processes such as hydrolysis, acidification, acetogenesis and methanogenesis take place separately (Ward et al., 2008). The two-stage system is considered a promising process to treat organic wastes with high efficiency in term of degradation yield and biogas production (Fezzani and Cheikh, 2010).

Table 2.5: Different types of bioreactors used for anaerobic digestion of organic waste.

Bioreactor type	**Type of substrate**	**Organic loading rate (kg/m^3/d)**	**Comments**	**References**
ASBR	Fruit and vegetable waste and abattoir wastewater	2.6	A decrease in biogas production was observed due to high amount of free ammonia at high organic loading rate (OLR).	Bouallagui et al. (2009b)
CSTR	Municipal solid waste	15	The reactor showed superior process performance as the OLR progressively increased up to 15 kg/m^3/d.	Angelidaki et al. (2006)
FSAD	Industrial food waste	17	Methane yield of 360 l/kg feed waste with 40 days retention time was observed.	Ike et al. (2010)
IBR	Kitchen waste	8	Showed best performance and biogas production rate was higher than the single reactor.	Guo et al. (2011)
SCR-Laboratory scale	Municipal solid waste and press water from municipal composting plant	20	The reactor performance for biogas production was higher up to 20 OLR but further increase in OLR did not affect the biogas production.	Nayono et al. (2010)
FAFBR-starch based	Primary treated sewage effluent with or without refractory organic pollutants		The efficiency and microbial activity at high OLR was higher than conventional anaerobic fluidized bed bioreactor.	Xing et al. (2010)
MFBR-rotating drum	Municipal solid waste	15	The reactor proved to be stable and helpful in mixing the waste at high OLR, which is usually not possible in mechanically stirred digesters.	Walker et al. (2009)
SMAD	Poultry litter	16	Self mixing at high OLR and high bio-methanization of the poultry litter was observed.	Rao et al. (2011)
SAMR	Sewage sludge, food waste and livestock wastewater	1.8	The reactor showed unstable performance during the initial stage, but performed superior after acclimation formation.	Jeong et al. (2010)
ASCD-self mixing	Olive mill wastewater and olive mill solid waste	14	The best performance in terms of methane productivity, soluble COD and phenol removal efficiencies and effluent quality was observed.	Fezzani and Cheikh (2010)
AHMPR-two phase	Organic waste	3	Compared to a single-stage methanogenic reactor, 11% higher energy was achieved.	Luo et al. (2011)
UASSR	Mixture of maize silage and straw	17	The UASS reactor showed the highest methanogenic performance for the digestion of solid biomass.	Mumme et al. (2010)

All the above types of bioreactors, along with a variety of methanizers such as continuously stirred tank bioreactor, tubular bioreactor; anaerobic sequencing batch bioreactor, upflow anaerobic sludge blanket and anaerobic filters are applied for the treatment of different types of waste (Bouallagui et al., 2005). Table 2.6 summarizes the important features of UASB reactor among other bioreactors.

Table 2.6: Comparison of UASB reactor with other reactors

Characteristics	**UASBR**	**CSTR**	**AF**	**AFB**
Start up period	4-16	--	3-4	3-4
Channeling	Low	Not present	High	Non-existent
Effluent recycle	Not required	Not required	Not required	Required
GSL separator	Essential	Not Required	Beneficial	Beneficial
Carrier packing	Not essential	Not essential	Essential	Essential
Loading rates (kg COD/m^3.day)	10-30	0.25-3	1-4	1-100
HRT (day)	0.5-7	10-60	0.5-12	0.2-5

(Adopted from Kaul and Nandy, 1997)

2.5.1 Anaerobic Fixed Film Reactor (AFFR)

Anaerobic fixed film reactors offer the advantages of simplicity of construction, elimination of mechanical mixing, better stability at higher loading rates, and capability to withstand large toxic and organic shock loads (Rao et al., 2005). In anaerobic fixed film reactors (Figure 2.5), the reactor has a biofilm support structure (media) such as activated carbon, PVC (polyvinyl chloride) supports, hard rock particles or ceramic rings for biomass immobilization and the wastewater is distributed from above or below the media. The main limitation of this design is that the reactor volume is relatively high compared to other high rate processes due to the volume occupied by the media. Another constraint is clogging of the reactor due to increase in biofilm thickness and/or high suspended solids concentration in the wastewater.

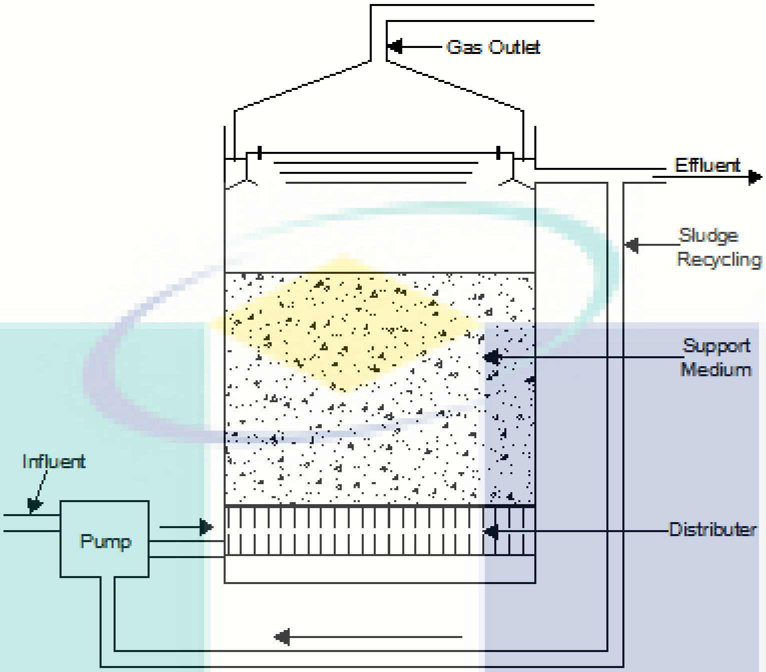

Figure 2.5: Anaerobic fix film reactor

2.5.2 Continuous Stirrer Tank Reactor (CSTR)

Continuous stirrer tank reactor works on the principle of medium rate anaerobic system and it is still widely used for anaerobic digestion (Zhang et al., 2006). The reasons of medium rate application are the simplicity of the design system as well as independence of biomass type. To intensify this simple technology and maintain a viable population of the slow growing methanogens, the CSTRs are usually combined with an internal or external biomass separation and recycling system.

Many solid waste fractions are also treated in CSTRs after slurring with liquid. In the continuously stirred tank reactor, the rate of feeding has to be continuous in order to believe maximum efficiency; however for practical reasons the reactor is fed intermittently; the most common frequency is to feed once a day (Gunaseelan, 1997). The main characteristic of a CSTR system is that its sludge retention time (SRT) is equal to its hydraulic retention time (HRT), and thus no biomass retention occurs (Zeeman and Saunders, 2001, and Kaparaju and Angelidaki, 2007). Normally the CSTR is operated at

HRT of 20-30 days and a loading rate of 1.7 kg-VS/m^3.day. A long HRT has to be employed in order to prevent biomass washout. In spite of that, the disadvantages of the CSTR are that the effluent will contain some fraction of fresh and undigested feed material as well as some of an active microbial population and consequently yielded lower methane concentration (Kasperski, 2008). In addition, this system requires large amount of waste thus the use of large volume reactor is required and eventually leads to high costs for the post treatment of the digester residue (Gunaseelan, 1997). Due to these limitations, there have been developments in the design of new reactors suitable for thick slurries and semi-solid waste. The schematic of CSTR is shown in Figure 2.6.

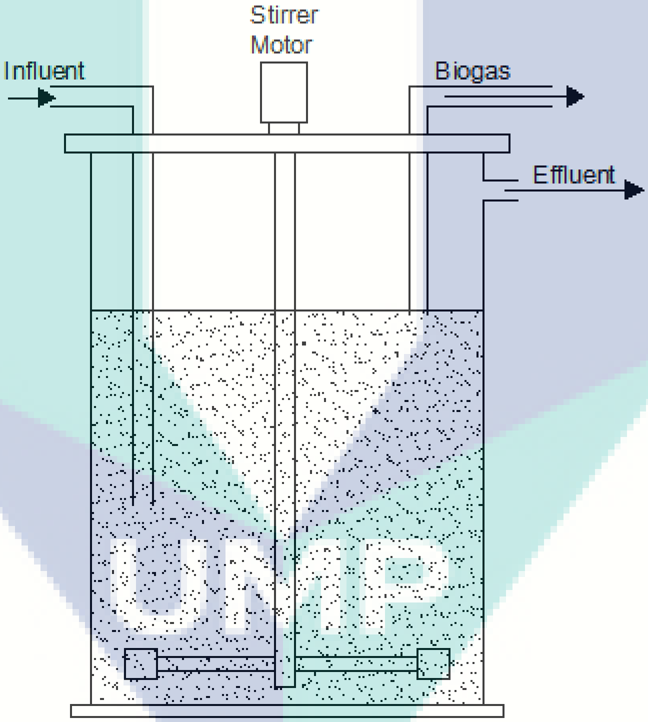

Figure 2.6: Continuous stirrer tank reactor

2.5.3 Anaerobic Fluidized Bed Reactor (AFBR)

In the anaerobic fluidized bed (AFBR) as shown in Figure 2.7, the media for bacterial attachment and growth is kept in the fluidized state by drag forces exerted by the up flowing wastewater. The media used in this system as small particle size sand, activated carbon, and etc. Under fluidized state, each media provides a large surface area for biofilm formation and growth. Thus, it enables the attainment of high reactor biomass hold-up and promotes system efficiency and stability (Sowmeyan and Swaminathan, 2008). This condition is eventually providing an opportunity for higher organic loading rates as well as greater resistance to inhibitors (Wu et al., 1998). Fluidized bed technology is more effective than anaerobic filter technology as it favors the transport of microbial cells from the bulk to the surface and therefore enhances the contact between the microorganisms and the substrate (Sowmeyan and Swaminathan, 2008).

These reactors have been noted to have several advantages over suspended microbial systems such as elimination of bed clogging; a low hydraulic head loss, better hydraulic circulation and a greater surface area per unit of reactor volume (Collivingnarelli et al., 1991, and Perez et al., 2007). Moreover, the capital cost is lower due to reduced reactor volumes.

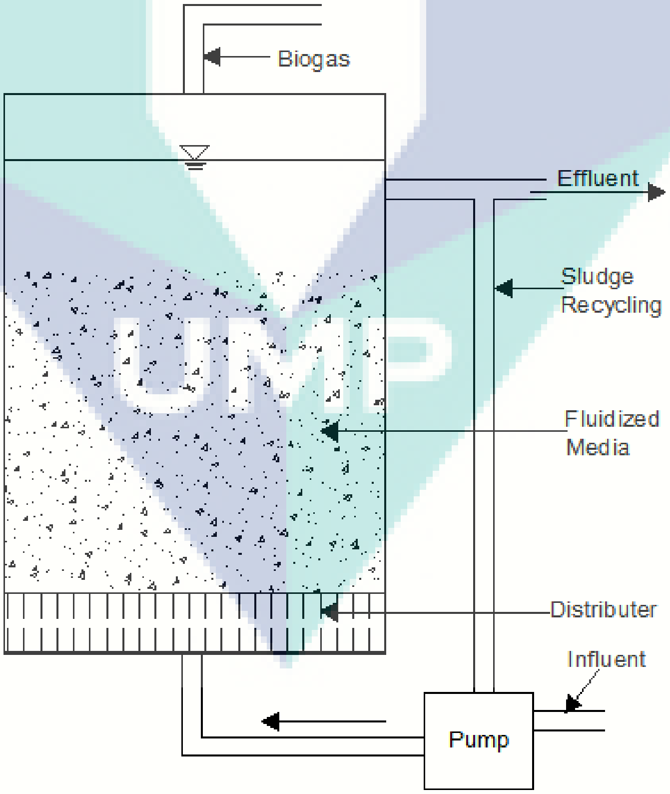

Figure 2.7: Anaerobic fluidized bed reactor

2.5.4 The Upflow Anaerobic Sludge Blanket Reactor (UASBR)

Recently modern anaerobic processes used for high rate reactors have been widely applied to the treatment of a wide variety of industrial wastewaters with a high soluble COD content (Tchobanoglous et al., 2003), including paper-pulp liquors (Elliott and Mahmood, 2007), spent sulphide liquors (Jantsch et al., 2002), and those wastewater from the food industry (Stabnikova et al., 2008). Among the high rate reactors, the UASB is the most commonly used process, with more than 500 installations in the world (Tchobanoglous et al., 2003). The schematic diagram of UASB reactor is shown in Figure 2.8.

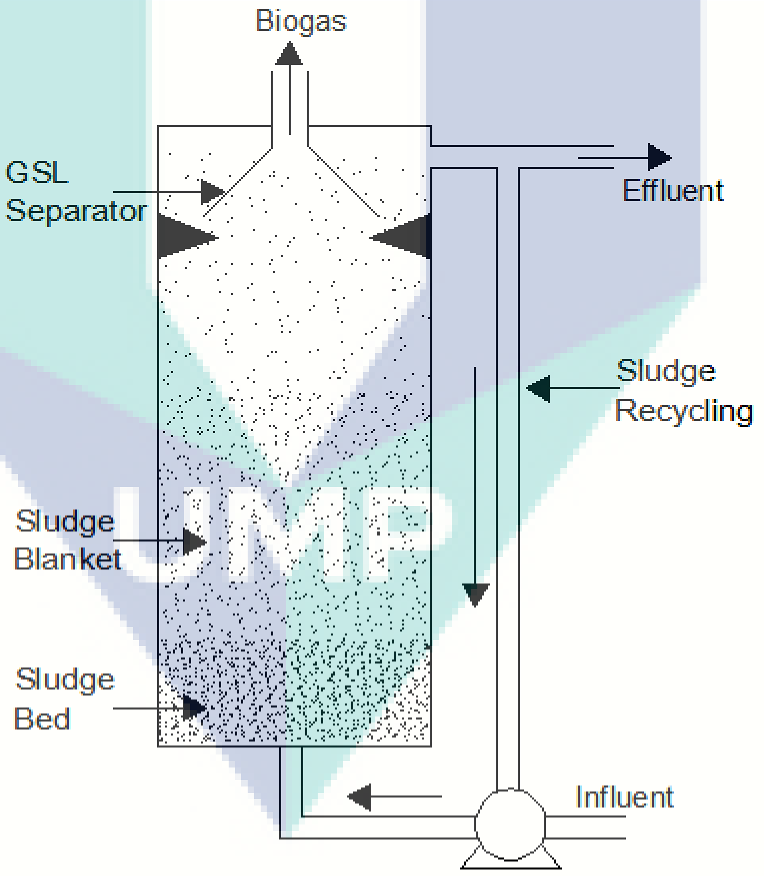

Figure 2.8: Upflow anaerobic sludge blanket reactor

The UASB reactor was developed by Lettinga Association (Lettinga et al., 1980) whereby this system has been successful in treating a wide range of industrial effluents including those with inhibitory compounds. The UASB reactor has been reported to exhibit high OLR, short HRT and low energy demand (Borja and Banks, 1994; Metcalf and Eddy, 2003). The principle of the UASB operation is to have an anaerobic sludge which exhibits good settling properties (Lettinga, 1995, and Jeison and Chamy, 1998) efficiently retains complex microbial consortium without the need for immobilization on a carrier material (for example, as a biofilm). Moreover, the formation of biological granules has good settling characteristics. Performance depends on the mean cell residence time and reactor volume depends on the hydraulic residence time, therefore, UASB reactor can efficiently convert wastewater organic compounds into methane. Among thousands of anaerobic full scale treatment facilities worldwide, approximately 60% are based on the UASB design concept, treating a various range of industrial wastewaters (Jantsch et al., 2002, and Karim and Gupta, 2003). The long HRTs are known to be unfavorable for sludge granulation in UASB reactors (Alphenaar et al., 1993a) whereas, very short hydraulic retention times give rise to possibility of biomass washout. Both scenarios are unfavorable to good performance of the UASB reactor, although granulation has been reported to be necessary for successful domestic and industrial wastewater treatment in UASB reactors (Aiyuk and Verstraete, 2004, and van Haandel et al., 2006).

2.6 THE UASB PROCESS POTENTIAL

Since the concept of the upflow anaerobic sludge blanket (UASB) reactor was proposed in 1970s, the UASB has become the most popular high-rate reactor for the anaerobic treatment of wastewater throughout the world (Ren et al., 2009). However, start-up and restart-up of this reactor is cumbersome due to the high sensitivity of the methanogenic microorganisms (Fang et al., 2010). Because of their high sensitivity to variations in organic loading and other external disturbances, the anaerobic digestion system is often prone to population shifts and process instability (Harada et al., 1996; van Lier et al., 2001). However, Upflow anaerobic sludge blanket (UASB) method has been characterized as the core technology for an anaerobic wastewater treatment method, widely used for the treatment of medium and high organic strength wastewater (Jantsch et al., 2002).

Previously, UASB process has been applied to a low-strength wastewater because of its advantages such as energy saving and low excess sludge (Yoochatchaval et al., 2008, and Syutsubo et al., 2008) however its application is often limited by poor biodegradability of complex organic substrates (Seghezzo et al., 1998; Goodwin et al., 2001; Wolmarans and de Villiers 2002, and Coetzee et al., 2004). Sato et al. (2007) in his study revealed that UASB could be the most suitable option in terms of expenses and treatment efficiency for sewage treatment in the warm regions. However, its applications have also been reported as to have a simple design, easy construction and maintenance, low operating cost; and ability to withstand fluctuation in pH, temperature, and influent substrate concentration (Vasileios and Alexandros, 2007). Recently, UASB reactors are being increasingly used for the treatment of various high-strength industrial wastewaters, many problems are encountered, for instance granulation may not occur and the seeded granular sludge may get washed out (Revanuru and Mishra, 2008). The advantages and disadvantages of UASB reactor have been summarized in Table 2.7.

Recently, the UASB process technique has been applied to a various wastewater treatment such as industrial wastewater, palm oil mill effluent, distillery wastewater, coffee production wastewater, petrochemical wastewater, domestic wastewater, slaughterhouse wastewater, piggery wastewater, dairy waste water, fish meal process wastewater, potato waste leachate as well as a low strength wastewaters like real cotton processing wastewater and synthetic wastewater. The key design and working parameters of UASB reactor used by various researchers have been shown in Table 2.8.

Table 2.7: Advantages and disadvantages of UASB reactor

Advantages

1. Good removal efficiency can be achieved in the system, even at high loading rates and low temperatures.
2. The construction and operation of these reactors is relatively simple and low demand for foreign exchange due to possible local production of construction material, plant components, spare parts and low maintenance.
3. Anaerobic treatment can easily be applied on either a very large or a very small scale.
4. When high loading rates are accommodated, the area needed for the reactor is small thus reducing the capital cost.
5. As far as no heating of the influent is needed to reach the working temperature and all plant operations can be done by gravity, the energy consumption of the reactor is less. Moreover, energy is produced during the process in the form of methane.
6. Reduction of CO_2 emissions due to low demand for foreign (fossil) energy and surplus energy production.
7. Much less bio-solids waste generated compared with aerobic process because much of the energy in the wastewater is converted to a gaseous form and resulting in very little energy left for new cell growth.
8. The sludge production is low, when compared to aerobic methods, due to the slow growth rates of anaerobic bacteria. The sludge is well stabilized for final disposal and has good dewatering characteristics. It can be preserved for long periods of time without a significant reduction of activity, allowing its use as inoculum for the start-up of new reactors.
9. Can handle organic shock loads effectively.
10. Low nutrients and chemical requirement especially in the case of sewage, an adequate and stable pH can be maintained without the addition of chemicals.
11. Macronutrients (nitrogen and phosphorus) and micronutrients are also available in sewage, while toxic compounds are absent.

Disadvantages

1. Pathogens are only partially removed, except helminthes eggs, which are effectively captured in the sludge bed. Nutrients removal is not complete and therefore a post treatment is required.
2. Due to the low growth rate of methanogenic organisms, longer start-up takes before steady state operation, if activated sludge is not sufficiently available.
3. Hydrogen sulphide is produced during the anaerobic process, especially when there are high concentrations of sulphate in the influent. A proper handling of the biogas is required to avoid bad smell and corrosion.
4. Post-treatment of the anaerobic effluent is generally required to reach the surface water discharge standards for organic matter, nutrients and pathogens.
5. Proper temperature control (15-35 $^{\circ}$C) required for colder climates.

Table 2.8: UASB reactor specifications and working parameters

Type of wastewater	Phase	Reactor Volume (L)	Diameter (cm)	Height (cm)	Flow rate (L/d)	HRT (d)	Upflow Velocity (m/h)	Sampling ports	Reference
Palm oil mill effluent	Two	12	13	90	11.76	1	--	6	Borja et al. (1996)
		5	9	78	4.9	1	--	6	
	Single	10	9-12	140	3	3.33	--	4	Chaisri et al. (2007)
	Single	14	18.5	52	383.26[a]	0.53	0.59	3	Siang (2006)
Distillery WW	Single	1.05	6.2	34.8	0.5	2	--	--	Goodwin et al. (2001)
	Two	10.2	9	160	4	2.5	--	15	Laubscher et al. (2001)
	Three	2.3	5	83	1.84	1.25	2	--	Keyser et al. (2003)[b]
	IUASB-Sinlge	143000	1102	1500	47667	3	--	--	Ince et al. (2005)
	TUASB-Sinlge	476000	2010	1500	119000	4	--	--	
	CUASB-Sinlge	190000	1270	1500	38000	5	--	--	
Dairy WW	Phase 1	12.3	15	70	3.5	3.5	--	4	Luostarinen & Rintala (2005)
	Phase 2	3.2	9	50	2.13	1.5	--	3	
	Single	31.7	15.4	170		--	0.11		Nadais et al. (2005)
	Intermittent	6	9.4	86.4	12	0.5	--	--	Nadais et al. (2005a)
	Single	5	10	70	5	1	--	5	Tawfik et al. (2008)
Fishery WW	Single	7.85	10	100	--		--	5	Huang et al. (2009)
Slaughterhouse WW	Single	31840	260	600	96485	0.33	--	5	Sayed et al. (1984)
	Single	3	6.7	85	24	0.13	--	--	Chávez et at. (2005)
	Three	7.2	15	41	7.9-12.4	0.91-0.58	--	3	Caixeta et al. (2002)
	Single	1000	--	--	10000	0.1	0.33-1.0	9	Torkian et al. (2003)
	Two	2	8	15	0.59	3.4	--	6	Ruiz et al. (1997)
Piggery WW	Single	5	15	30	1	5	--	--	Sánchez et al. (2005)

	Single	3.78	6W, 6L	105	6	0.63	2 as Vs	6	Huang et al. (2005)
	Single	800 x 10^3	940	1160	--	--	--	--	Miranda et al. (2005)
	Two	1	3.9	84	3	0.33	--	4	Hendriksen & Ahring (1996)
		2	5.7	78	4.37	0.46	--	4	
Municipal WW and waste	Two	2.5	6	100	2	1.25	--	5	Ağdağ & Sponza (2005)
	Single	40	16	240	121	0.33	--	--	El-Gohary & Nasr (1999)
	Two	25	11	300	119	0.21	--	--	
	Single	55	28	89	172	0.32	--	1	Behling et al. (1997)
	Single	8	10	100	11.4	0.7	--	5	Singh & Viraraghavan (1998)
	Three[c]	46	15W, 25L	125	--	0.17-0.13	0.31-0.43	4	Moawad et al. (2009)
	Single	15.7	10	200	80	0.196	0.426	7	Uemura & Harada (2000)
	Single	2.3	5	90	6	0.33	1	--	Aiyuk & Verstraete (2004)

[a] total flow rate (influent and recycled)

[b] design adopted from Trnovec & Britz (1998)

[c] only UASB data

2.6.1 The Process Control Parameters for UASB Reactor

Organic loading rate and hydraulic retention time

Organic loading rate is an important parameter in which affecting microbial ecology and performance of UASB systems (Rincón et al., 2008). This parameter actually integrates reactor characteristics, operational characteristics, and bacterial mass and activity into the volume of media (Torkian et al. 2003). Various studies have proven that higher OLRs will reduce COD removal efficiency in wastewater treatment systems (Patel and Madamwar, 2002; Torkian et al. 2003 and Sánchez et al., 2005). However, with the increasing OLR, the gas production will increase to a stage where methanogens could no longer work quickly to convert acetic acid to methane. Moreover, organic loading rate can also be related to substrate concentration (influent COD and suspended solids) and HRT. The high COD and suspended solids will increase the OLR and also HRT will be increased in which anaerobic process will not remain stable for the continuous production of biogas. The HRT also related to flow rate of influent stream and the volume of reactor. On the other hand, if flow rate will be too high then the digester will be filled very fast in which HRT will be shorten. Laubscher *et al*. (2001) used POME as a substrate in 10 liter UASB reactor and maintained a flow rate of 3 L/day. In another study, Chaisri (2007) used distillery wastewater and maintained a flow rate of 4 L/day. The difference in flow rate is due to the type and substrate concentration.

pH and Alkalinity

A range of pH values suitable for anaerobic digestion has been reported by various researchers (Huber et al., 1982). However, the optimal pH for methanogenesis has been found to be around 7.0 (Yang and Okos, 1987). Meanwhile, Ağdağ and Sponza (2007) reported a very narrow range of suitable pH (7.0–7.2) in the industrial sludge added bioreactors during the last 50 days of the anaerobic incubation. Similarly, Ward et al. (2008) found that a pH range of 6.8–7.2 was ideal for anaerobic digestion. Lee et al.

(2009b) reported that methanogenesis in an anaerobic digester occurs efficiently at pH 6.5–8.2, while hydrolysis and acidogenesis occurs at pH 5.5 and 6.5, respectively (Kim et al., 2003). From the batch experiments, Park et al. (2008) showed that the appropriate pH range for thermophilic acidogens was 6–7. Dong et al. (2009) suggested that the hydrogen production will be at a maximum if the initial pH of a bio-system is maintained at 9. However, similar results can also be achieved at pH 5–6 (Kapdan and Kargi, 2006). Liu et al. (2008) showed that the most favorable range of pH to attain maximal biogas yield in anaerobic digestion is 6.5–7.5.

The alkalinity represents the capacity of a substrate to neutralize the acidity. The alkalinity plays an important role in anaerobic system in which volatile fatty acid concentration increased due to metabolic transformation and pH drop. In such case the waste wastewater with high alkalinity can neutralize pH and system again become stable. (Isik and Sponza, 2005). The increase in acidity of the system is due to the high organic loadings. It is reported that for a stable anaerobic process, the alkalinity should be be in the range of 250 to 950 mg/L (Singh et al., 1999).

Carbon to Nitrogen Ratio (C/N)

The C/N ratio in the organic material plays a crucial role in anaerobic digestion. The unbalanced nutrients are regarded as an important factor limiting anaerobic digestion of organic wastes. For the improvement of nutrition and C/N ratios, co-digestion of organic mixtures is employed (Cuetos et al., 2008). Co-digestion of fish waste, abattoir wastewater and waste activated sludge with fruit and vegetable waste facilitates balancing of the C/N ratio. Their greatest advantage lies in the buffering of the organic loading rate, and anaerobic ammonia production from organic nitrogen, which reduce the limitations of fruit and vegetable waste digestion. The C/N ratio of 20–30 may provide sufficient nitrogen for the process (Weiland, 2006), and Bouallagui et al. (2009a) suggested that a C/N ratio between 22 and 25 seemed to be best for anaerobic digestion of fruit and vegetable waste, whereas, Guermoud et al. (2009) and Lee et al. (2009b) reported that the optimal C/N ratio for anaerobic degradation of organic waste was 20–35.

Temperature

Many researchers have reported significant effects of temperature on the microbial community, kinetics process and stability as well as methane yield (Dela-Rubia et al., 2002; Bouallagui et al., 2009a, and Riau et al., 2010). Lower temperatures during the process are known to decrease microbial growth, substrate utilization rates, and biogas production (Kim et al., 2006, and Trzcinski and Stuckey, 2010). Moreover, lower temperatures may also result in an exhaustion of cell energy, a leakage of intracellular substances or complete lysis (Kashyap et al., 2003). In contrast, high temperatures lower the biogas yield due to the production of volatile gases such as ammonia which suppresses methanogenic activities (Fezzani and Cheikh, 2010). An anaerobic digestion at low temperatures was found to require long HRT with satisfactory production and composition of biogas. In other study, Sánchez et al. (2001) studied the influence of temperature and substrate concentration on the anaerobic batch digestion of piggery wastewater. They compared the process at mesophilic temperature (35 ^{o}C) and ambient range of 16.8–29.5 ^{o}C, and the influent concentrations in the range of 3.3–26.3 g-TCOD/L. The process at mesophilic temperature was found more stable than at ambient temperature, which obtaining higher values of COD removal efficiency.

Generally, anaerobic digestion is carried out at mesophilic temperatures (El-Mashad et al., 2003). The operation in the mesophilic range is more stable and requires a smaller energy (Fernandez et al., 2008). Castillo et al. (2006) found that the best operational temperature was 35 °C with an 18 day digestion period meanwhile a little fluctuation in temperature from 35 °C to 30 °C has caused a rate reduction of biogas production (Chae et al., 2008). Overall, a temperature range between 35–37 °C is considered suitable for the production of methane and a change from mesophilic to thermophilic (45-55 °C) temperatures can cause a sharp decrease in biogas production until the necessary populations of methane producing bacteria. Briski et al. (2007) reported that for biodegradation, the temperature must be below than thermophobic (60-65 °C) temperature because above 65 °C denaturation of enzymes occurs. However, thermophilic conditions have certain advantages, such as a faster degradation rate of organic waste, higher biomass and gas production, less effluent viscosity and higher pathogen destruction (Zhu et al., 2009). Tsukahara et al. (1999) used liquidized food waste in UASB at 35 ^{o}C (mesophilic condition) and reported that 60% TOC removal efficiency was achieved during first week at volumetric loading rate (VLR) of 3.60 g-TOC/L-reactor.day and a week after the TOC removal efficiency increased up to a maximum value of 71 % at VLR of 5.95 g-TOC/L-reactor.day.

Upflow Velocity

The upflow velocity (V_{up}) is also an important operational parameter in upflow digesters (UASB reactor and upflow anaerobic fix film reactor). It maintains the mixing and hydraulic retention time of the substrate and biomass. An upflow velocity is directly proportional to reactor height and inversely proportional to hydraulic retention time. It determines the appropriate mixing of biomass with the height of the reactor with or without channeling. The permissible limit of upflow velocity is 0.5-1.5 m/h as described by different researchers (Torkian et al., 2003; Huang et al., 2005 and Siang, 2006). Siang (2006) maintained a 0.59 m/h V_{up} at and HRT of 13 h. Meanwhile, Keyser et al. (2003) reported a 2 m/h V_{up} at HRT of 1.25 days which is quite higher than limits but they used distillery wastewater as substrate and little fluctuations in upflow velocity might be possible with different types of wastewaters.

Nutrients

It is widely known that the growth and activity of the methanogenic consortium in anaerobic reactors strongly depend on environmental factors, such as nutrients availability (Cresson et al., 2006). The existence of nutrients such as nitrogen, phosphorus and sulfur in the influent is fundamentally needed to accomplish a successful development of granules. Basically, throughout the first phases of granules formation, surplus nutrients into influent can enhance the process and have no deadly effect (Ma et al., 2009). While lack of nutrients in substrate can adversely affect the granules formation process, whereby it has been reported that at a concentration of less than 300 mg/L of nitrogen will be resulted in granules growth dwindling (Singh et al., 1999). Considering that the composition of ammonia is totally produced as a result of nitrogen correlation with hydrogen to introduce electrons donor, it can be concluded that ammonia is playing a significant role on pH in medium (Bettazzi et al., 2010). Also it is important to mention that methanogens organisms are generally utilizing ammonia as a food source which is giving more acceleration for granules activities in the methanogenic phase (Yang et al., 2005).

Moisture

High moisture contents usually facilitate the anaerobic digestion; however, it is difficult to maintain the same level of water throughout the digestion cycle (Hernandez-Berriel et al., 2008). High water contents are likely to affect the process performance by dissolving readily degradable organic matter. It has been reported that the highest methane production rates occur at 60–80% of humidity (Bouallagui et al., 2003). Hernandez-Berriel et al. (2008) studied methanogenesis processes during anaerobic digestion at different moisture levels of 70% and 80%. They found that the onset of the methanogenic phase took place after ten weeks in both cases, at 70% and 80% of moisture content. However, bioreactors under the 70% moisture regime produced a stronger leachate and consequently a higher methane production rate. At the end of the experiment, 83 ml methane per gram dry matter were produced at the 70% moisture level, while 71 ml methane per gram dry matter were produced with the 80% moisture. Nonetheless, bioreactors from both moisture regimes showed similar ratios (0.68) of biochemical oxygen demand (BOD) to chemical oxygen demand (COD).

Mixing

Mixing will not only provide a good contact between microbes and substrates, it also helps to reduces resistance to mass transfer, minimizes buildup of inhibitory intermediates and stabilizes environmental conditions (Grady et al., 1999). When mixing is inefficient, overall HRT of the substrate will be impaired by pockets of material at different stages of digestion whereby every stage has a different pH and temperature (Stafford, 1982, and Kaparaju et al., 2008). Mixing can be accomplished through mechanical mixing, biogas recirculation or through slurry recirculation (Karim et al., 2005). Studies have been done to observe the effects of mixing to the performance of anaerobic digesters previous researchers and it was found that mixing improved the performance of digesters treating waste with higher concentration (Karim et al., 2005a). Slurry recirculation has been shown to have better results compared to impeller and biogas recirculation mixing mode (Karim et al., 2005b).

Mixing also has been reported to improv the gas production as compared to unmixed digesters (Karim et al., 2005a). Stafford, (1982) has long reported that on intermittent mixing is advantageous over vigorous mixing (Kaparaju et al., 2008). Sludge granules are formed due to fluidization (Guiot et al., 1992) which is achieved by mixing of the sludge by flow and gas release. Rapid mixing is not encouraged because methanogens can be less efficient in this mode of operation (Gerardi, 2003). However, Karim et al. (2005a) mentioned that mixing during start-up is not beneficial due to the fact that digester pH will be dropped, resulting in performance instability as well as leading to a prolonged start-up period. The upflow reactors with big diameters could face the problem of channeling where as upflow velocity, sometimes, cannot improve the mixing of more viscous substrate. So, mixing becomes the important functional parameter for such cases.

The most commonly used operational parameters like pH, operational temperature, hydraulic retention time and organic loading rates have extensively been discussed in Table 2.9 along with COD removal and biogas production.

Table 2.9: Operation and performance of UASB reactor with various wastewaters

	Type of wastewater	Reactor type	Phase	Influent COD (g/L)	OLR (kgCOD/ m.d)	HRT (days)	Temperature (°C)	COD removal (%)	Biogas (L/d)	CH_4 (L/d)	Average CH_4 %	Reference
POME	POME	UASB	Single	42.5	10.63	4	35	96	11.5	6.9	60	Borja & Banks (1994)
	POME	UASB	Two	30.6	30	1.02	35	90	10	7	70	Borja et al. (1996)
	POME	UASB	Single	50	15.5	3.33	28	80.5	14	7	50	Chaisri et al. (2007)
Distillery WW	Recalcitrant distillery WW	UASB	Single	10	19	0.53	55	<67	6.4	3.5	55	Harada et al. (1996)
	Distillery WW	AF-UASB	Two	8.51-16.8^a	4	19-10	36±1.5	47	0.091-0.39	0.06-0.26	65	Blonskaja et al. (2003)
				13.6^b	2.2	20	36±1.5	93	5.3	3.45	65	
	Malt whisky distillery pot ale	UASB	Single	21.05^c	10.2	2.1	35	93	4.7	--	--	Goodwin et al. (2001)
				32.86^d	4.69	7	35	88	1.3	--	--	
	Malt whisky WW	UASB	Two	20.92	17.2	1.22	35±1.5	92	310^e	238^f	77^f	Uzal et al. (2003)
	Grap wine distillery WW	UASB	Single	30	18	1.67	34-36	90±3	--	--	--	Wolmarans & de Villiers (2002)
	Grain distillation WW	UASB	Two	5.1^g	18.4	0.28	35	90±3	--	--	--	Laubscher et al., 2001
	Winery effluent	UASB	Three	6.4	5.1	1.25	35	86	2.3^h	--	--	Keyser et al. (2003)
	Raki & Cognac distilleries	IUASB	Single	33	11	3	36±1	85	0.078^k	0.045	74	Ince et al. (2005)
		TUASB		32	8.5	4	36±1	60-80	0.078^k	0.045	74	
		CUASB		23	4.5	5	36 ± 1	70-80	0.071^k	0.041	74	
Dairy WW	Dairy WW	UASB	Single	37	6.2	6	35	98	--	--	--	Gavala et al. (1999)
	Dairy palour WW	UASB-Septic	Phase 1	0.63	0.179	3.5	20	73	--	--	--	Sari & Jukka (2005)
			Phase 2	0.36	0.24	1.5	10	64	--	--	--	
	Dairy & Domestic WW	UASB-AS	Single	2.01	3.4	1	35	69	--	--	--	Tawfik et al. (2008)
	Dairy WW	UASB	Single	79	7.5	0.66	35±1	74	--	16	--	Nadais et al.

												(2005)
	Dairy WW	UASB	Intermittent	13.5	22	2	35±1	97±1	--	54	--	Nadais et al. (2005a)
	Digested cowdung slurry	UASB	Single	1.8	13.5	0.13	30±2	>90	--	--	--	Ramasamy et al. (2004)
	Dairy WW	UASB	Continuous	12.48	12.48	1	35±1	90	--	--	--	Nadais et al. (2006)
			Intermittent	12.48	12.48	1	35±1	90	--	--	--	
	Cheese whey	UASB	Two	55.1	11.1	4.95	--	95	--	23.4^{m}	--	Ergüder et al. (2001)
	Dairy manure	UASB	Two	17.8	8.9	2	35±1	87.3	--	0.27^{n}	--	García et al. (2008)
Fishery WW	Mixed sardine and tuna canning	UASB	Single	2.72	8	0.33	--	80-90	--	--	--	Palenzuela-Rollon et al. (2002)
Slaughterhouse WW	Slaughterhouse waste	UASB	Single	1.2	3.5	0.33	20	70	10000	6500	65-70	Sayed et al. (1984)
	Poultry slaughter WW	UASB	Single	5.5	28.7	0.19 as CRT	24.7	95	--	--	--	Chávez et at. $(2005)^{p}$
	Slaughterhouse waste	UASB	Three	4.2	4.6	0.92	35	89	11.9	--	--	Caixeta et al. (2002)
				6.5	8.7	0.75		90	10.9	--	--	
				6.3	10.8	0.58		86	10.6	--	--	
	Slaughterhouse WW	UASB	Five	2.87	30	0.1	33	90	--	280^{q}	--	Torkian et al. (2003)
	Slaughterhouse WW	UASB-AF^{1}	Two	7.6	2.23	3.4	37	93	1.03	0.6	70.6	Ruiz et al. (1997)
Piggery WW	Piggery waste	UASB	Single	8.12	1.62	5	30-35	75	4.1	2.37^{r}	57.8	Sánchez et al. (2005)
	Pre-settled piggery WW	UASB-AS	Two	2	3.17	0.63	30±1	91	1.51	0.6	39.7	Huang et al. (2005)
Muncipal WW	Municipal landfill leachate	UASB-CSTR	Two^{2}	20	16	4.5	37±3	79	9.5	5.7	60.05	Ağdağ & Sponza (2005)
	Domestic WW	UASB	Single	0.39	1.21	0.32	30	85	26	--	--	Behling et al. (1997)
	Domestic WW	UASB-SBR^{1}	Three	0.37	2.93	0.125	--	57	--	--	--	Moawad et al. (2009)

	Muncipal WW	UASB-AS	Five[3]	0.56	0.09	0.17	--	85	--	--	--	von Sperling et al. (2001)
	Muncipal WW	UASB	single	3.2	1.05	0.42	20±1	86	1.97	1.1	79	Singh & Viraraghavan (1998)
	Sewage WW	UASB	Single	0.15-0.5	0.77-2.55	0.196	25-13	68±4	--	3.5	--	Uemura & Harada (2000)
	Sewage WW	AF-UASB[1]	Two	0.47-1.23	1.4-3.7	0.33	12-23	<50	--	--	--	Sawajneh et al. (2010)

[1] only UASB data
[2] 1st run data, total 3 runs
[3] phase 3 data
[a] acidogenic phase
[b] methanogenic phase
[c] with 70% pot ale
[d] with 100% pot ale
[e] at influent of COD 16 g/L
[f] stoichiometric calculations for CH_4
[g] at controlled conditions
[h] at OLR of 6.3 kg-COD/m^3.day
[k] as per gVSS
[m] L-CH_4/L of cheese whey
[n] L-CH_4/g-$COD_{removed}$
[p] all data interms of BOD
[q] as per SCOD
[r] per liter of influent

2.7 Important Calculation Methods for Anaerobic Treatment Process

2.7.1 The F/M Ratio

Before starting the experiment, the first step achieved was the determination of food to micro-organism ratio (F/M). The F/M ratio is the key factor controlling anaerobic digestion. At a given temperature, the bacterial group can only consume a limited amount of food each day. In order to consume the required number of kilograms of waste, one must supply the proper number of kilograms of bacteria. The ratio of the kilograms of waste supplied to the kilograms of bacteria available to consume the waste is the food to F/M. Lower F/M ratio will result in a greater percentage of the waste being converted to gas. Unfortunately, the bacterial mass is difficult to measure since it is difficult to differentiate the bacterial mass from the influent waste. The task would be easier if all of the influent waste were converted into biomass or gas. In that case, the F/M ratio would simply be the digester loading divided by the concentration of volatile suspended solids (biomass) in the digester (kg-COD/kg-VSS.day). For any given loading, efficiency can be improved by lowering the F/M ratio and increasing the concentration of biomass in the digester. Also for given biomass concentration within the digester, the efficiency can be improved by decreasing the loading.

$$\frac{\mathrm{F}}{\mathrm{M}} = \frac{\text{Organic loading rate}}{\text{Volatile solid}} \qquad \text{Eq. 2.1}$$

Where,
Organic loading rate = COD of the influent stream (kg-COD/L.day)
Volatile solids = Volatile suspended solids concentration in the reactor (kg-VSS/L)
F/M = kg-COD/kg-VSS.day

This is most important step before each organic loading. While calculating F/M ratio, VSS of the digester were compared with the designed organic loading rate and the

F/M results allowed us to make new feed. The F/M ratio ranges from 0.1-0.6 where low ratio can participate in high biogas production.

2.7.2 The Hydraulic Retention Time

The hydraulic retention time calculation before proceeding experiments is also an important process control parameter. It shows the total time required by the liquid to degrade. In other words, HRT is the time consumed by a liquid entering from inlet to the outlet until all soluble matter could be degraded. The HRT plays an important role while anaerobic digestion of which the liquid has to stay within the digester until degradation. This can be assumed both for aerobic facultative ponds and anaerobic treatment. Perhaps, it is the measure of the average length of time that a soluble compound remains in a bioreactor.

Followed by the F/M ratio, the hydraulic retention time (HRT) can be calculated as,

$$\mathrm{HRT} = \frac{\mathrm{CODin}}{\mathrm{OLR}} \qquad \text{Eq. 2.2}$$

Where,
HRT = Hydraulic retention time (days)
OLR = Organic loading rate (kg-COD/L.day)
CODin = Influent COD (Kg-COD/L)

2.7.3 The Flow Rate

The calculated organic loading rate and hydraulic retention time from Equation 1 and 2 respectively thus contribute in flow rate (Q) calculation. The flow rate calculation also plays an important role in UASB treatment process. The HRT and flow rate examine the exact influent stream from feed inlet to the effluent outlet. Normally, flow rate is controlled by means of a peristaltic pump with corresponding tube hosing of different diameter. The lower will be the diameter of tube hose; less will be the flow rate of the influent stream and vice versa. The flow rate is designed according to the working volume of the reactor. The flow rate can be calculated as follows,

$$Q = \frac{Vw}{HRT} \qquad \text{Eq. 2.3}$$

Where,

Q = Flow rate of influent stream (L/day)

V_w = Working volume of the reactor (L)

HRT = Hydraulic retention time (days)

2.7.4 The Upflow Velocity

For calculating upflow velocity (V_{up}), flow rate and cross-sectional area (A) are required. The cross-sectional area is a known parameter and flow rate derived from Equation 3 will help in calculating the upflow velocity. The upflow velocity determines the mixing of the biomass. The upflow velocity can be calculated by the following expression.

$$V_{up} = \frac{Q}{A} \qquad \text{Eq. 2.4}$$

Where,

V_{up} = Upflow velocity (m/h)

Q = Flow rate of influent stream (m^3/ h)

A = Cross-sectional area of the reactor (m^2)

In another equation, the upflow velocity can be calculated as,

$$Vup = \frac{H}{HRT} \qquad \text{Eq. 2.5}$$

Where,

V_{up} = Upflow velocity (m/h)

H = Height of the reactor up to working volume (m)

HRT = Hydraulic retention time (hours)

2.8 THE CEMENT KILN DUST

Cement kiln dust (CKD) is a fine-grained, alkaline matter generated as a by-product of the cement manufacturing process. In general, the key raw material used to produce cement clinker is limestone, silica, alumina and iron. During calcination process, these raw materials are subjected to a high temperature thus altering their chemical composition. The annual clinker production from a cement plant ranges from 528-900 metric tons depending upon the production capacity of plant and 16-60 kg-cement kiln dust produced per ton of clinker (PCA, 2007). Annual use of CKD for beneficial applications has ranged from 574-1160 metric ton in U.S.A. and its beneficial uses centered on four primary applications: stabilization of soils, stabilization of waste, cement additive/blending, and mine reclamation (Adaska and Taubert, 2008). In waste stabilization, the use of CKD is essential as the average levels of trace metals found in the CKD were significantly below the regulation limits (PCA, 1992).

Furthermore, CKD has been long reported as a substitute for lime in stabilizing wastewater streams (Nicholson 1988a and b). This is primarily because of its high neutralizing potential. Meanwhile, Burnham (1987) also reported by adding 35% of CKD in sewage sludge has satisfactorily met the specified pathogen control level in sewage sludge. The CKD addition in wastewater helps in increasing the pH and accelerated drying of wastewater along with an exothermic reaction. (Nicholson and Burnham, 1988). In practice, the neutralization of acidic wastewater streams often involves addition of chemicals to raise pH levels as well as to precipitate soluble metals. By elevating the pH of the waste streams the dissolved metals will be precipitated in the form of hydroxides which then can be removed through either settling or filtration processes (El-Awady and Sami, 1997). Mackie et al. (2010) showed potential application of CKD with low free lime contents to be effective in neutralizing acidic wastewater. Moreover, the most common material used in treatment of acid rock drainage from mines is quicklime in the form of lime slurries (Younger et al., 2002). Until now, none study on the feasibility of CKD as neutralizing agent for the treatment of POME has been conducted. For example, Smith and Campbell (2000) used CKD instead of sodium hydroxide (NaOH) for the treatment of pulp and paper industry wastewater. In another study, Ayoub et al. (2002) used lime to raise pH of pulp and paper wastewater effluent above 11.4. Table 2.10 shows the characteristics of CKD.

Several cases of reactor failure have been reported in earlier studies of wastewater treatment due to the accumulation of high volatile fatty acids (VFAs), causing a drop in pH which inhibited methanogenesis (Patel and Madamwar, 2002; Parawira et al., 2006). According to Boone and Xun (1987) most methanogenic bacteria have optimal growth between pH 7 and 8, whereas VFA degrading bacteria have lower pH. It was found that digester could tolerate acetic acid concentrations up to 4000 mg/L without inhibition of gas production (Stafford, 1982). To control the level of VFA in system, alkalinity has to be maintained by recirculation of treated effluent (Borja et al., 1996; Najafpour et al., 2006) or by the addition of lime, caustic soda or bicarbonate of soda (Caixeta et al., 2002; Gerardi, 2003). CKD has higher contents of CaO (lime) which helped in maintaining the pH of digester wastewater. Moreover, the CKD is a freely available material from cement industry. Furthermore, NaOH and $NaHCO_3$ are very expensive than CKD and the cost will be minimized by using CKD as neutralizing the wastewater.

Table 2.10: Characteristics of cement kiln dust

Parameter	**Dry-kiln CKD (% by weight)**	**Dry-kiln CKD (% by weight)**
pH	13.6*	-
CaO	51	44.9
SiO_2	11.6	9.64
Al_2O_3	6.1	3.39
Fe_2O_3	3.3	1.10
MgO	1.1	1.29
K_2O	1.69	2.4
SO_3	5.4	6.74
Particle size	<25 µm	1-40 µm

* unitless parameter
(adapted from Adaska and Taubert, 2008)

CHAPTER 3

MATERIALS AND METHODS

3.1 INTRODUCTION

For the purpose of this study, three samples; food waste, POME and CKD were used. All the samples were characterized prior to use and the characterization methods were explained in detail in the following subchapter. This chapter also includes the UASB reactor design involved in this study.

3.2 SAMPLES COLLECTION

Food wastes were collected from main cafeteria of University Malaysia Pahang (UMP). The volume of food waste generation from all students is not the part of this study. The food waste was collected to determine the treatment feasibility using UASB reactor. The food waste collection bins have been separated into organic waste, plastics and bones. The food waste, consisting of rice, coconut, vegetables, fruit, chicken meat, fish and egg shells, was used for anaerobic digestion. Food waste samples were collected at afternoon 15:00 and evening 22:00 because of high waste generation at these periods. Food waste samples were collected manually from the bins and being segregated for their components as shown in Table 3.1. The bones and any other material like plastic or glass was separated manually from the collected samples except those parts which were not able to be taken out by any means. Organic fraction was taken out from composite sample and analyzed for its components. Figure 3.1 shows the solid waste collection point of which the food waste was selected.

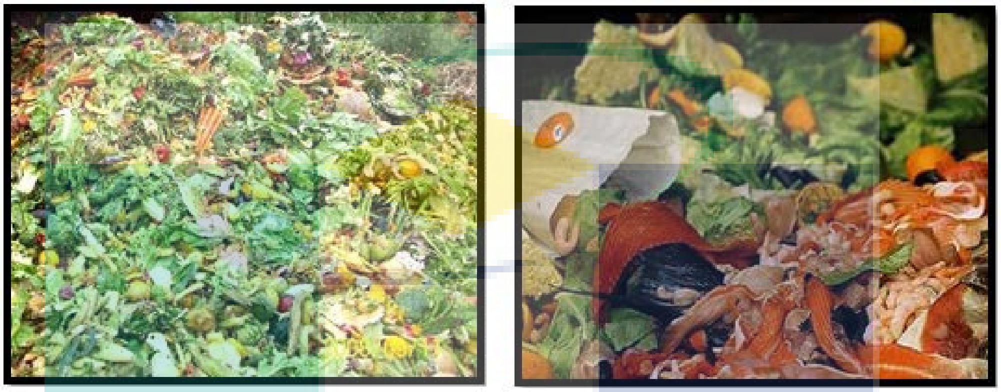

Figure 3.1: Food waste collection point

Meanwhile the raw POME samples were continuously collected from de-oiling pond at KSLH oil palm mill. This mill is the subsidies of Felda Oil Palm Industries situated 62 Km west from Kuantan city. The mill is capable of process 1200-1450 tons of fresh fruit bunches (FFB) daily. 54 tons of FFB were processed per hour for duration of 24 hours without stopping unless if there was any major problem occurred. Average crude palm oil (CPO) and POME production of KSLH is 20% and 70-80% of FFB processed respectively. The CPO produces is separated for further clarification and POME is first discharged to cooling pond because its discharge temperature from mill is 70-90 °C. Then this POME is transferred into de-oiling pond for further oil removal. After de-oiling, the POME is then transferred to the treatment ponds which cover an area of 25 hectors. The treatment ponds are consist of two mixing ponds where old and fresh POME is mixed, two facultative ponds, four anaerobic ponds and six algae ponds. The COD of POME ranges from 60-120 g/L. These ponds are only used to remove COD from POME and there is no biogas production facility on these ponds.

The process of oil extraction at KSLH can be briefly described as follows. The FFB are transported to mill each consists of hundreds of fruitlets each containing a nit surrounded by a bright orange pericarp which contains the palm oil. These FFB are sterilized with steam at a pressure of 3 bar. The steam condensate coming out of sterilized constitutes one of the two major sources of liquid effluent. The quantity produced varies from one mill to another, with a minimum of about one ton per ton of CPO produced. The fruitlets are stripped from bunches and further heated and crushed to a near homogenous mass into a tank before subject to a screw press. The screw press is used to separate spent mesocarp and nuts. The empty fruit bunches (EFB) are burnt in an incinerator and the resulting ashes used as fertilizer or being dumped off accordingly. The spent mesocarp and nuts sent for further processing from which another product, the palm kernel, is obtained. In this processing section, a hydrocyclone is used to separate the kernels from empty shells after cracking the nuts in a nut cracker. Approximately 0.2 ton of liquid effluent per ton of CPO produced is generated in this process.

The oil from screw press is diluted with hot water which assists in its separation in a settling tank. The floating or oil containing phase is further purified by centrifugation and subsequently sent to storage with the impurities. The bottom phase of settling tank is send to a sludge centrifuge where about 1.5 tons of sludge is produced per ton of CPO processed. The POME sample was stored at 4 °C before use. Different dilutions of POME were done by using tap water. Large and bulky materials in the raw POME samples were removed before dilution. Figures 3.2 and 3.3 showed the front internal view of the palm oil mill and the sampling point of raw POME respectively.

Figure 3.2: A view of Kilang Sawit Lepar Hiler Gambang, Kuantan

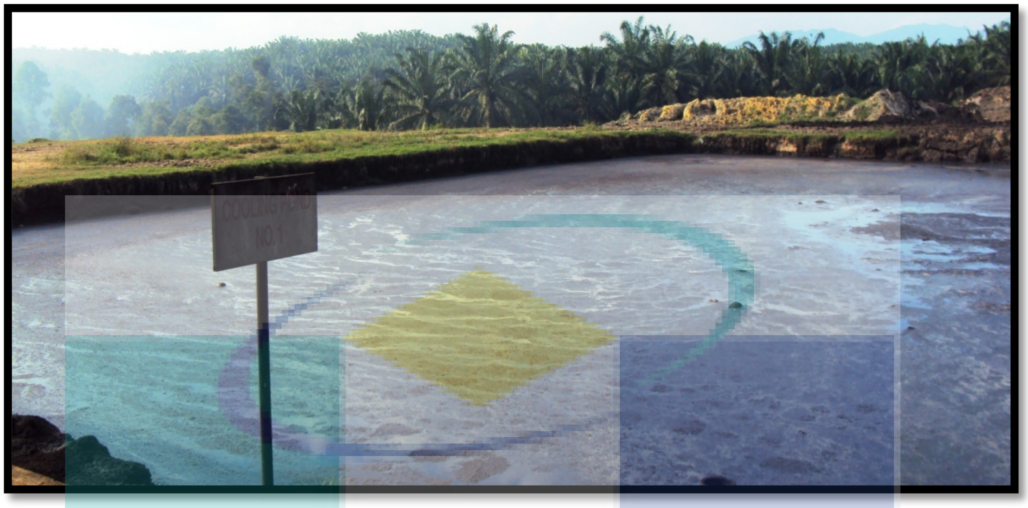

Figure 3.3: POME de-oiling pond for POME sample collection

Finally, the available dry-kiln CKD sample was collected from Pahang Cement which is located 30 Km east from Kuantan city. The plant has rotary kiln which operates in the temperature range between 1370–1480 ^{o}C. The CKD sample was washed with water and filtered to obtain approximate particle sizes. The properties of CKD were determined by various researchers are shown in the previous Chapter 1.

3.3 CHARACTERIZATION OF FOOD WASTE AND POME

The collected waste from waste was initially characterized by general segregation. The characterization was done by weight basis of the sample. All samples were taken from three different spots of final collection bins and the waste characteristics are shown in Table 3.1.

Meanwhile, the raw POME from the cooling pond was characterized by the parameters such as pH, BOD, COD, TS, SS, TN, TP and OG (oil and grease). Table 3.2 shows the physico-chemical properties of raw POME.

Table 3.1: Characterization of mixed food waste

Type of food waste	Weight (%)*
Rice	40±1.732
Fruits	34±2.646
Vegetables	6±1.732
Chicken meat	15±0.976
Egg and egg shells	1.6±0.10
Fish	3±0.05
Coconut	0.4±0.10
Total	100

*% weight mean±SD

Table 3.2: Physico-chemical properties of raw POME

Parameter*	Raw POME
pH	4.1-4.9
BOD_5	34-37
COD	67-70
TS	39-42
SS	18-21
TN	0.69-.80
TP	0.15-0.18
Oil and Grease	3-5

* All parameters are in g/L except pH

3.4 PREPARATION OF SAMPLES

The food waste sample was separated into three parts and subjected to filteration before liquidization was carried out as described in previous literature (Minowa et al., 1995; Sawayama et al., 1997). A galvanized steel autoclave of Hirayama company (model: INCLAVE HVE 50) was used for food waste liquidization. After purging with nitrogen gas, the autoclave was maintained with nitrogen at 2 MPa. The reaction was set to start by heating the autoclave up to 175 °C with an electric furnace. This temperature was maintained for an hour before the temperature was brought down to the room temperature.

After liquidization, the food waste was analyzed for physico-chemical properties. The biochemical oxygen demand (BOD_5), chemical oxygen demand (COD), total solids (TS), volatile suspended solids (VSS) and total suspended solids (SS) was determined for each sample and the results are shown in Table 3.3.

Table 3.3: Physico-chemical properties of LFW

Parameter*	**Concentration**
pH	4.5-4.9
BOD_5	18.8-20.4
COD	51.53.2
VSS	42-44.5
TSS	51-53.6
TKN (%TS)	11.15-11.9

*All values are in g/L except pH

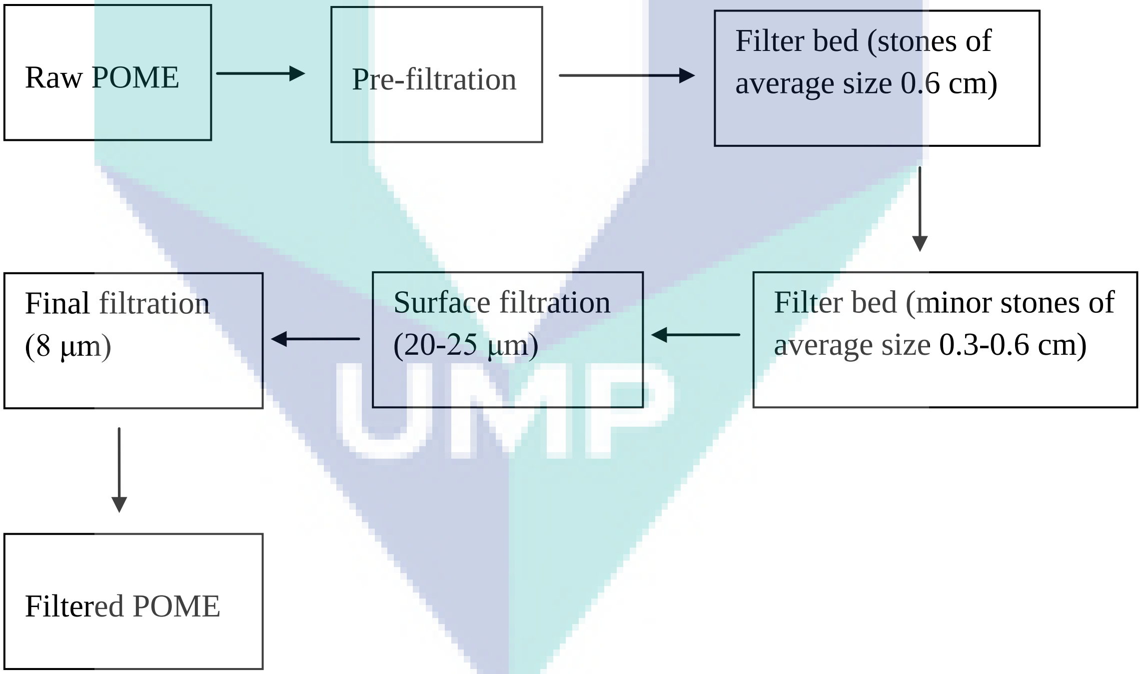

Figure 3.4: Sample preparation process of POME

Meanwhile for the POME sample, it was subjected to the simple filtration in order to remove the coarse solids. Then it was further passed through a filter bed, which was consisted of minor stones with average size of 0.6 cm. The collected filtrate was passed through another filter bed that was consisted of mixture of minor stones and sand (average diameter size of 300–600 µm) in the ratio of 1:2. The filtrate from second filter bed was then subjected to simple surface filtration. This filtration activity was done by using a Whatman No. 41 filter paper (20–25 µm) and finally a Whatman No. 40 filter paper (8 µm) under vacuum. Finally, the raw and filtered POME was subjected to pH, COD, BOD_5, TS, SS, TN, TP and OG determination. Table 3.4 shows the values obtained for each parameter determined for POME.

Table 3.4: Physico-chemical properties of raw and filtered POME

Parameter*	**Raw POME**^{a}	**Filtered POME**
pH	4.40	4
BOD	35	26
COD	68	55
TS	40	25
SS	19	14
TN	0.76	0.7
TP	0.17	0.16
Oil and Grease	4.0	1.8

* All parameters are in g/L except pH

^{a} Average values of 3 replicates

3.5 THE UASB REACTOR DESIGN

To date, researchers used different specification of UASB reactor for various wastewaters treatment. The UASB reactor specifications have also been shown in previous Chapter 2 (Table 2.9). In this study, the specifications of UASB reactor used were described by previous researchers (Lettinga et al., 1980, 1991; Borja and Banks, 1996; Nadais et al., 2005; Sari and Jukka, 2005; Revanuru and Mishra, 2008; Ma et al., 2009 and Sawajneh et al., 2010) and the specifications used are shown in Table 3.5.

Figure 3.5 shows an UASB reactor used in this study. The reactor was fabricated in university workshop. The dimension of UASB reactor used in this study was measured determining the reactor’s total and working volume. The reactor was made of cylindrical

PVC tubing with 4 mm wall thickness. Four sampling ports were installed at 15, 30, 45 and 60 cm from bottom of the reactor. These sampling ports were made by 6 mm of pneumatic fittings which were connected to manual valves in order to prevent leakage. An effluent post, an overflow pipe and baffles were installed 5 cm, 10 cm and 20 cm below the top of reactor respectively. The baffles angle of inclination was 45 degree with inner diameter 40% lesser than the actual reactor diameter.

Table 3.5: Specifications of UASB reactor

Specification	Measurement	Units	Remarks
Internal diameter, D_{in}	7.62	cm	
Height, h	85	cm	Total height of reactor
Total volume, V_r	3.87	liter	
Working volume, V_w	2.70	liter	
Cross-sectional area, A	45.58	cm^2	
Distance between each sampling port	15	cm	4 sampling ports
Diameter of sampling port	6	mm	
Diameter of other ports	8	mm	Effluent, over flow, feed inlet and gas outlet
Baffle inner diameter	4.60	cm	
Baffle wall inclination	45	degree	
Baffle thickness	3	cm	

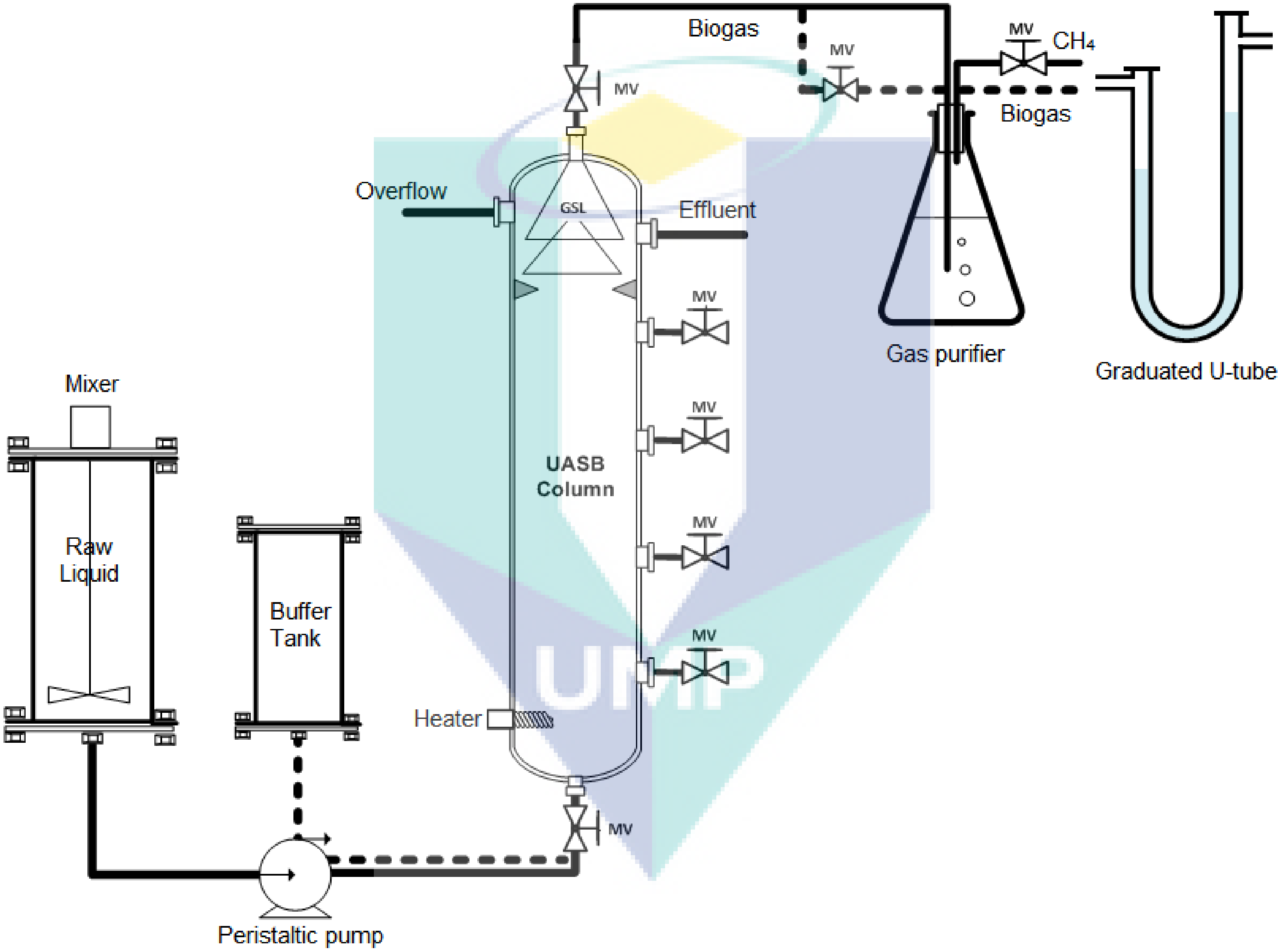

Figure 3.5: Schematic diagram of UASB reactor. MV stands for manual valve.

Meanwhile, a dual gas-solid-liquid (GSL) separator was installed at 4 cm above the baffles. The height of GSL was adjustable by means of reactor top cover. The Top cover with long male threads was screwed in the reactor column. For the purpose of drainage and feeding to maintain the reactor's pH, one particular feed inlet was made at the bottom of the reactor. The diameter of the reactor was enough to avoid from channeling.

3.6 SLUDGE SEEDING

3.6.1 Seeding for LFW

The reactor was seeded with anaerobic digested sewage sludge which was taken from an anaerobic digester of the municipal sewage treatment plant that situated in Kuantan.

The digested sewage sludge was first sieved (< 2mm) in order to remove any bigger particles before subjected to the reactor. A 50% of sludge which had suspended solids composed of 3.09 g-TSS/L and 2.09 g-VSS/L was filled into the reactor. The microbial activity of the sludge seed was tested by adding 5 ml of sledge mixture to 50 ml sucrose and acetate (Alper et al., 2005). The gas production was analyzed after 24 hours. The sledge seed used was anaerobically active and produced CH_4, H_2S and N_2.

3.6.2 Seed Sludge Formation for POME

As the purpose of this study was to evaluate the feasibility of CKD for potential treatment of POME, so it was necessary to use activated sludge for proper monitoring of the pH variation with CKD. A stock solution was prepared for sludge activation. Stock solution consists of macro and micro nutrients which help the bacteria to activate for quickly. The stock solution was prepared by the following macro and micro nutrients for pre-treated POME sample (values are in g/L): NH_4CL, 174; KH_2PO_4, 28.3; $(NH_4)_2 SO_4$, 28.3; $MgCl_2$, 25; KCL, 45; yeast extract, 3; $FeCl_2.6H_2O$, 2; H_3BO_3, 0.05; $ZnCl_2$, 0.05; $CuCl_2.2H_2O$, 0.038. The samples after nutrient addition were kept at room temperature for

20 days. The total volatile solids concentration of seeded sludge was 10 g/L. Nitrogen and phosphorus were added in the form of NH_4Cl and KH_2PO_4 to give a COD:N:P ratio of 250:5:1 for successful activation of seed sludge.

3.7 THE CKD SLACKING PROCESS

A bench-scale experiment to model the lime slaking process by using a standard jar test apparatus was conducted in this study. Several dosages of each CKD sample, normalized to its CaO contents, were added to the jars with 750 mL of the room temperature water (approximately 27 °C). Water was added slowly at a constant rate to avoid the CKD particles drown. Drowning occurs when water is added too quickly thus creates an outer shell of hydrated lime $Ca(OH)_2$ that slows or prevents the interior of the particles from hydrating (Boynton, 1980). The mixer speed was adjusted to 10 rpm for slow and steady mixing. Each sample was then rapid mixed at 200 rpm for 30 min, with 250 mL of cold water (4 ◦C) added at the end and mixed for 1 min to ensure maximum dissolution of $Ca(OH)_2$.

The CKD dosages evaluated were 8.06, 13.14, 18.76, 22.46, 26.7, and 32.96 g-CaO/L. After mixing, samples of the slurry were taken and analyzed for pH and conductivity. Separate samples were taken and passed through a 0.45 µm filter (Whatman) and the same analyses were performed on the filtrate. The slaked solutions were filtered through a 1.5 µm Whatman filter in order to determine total suspended solids (TSS) concentrations of the samples. The filters were weighed before and after filtration and these measurements were compared to the mass of CKD that had been added at the beginning of the jar tests in order to estimate the percent dissolution of the CKD material.

3.8 EXPERIMENTAL SETUP

3.8.1 Liquidized Food Waste (LFW) Experiment

The treatment of LFW was carried out by using UASB reactor (Figure 3.4). Thermophilic condition was adapted by gradual increase in the temperature from 30-55 °C. The feed was introduced by peristaltic pump. The reactor pH was maintained through separate buffer solution dosing tank. The diluted liquid phase of LFW was fed into the reactor with stepwise increase in OLR from 2.0-12.5 kg-COD/m^3.day and corresponding HRT was 10-4 days as shown in Table 3.6. The OLR was of the influent was calculated from VSS of substrate multiplied by the F/M ratio (0.10-0.60). The reactor was monitored daily for flow rate, TOC, COD, SS, and VSS, volatile fatty acids (VFAs) while biogas and CH_4 were measured after every one day and temperature and pH were monitored quarterly a day.

Table 3.6: Experimental setup for the anaerobic treatment of LFW

Operation Run	Duration per run (days)	OLR (kg-COD/m^3.day)	HRT (days)	Temperature (°C)
1	6	2	10	30
2	10	3	10	33
3	8	4	10	37
4	6	5	10	50
5	14	6.25	8	55
6	12	8.33	6	55
7	10	10.00	5	55
8	6	12.50	4	55

3.8.2 The Palm Oil Mill Effluent Experiment

The experiment was performed by upflow anaerobic sludge blanket (UASB) reactor. The reactor was continuously operated under mesophilic conditions (27-32 °C). The feed was introduced through bottom of the reactor by peristaltic pump with a flow rate of 0.52 L/day at a fix HRT of 4 days.

The UASB reactor was inoculated with 350 mL seed sludge and acclimatization of sludge with POME was done by daily bench fed of diluted sludge (5 g-COD/L) for five days. The average volatile suspended solids (VSS) of the sludge after 5 days bench fed were 11.3 g/L. Continuous feeding was started with an initial organic loading rate (OLR) of 1.5 kg-COD/m^3.day and HRT of 4 days. The HRT was kept constant throughout the experimental period. The POME was fed in the reactor with stepwise increase in OLR from 1.5-4.0 kg-COD/m^3.day. The OLR of the influent was calculated from VSS of substrate multiplied by the F/M ratio (0.10-0.60). The detail of experimental setup is shown in Table 3.7. The reactor was monitored daily for volatile fatty acids, alkalinity, COD; SS, pH and temperature while biogas and methane production were measured three times a week.

Table 3.7: Experimental setup for the anaerobic treatment of POME

Operation Run	Duration per run (days)	OLR (kg-COD/m^3.day)	HRT (days)	Temperature (°C)
1	12	1.5	4	30±2
2	12	2.5	4	30±2
3	18	4	4	30±2

3.9 ANALYTICAL METHODS

The COD, BOD, TOC, TN, TKN, TP, VFA, total solids, total suspended solids and volatile suspended solids were measured by the Standard Methods of Waste and Wastewaters APHA, 1995).

The chemical oxygen demand was measured by direct digestion method, using HACH apparatus LR (3–150 mg/L COD); HR (20–1500 mg/L COD) and HR plus (20–15000 mg/L COD and above). The COD measurement was always carried out by diluting the original sample to meet the vial results. The vials (blank and sample) after adding the sample were placed in COD digester reactor for 2 hours at 150 ^{o}C which were kept remained in reactor until the reactor temperature was lower down to 120 ^{o}C. After removing vials from reactor, these were allowed to cool down further at room temperature for accurate results. The program 435HR COD was used accordingly for results.

Total organic carbon was measure by direct method in HACH. A 10 mL sample with 0.4 ml buffer solution at pH 2 were added to a flask and kept for moderate stirring up to 10 minutes. After that TOC reagent powder pillow was added to acid digestion vial along with 0.3 ml of organic free water and 0.3 mL of sample. The blue HR/MR indicator rinsed with de-ionized water was placed in vial and kept in COD reactor for 2 hours at 103 ^{o}C. After removing the vial from reactor, it was placed in tube holder till its room temperature. A blank vial was also prepared by the same method as used for original sample. The HACH was calibrated by using program 426 HR (100-700 mg/L) TOC and results were taken by placing original sample vial. The obtained results were obtained by multiplying the original results with the dilution factor.

Total nitrogen was measured by persulfate digestion method using HACH program 350N, LR (0.5-25 mg/L) TNT. The nitrogen hydroxide reagent vials with added total nitrogen persulfate pillow and 2 mL sample were kept in COD reactor for 30 minutes at 105 ^{o}C. Then total nitrogen reagent pillows A, B and C were added by giving them the reaction time of 3, 2 and 5 minutes respectively in HACH apparatus. The blank was also run with same manner as used for original sample for apparatus calibration. The obtained results were obtained by multiplying the original results with the dilution factor.

Total Kjeldahl nitrogen (TKN) was also measured by Nessler method where digestion of sample is carried out by means of digesdahl digestion apparatus. After digestion 10 mL of sample was taken in a 25 mL graduated cylinder and same amount of de-ionized water was taken for blank. Then one drop of TKN indicator was added to each cylinder. After that drops of 8.0 N KOH were added until a flash of blue color is appeared. For taking permanent blue color, a drop of 1.0 N KOH was added and mixed by gentle shaking of cylinder then other drop was added and mixed until permanent blue color appeared. The cylinders were then filled up 20 mL mark with de-ionized water and three drops of mineral stabilizer were added to each cylinder. The three drops of polyvinyl alcohol dispersing agent were also added and mixed in each cylinder and filled with de-ionized water up to 25 mL mark. 1.0 mL of Nessler reagent was pipette in each cylinder and cylinders were inverted repeatedly. The cylinders were kept consecutively in HACH apparatus pre selected program 399 nitrogen, TKN and two minute reaction time was given to each of the cylinder. After that the 10 mL volume from each of the cylinder was poured in separate square sample cell. The blank square sample cell was placed in apparatus for zero and then original sample was placed for results. The measured amount was multiplied with a factor 75 and then divided with volume of digest for analysis and volume taken in square sample cell.

Total phosphorus was measured by molybdovanadate method with acid persulfate digestion. Two total phosphorus test 'N tube vials used and each vial was added with 5 mL of de-ionized water and sample solution respectively and potassium persulfate powder pillows were added to each vial. The de-ionized water was used for blank. The vials were then placed in reactor preheated at 150 ^{o}C for 30 minutes. After digestion the vials were placed in vial holder and when these vials cooled down at room temperature, 2 mL of 1.54 N sodium hydroxide solution was added to each vial. Then 0.5 mL of molybdovanadate reagent was added to each vial. The vials were then placed in HACH apparatus for seven minutes reaction period. The total phosphorus was calculated by selecting a 542 HR P total program. The results obtained were in mg/L PO_4^{3-} and results for only phosphorous (P) were obtained by multiplying a factor of 0.3261 to the original results. The obtained results were obtained by multiplying the original results with the dilution factor.

Volatile fatty acids (VFAs) were measured by esterification method using HACH apparatus. A 25 mL sample was centrifuged and supernatant was collected in to separate sampling bottle. The 0.5 mL of this supernatant was taken into another 25 mL sampling bottle. Ethylene glycol 1.5 mL was added to the sample and swirled for mixing. The 0.2 mL of 19.2 N sulfuric acid standard solution was added and mixed prior to putting the sample bottle into water bath for boiling. The HACH program 770 volatile acids was selected and boiled sample was placed in it thus giving a reaction time of three minutes. After beep the sample bottle made cool down until 25 ^{o}C and 0.5 mL of hydroxylamine hydrochloride solution was added and mixed. Then 2 mL of 4.5 N sodium hydroxide solution was pipette into the sample and 10 mL of ferric chloride solution as also added and mixed. 10 mL of deionized water was added to the sample and well mixed. The 10 mL of sample was taken into clean dry square sample cell. Similarly whole procedure was also adapted in order to prepare the blank sample. This square sample cell was placed in the apparatus for giving a 3 minute reaction time. The volatile acids were noted in mg-acetic acid/L by placing the sample cell into apparatus.

For total solids, the porcelain dishes were washed and oven dried and cooled by means of desiccators for accurate weighing. Then sample was weighted along with empty porcelain dish prior to follow the further procedure of dehydrating and drying. The water bath was used for dehydration of the solution in dish for 30 minutes at 80 ^{o}C and it was placed in oven at 104 ^{o}C for 3-4 hours. The difference in result was divided by the sample volume used and then multiplied with factor to obtain the results in mg/L.

The total suspended solids were measure by centrifuge method and the centrifuge was run at 1400-2000 rpm for 5 minutes and the process was repeated again for better solids settling in the centrifuge tube. The centrifuged contents were transferred in porcelain dish and placed in oven for 2-4 hours at 104 ^{o}C. The weight difference was then divided by the sample volume and multiplied with the unit conversion factor to obtain the results in mg/L.

The TSS residue obtained after oven drying was then processed for ignition at 550 ^{o}C to obtain volatile suspended solids concentration. The difference in weight loosed in ignition and before ignition was measured and multiplied by the unit conversion factor to obtain the results in mg/L.

The pH was measured using HACH pH meter (sension 1) and conductivity by conductivity meter (HACH). Biochemical oxygen demand (BOD_5) was measured by dilution method using BOD buffer pillows. The difference in dissolved oxygen before and after incubation was calculated for BOD measurement. The dissolved oxygen was measured through DO meter (YSI 5100). Oil and grease were determined according to standard methods (Clescerl et al., 1998). Alkalinity was measured by the direct titration method (Jenkins et al., 1983).

Gas production was measured by liquid displacement method. An inverted burette was connected to one end of a flexible tube and the other end of flexible tube was then jointed with another burette and this full set of U-tube was filled with water up to a graduated level. The gas was entered from one side of the burette and gas volume was measure by the liquid displacement to other side by pressure exerted from the gas. The change in volume in burette was noted for total gas volume produces at that time. Methane gas was calculated by the same method but the gas was first passed through soda lime in order to remove carbon dioxide and remaining methane gas was then subjected to pass through U-tube water displacement set for measurement. The methane gas contents were measured once a day during experiment. The biogas production by displacement method was calculated as follow,

$$V = V_A - V_B \tag{3.1}$$

Where,

V = Total volume of gas collected

V_A = Volume of water in tube before displacement

V_B = Volume of water in tube after displacement

3.10 STATISTICAL ANALYSIS

The data was analyzed for three replicates by using Microsoft Excel 2007. All necessary statistical data was derived in this software. The mean, standard deviation and standard error results are were calculated from replicates and applied to each figure and table values. Correlation (r) was calculated to analyze the effect of one parameter on other.

CHAPTER 4

RESULTS AND DISCUSSION

4.1 INTRODUCTION

This chapter describes the experimental results and discussion of liquidized food waste (LFW), and palm oil mill effluent (POME) treatment with cement kiln dust (CKD) using UASB reactor. In the LFW part, results of total organic carbon, effect of temperature on VFA and pH, effect of organic loading rate, solids removal and biogas production are described. In the POME treatment part, the results are described in two main sections viz. reactor stability and reactor performance. The reactor stability shows the effect of CKD on reactor pH, VFA and alkalinity, flow rate, upflow velocity, HRT and food to micro-organism ratio. The reactor performance in terms of COD removal, SS removal and biogas production has been described.

4.2 THE LIQUIDDIZED FOOD WASTE

4.2.1 Total Organic Carbon

Figure 4.1 shows the variation in TOC removal efficiency with influent TOC concentrations over the experimental period. The temperature range during the experiment has also been shown. During first week, at the TOC concentration of 21 g/L and 14 g/L, the TOC removal efficiencies were 32% and 43% respectively. Later on, when TOC concentration decreased, its removal efficiency increased and reached a maximum value of

79% till 51 days. As TOC concentration increased from 7 g/L to 18 g/L, the removal efficiency reduced to 42%, hence showed a negative correlation factor (r=-0.92). The values of standard deviation and error for influent TOC and TOC removal has been shown in Appendix A1.

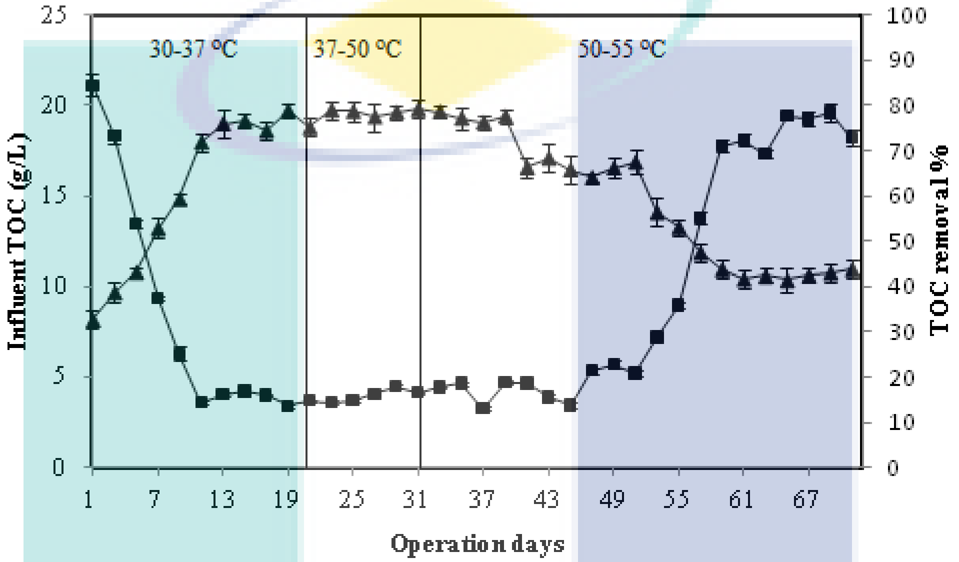

Figure 4.1: Influent TOC concentrations and removal efficiency during anaerobic treatment of LFW using UASBR; ■ Influent TOC, ▲ TOC removal, (values in mean ± SD)

Thermodynamically high TOC contents in LFW might be due to solubilization of waste. This takes more time at start up, which in fact, reduces the conversion of long chain polymers into short chain polymers and simple monomers, although the initial organic loading rate was limited to 2.0-3.0 g-COD/L.day at 10 days HRT. So, it might be suggested that, lower the total organic contents in the reactor, higher will be the TOC removal efficiency and vice versa.

4.2.2 Chemical Oxygen Demand Removal

The COD removal efficiency with organic loading rate is given in Figure 4.2. During the first week of experiment, COD concentration was 20 g/L. The lower organic loading rates (2.0-3.0 g-COD/L.day) were, in fact, due to lower COD concentrations at early stages in LFW treatment. COD concentration was increased stepwise by increasing OLR and reducing HRT. COD removal efficiency was low in first week (40-55%) but it was recovered during second week (59-68%), although the OLR was increased from 2.0-5.0 g-COD/L.day.

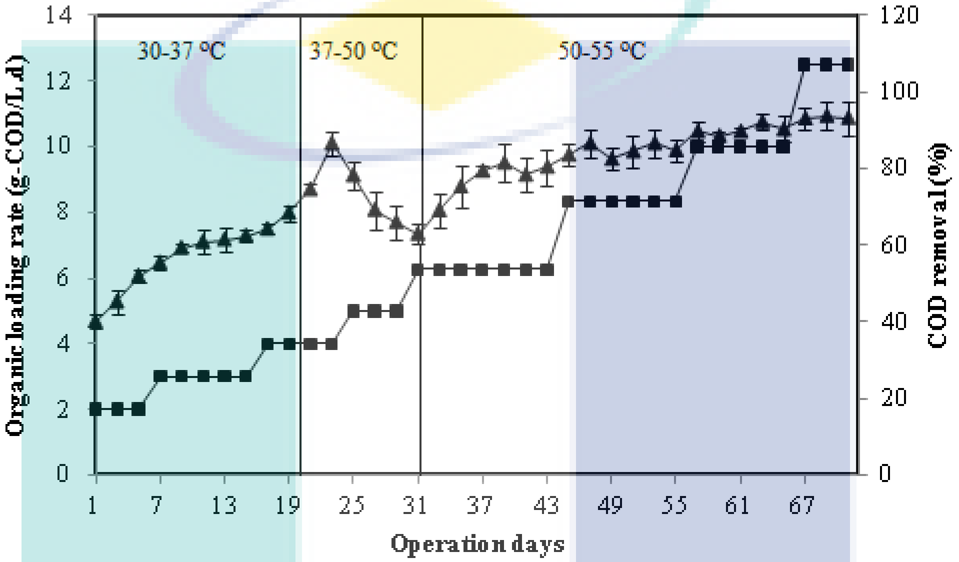

Figure 4.2: Effect of OLR on COD removal efficiency during anaerobic treatment of LFW; ■ OLR, ▲ COD removal, (values in mean ± SD)

During fourth and fifth week, the COD removal efficiency was decreased from 68% to 62%.This was happened due to sudden shift in temperature from 37 ^{o}C to 50 ^{o}C which caused a shock in substrate biomass but the system was maintained its stability after that. As organic loading rate was further increased, COD removal efficiency reached to 94% at an organic loading rate of 12.50 g-COD/L.day. Choorit and Wisarnwan (2007) reported 71.10% and 70.32% COD reduction under mesophilic (37 ^{o}C) and thermophilic (55 ^{o}C) conditions respectively, treating palm oil mill effluent in continuous stirrer tank reactor. Rittmann and McCarty (2001) investigated that the optimal growth rate of bacterial strain occurred over a limited temperature range and when this temperature range exceeded, growth rate reduced rapidly. The overall COD removal efficiency of the reactor was found 75%. Appendix A2 shows the standard deviation and errors for three replicates of removal efficiencies.

A comparison of the results achieved with those reported for other anaerobic treatment systems and for similar wastewaters illustrates that our results are satisfactory, though an actual comparison between various data sets can be based on experiments where the same wastewater and reactors of comparable size with the same operation temperature are used.

4.2.3 Effect of Organic Loading Rate

Organic loading rate affects on several parameters like COD, biogas, methane, volatile fatty acids and reactor pH. Stepwise increase in OLR from 2.0-12.50 g-COD/L.day resulted in satisfactory COD removal efficiency with a strong correlation (r=0.78) which indicates that by increasing the organic loading rate, COD removal percentage will be higher as shown in Figure 4.2 above.

The appropriate design of organic loading rate is necessary for better reactor performance and process stability. In this study, the organic loading rate was increased gradually by reducing HRT from 10 to 4 days while influent COD concentrations were kept constant each run (Table 4.1). Movahedyan et al. (2007) maintained OLR up to 1.5 g-COD/L.day for one week and then increased up to 10 g-COD/L.day for 40 days. Nevertheless, biogas and methane production was low at lower organic loading rates. Michaud et al. (2002) and Rincon et al. (2006) reported that lower organic loading rates resulted in lower COD removal and biogas yield. As strong wastewater containing high organic load, significant amounts of fatty acids can develop from partial degradation of substrate therefore can inhibit the methanogenic population in the reactor (Uyanik et al., 2002).

The high OLR reduces HRT at same influent COD concentration which causes high flow rate of influent stream. This high flow rate causes channeling through the biomass bed, resulting in poor substrate-biomass contact and minimum degradation of the incoming COD. This provides additional support to the earlier assumptions that under plug-flow conditions, incoming substrate remains in the reactor, for one retention time, allowing maximum time for conversion. The high substrate concentration results due to the lack of liquid distribution which may inhibit bacterial activity (Sallis and Uyanik, 2003). Kalyuzhnyi et al. (1998) worked on UASB reactor for chip-processing industry wastewater and achieved treatment efficiency between 63-75% at an OLR of 14 kg-COD/m^3.day. Grover et al. (1999) used anaerobic baffled reactor, treating pulp and paper liquors at 35 ^{o}C and observed maximum COD reduction of 60% at an organic loading rate of 5 kg-COD/m^3.day and 2 days HRT.

Table 4.1: Different operating and performance parameters for during start-up of UASBR (mean values ± SD, n=3)

Organic loading rate (g-COD/L.day)	TSS$_{in}$	TSS removal (%)	Reactor VSS (g/L)	F/M (kg-COD/kg-VSS.day)
2	5.15	71±4.3	5.7±0.41	0.35±0.02
3	5.15	70±1.0	13.0±0.63	0.23±0.01
4	5.15	67±3.3	13.8±0.64	0.29±0.01
5	5.15	60±3.5	11.1±0.29	0.45±0.01
6.25	6.45	66±4.2	16.9±0.59	0.37±0.01
8.33	8.58	77±3.9	20.3±0.69	0.41±0.01
10	10.3	80±3.9	20.4±0.53	0.49±0.01
12.50	12.88	84±4.7	26.6±0.66	0.47±0.01

4.2.4 Solids Removal

High rate anaerobic treatments have been proposed for wastewater having soluble organic pollutants. The LFW, having total and volatile suspended solids respectively 51.5 g/L and 43.5 g/L, was directly fed to the reactor without pretreatment. For each OLR, the influent COD was being decreased by dilution with water so the solids in influent stream also decreased as shown in Table 4.1. The reactor VSS were approximately ranged from 5-11 g/L at OLR 2-5 g-COD/L.day. But, VSS concentration was increase due to micro-organisms development and increasing OLR. The reactor VSS were observed as 26.6 g/L up to a maximum OLR of 12.5 g-COD/L.day. The change in F/M ration and TSS removal with increasing OLR has also been shown in Table 4.1.

At start up, effluent TSS (Table 4.2) were low (1.50-1.72 g/L) at lower organic loading rates (2-5 g-COD/L.day). As organic loading increased, effluent TSS were also increased upto 2.02 g/L, thus, showing that OLR affects the TSS removal with correlation (r=0.67). The effluent VSS removal showed a similar tendency as TSS, where effluent VSS were initially low (1.20-1.38 g/L) at start-up which further increased to 1.58 g/L with increasing organic loading rate and showed a correlation (r=0.73).

Table 4.2: Experimental results obtained after each run during start-up of UASBR (effluent mean values ± SD, n=3)

Run	Duration per run (days)	Organic loading rate (g-COD/L.day)	HRT (days)	COD in (g/L)	Effluent TSS (g/L)	Effluent VSS (g/L)	Effluent TOC (g/L)
1	6	2	10	20	1.50±0.17	1.20±0.12	6.83±0.08
2	10	3	10	30	1.54±0.14	1.27±0.10	3.51±0.11
3	8	4	10	40	1.72±0.13	1.38±0.09	2.82±0.10
4	6	5	10	50	2.06±0.15	1.65±0.13	3.20±0.12
5	14	6.25	8	50	2.21±0.22	1.57±0.18	3.17±0.08
6	12	8.33	6	50	2.01±0.12	1.62±0.16	3.64±0.18
7	10	10.00	5	50	2.01±0.09	1.61±0.05	7.46±0.15
8	6	12.50	4	50	2.02±0.13	1.58±0.17	8.21±0.21

The total and volatile suspended solids in the effluent are shown in Table 4.2. The effluent TSS was in the range of 1.50-2.21 g/L and effluent VSS were 1.20-1.65 g/L. Thus, LFW having high contents of solids can be suitable for UASB treatment process. Moreover, high rate anaerobic treatment technologies are projected for wastewater, containing soluble organic matter. As organic hydrolysis step is not required in such cases were faster conversion rates to methane can be obtained (Harper and Pohland, 1986).

However, HRT could not be decreased beyond 4 days and organic loading could not be increased beyond 12.5 g-COD/L.day in this study. This may be due to the fact that hydrolysis of the particulate organics present in the wastewater took longer time. The results support the views of Lettinga et al. (1984) where high levels of suspended solids can be treated anaerobically but only at reduced organic loads. However, the production rate of SS substrates in the anaerobic reactor is very low as compared to aerobic process (Speece, 1996).

4.2.5 Biogas and Methane Production

Figure 4.3 shows the change in biogas and methane production along with organic loading rate. At an OLR of 2.0 g-COD/L.day, biogas and methane production were 0.49 L/g-$COD_{removed}$ and 0.29 L/g-$COD_{removed}$ respectively. The biogas production gradually increased with increasing organic loading rate and the average biogas production was 1.16 and 1.35 L/g-$COD_{removed}$ at loading rates of 6.25 and 12.5 g-COD/L.day respectively.

The methane yield was between 0.29 to 0.91 L/g-$COD_{removed}$ at an OLR of 2.0-12.50 g-COD/L.day respectively. Similar results has given by Ross et al. (1992) that gas can be produced from the degradation of sewage sludge at a HRT of 20 days and at a temperature of 35 °C. Chiang and Dague (1992) obtained a constant methane yield between 0.3 and 0.34 L-CH_4/g-$COD_{removed}$ at a low OLR (1 g-COD/L.day), but it was between 0.26 and 0.29 at a higher OLR (12 g-COD/L.day). It appears that many data are in agreement with the observation made by Elmaleh et al. (1984) which described the methane yield as a decreasing function of the OLR.

A strong correlation (r=0.82) was found between organic loading rate and biogas production during LFW anaerobic treatment in UASB. Results indicate that high organic loading rates result in more biogas production as was also proposed by Angelidaki et al. (2006) but the reduction in biogas is caused due to the sudden change in temperature and VFA accumulation in the reactor. The values of standard deviation and error of three replicates for biogas production has been shown in Appendix A3.

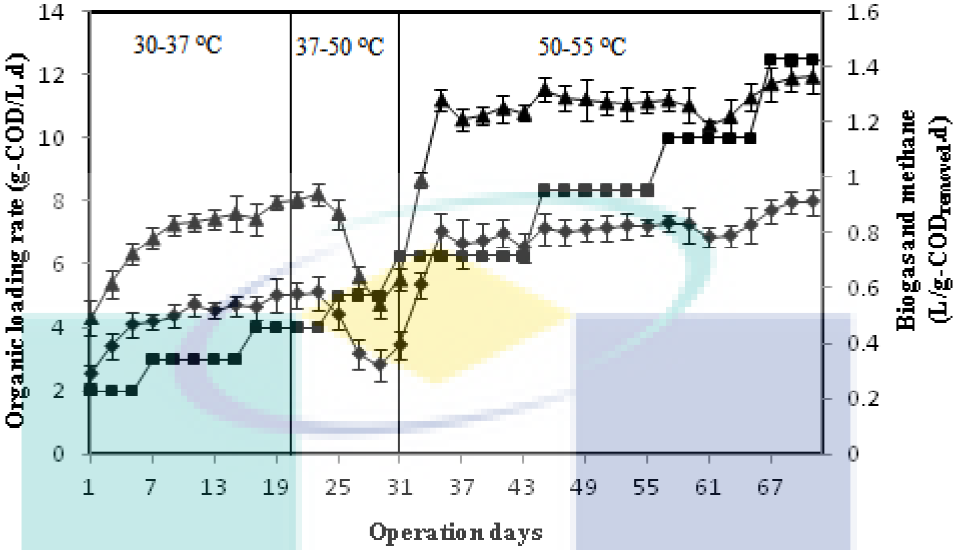

Figure 4.3: Effect of OLR on biogas and methane production during anaerobic treatment of LFW; ■ OLR, ▲ Biogas, ♦ Methane, (values in mean ± SD)

Biogas production (0.24-0.89 L/g-COD$_{removed}$.day) at mesophilic condition was low, but it might be considered that at the same time, the COD concentration in the wastewater was also less (2-5 g-COD/L.day). So, the decrease in the biogas production is mainly due to temperature and VFA accumulation. Nevertheless, thermophilic condition supported to high biogas production, as incoming COD concentrations and pH were monitored continuously. Biogas production is lower at temperature, less than 15 ^{o}C and higher than 65 ^{o}C which inhibited microbial activity (Komemoto et al., 2009). Effect of temperature on the performance of anaerobic digestion was investigated. Yu et al. (2002a) found that substrate degradation rate and biogas production rate at 55 ^{o}C was higher than operation at 37 ^{o}C. Studies have reported that thermophilic digesters are able to tolerate higher OLRs and operate at shorter HRT while producing more biogas (Ahn and Forster, 2002; Kim et al. 2006 and Yilmaz et al., 2008). However, failure to control temperature increase can result in biomass washout (Lau and Fang, 1997) with accumulation of volatile fatty acid due to inhibition of methanogenesis.

4.2.6 Effect of Temperature on VFA and pH

The production of VFAs under mesophilic conditions was low as shown in Figure 4.4. A strong correlation (r=0.95) was found between temperature and volatile fatty acids. At temperatures of 30-37 ^{o}C, volatile fatty acids concentration was 3.6-4.0 g-acetic acid/L. In contrast, as temperature was shifted from 37-50 ^{o}C, VFA concentration in the reactor increased up to 9.0 g-acetic acid/L.

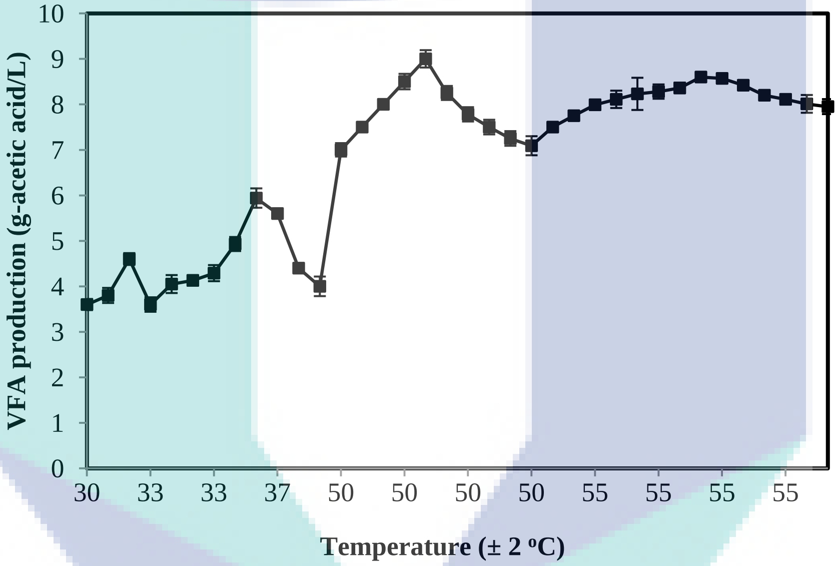

Figure 4.4: Effect of temperature on VFA concentrations during anaerobic treatment of LFW using UASBR, (values in mean ± SD)

Meanwhile, biogas production was reduced from 0.95 to 0.54 g-acetic acid/L (Figure 4.3) and pH was also dropped to 5.96 (Figure 4.5) which indicated that sudden change in temperature affects the reactor performance in terms of VFA accumulation, biogas production and drop in pH value of food substrate. The reactor still maintained its stability at a high temperature of 55 ^{o}C and the pH reached its optimal level (7.5±0.5). Later on, VFA concentrations increased gradually with increasing OLR but no reduction in biogas production was observed. Similar findings were reported by Wong et al. (2008) wherein they used different concentrations of VFAs and observed the biogas and methane production. The accumulation of volatile fatty acid reduced the biogas production and sometimes, caused system deterioration. A problem exists where the acid formers overproduce organic acids which lower down the pH below the optimum value (< 6.5), where methane bacteria cannot function. At this stage, methane formation stopped and lead to a buildup of sludge with a low pH.

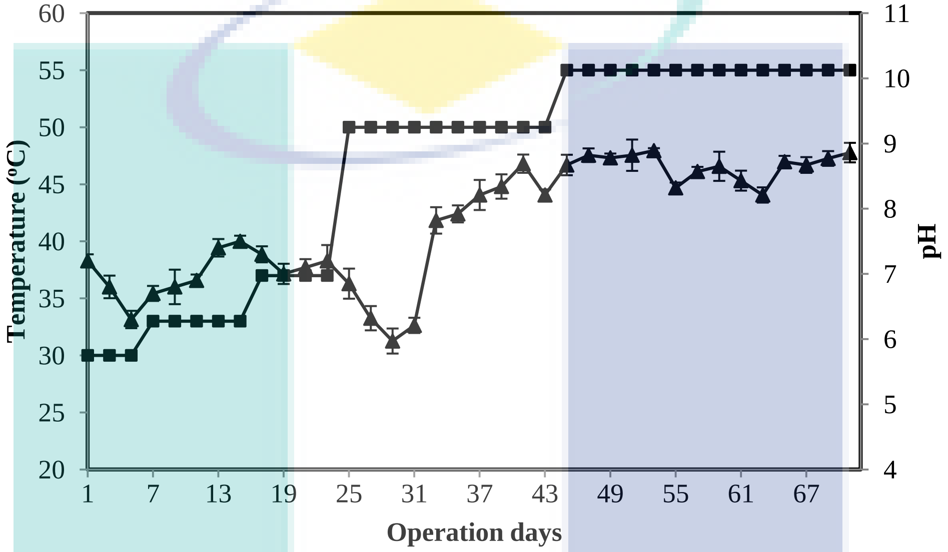

Figure 4.5: Effect of temperature on pH during anaerobic treatment of LFW using UASBR; ■ Temperature, ▲ pH, (values in mean ± SD)

Moreover, reduction in pH is also due to the increase in temperature as shown in Figure 4.5 where, pH value of the LFW dropped suddenly when temperature was raised from 37-50 °C, which suggested that an optimum value of pH can also improve the reactor performance in terms of low VFA accumulation and high biogas production. Ivan and Herbert (1998) investigated that the temperature shock causes severe biomass washout, lowering of the pH and accumulation of fatty acids. However, reactor stability was recovered within eight days, after the pH was rectified by the addition of alkaline solution. The three replicate standard deviation and errors for VFA and pH has been expressed in Appendix A4 and A5 respectively.

At high temperatures, production of volatile fatty acid is higher compared to mesophilic temperature range (Yu et al., 2002a). Many researchers prefer to have digesters operating in mesophilic temperature due to for better process stability. Nevertheless, investigation on digester stability by Kim et al. (2002) proved that disadvantages of thermophilic digesters can be resolved by keeping microbial consortia in close monitoring. The phenomenon of high VFA production is consistent with the reports of Kim et al. (2003) and Shirai (2004) that fatty acid bacterium has the ability to adapt to a wide range of temperatures (mesophilic to thermophilic range).

Full-scale thermophilic (50–55 ^{o}C) anaerobic digestion of wastewater from an alcohol distillery was reported by Vlissidis and Zouboulis (1993). More than 60% removal of COD was achieved with 76% of biogas comprising of methane thus making it a valuable fuel. Stevens and Schulte (1979) studied the effect of the temperature at solids retention times of between 6 and 55 days at organic volumetric loading rates of between 0.61 and 4.81 kg-VS/m^3.day, in a complete mixed anaerobic digester. They concluded that at organic rates in the range of 0.61–1.80 kg-VS/m^3.day and temperatures lower than 25 ^{o}C, the operation proceeded satisfactorily.

4.3 TREATMENT OF POME WITH CKD

4.3.1 The Quality of Hydrated CKD Slurry

The parameters pH and conductivity were used to characterize the reactivity of the CKD when slacked in water. The pH was chosen as major indicator of CKD reactivity for this experiment because of the importance of this parameter for neutralizing the acidic wastewater streams. Conductivity is generally a measure of the ionic strength of a solution, or how many free ions are in solution. Robinson and Burnham (2001) found the dissolution rates of hydrated limes that conductivity increased close to linearly with lime dosage.

Conductivity has also been directly related to the concentration of $Ca(OH)_2$ in a pure solution, though it also varies with temperature (Boynton, 1980). Shown in Figure 4.6 are the pH results for the filtered samples and in Figure 4.7 the conductivity results for the filtered samples. The data in Figures 4.6 and 4.7 represent the average outcome of three replicates, with error bars representing standard deviation less than one.

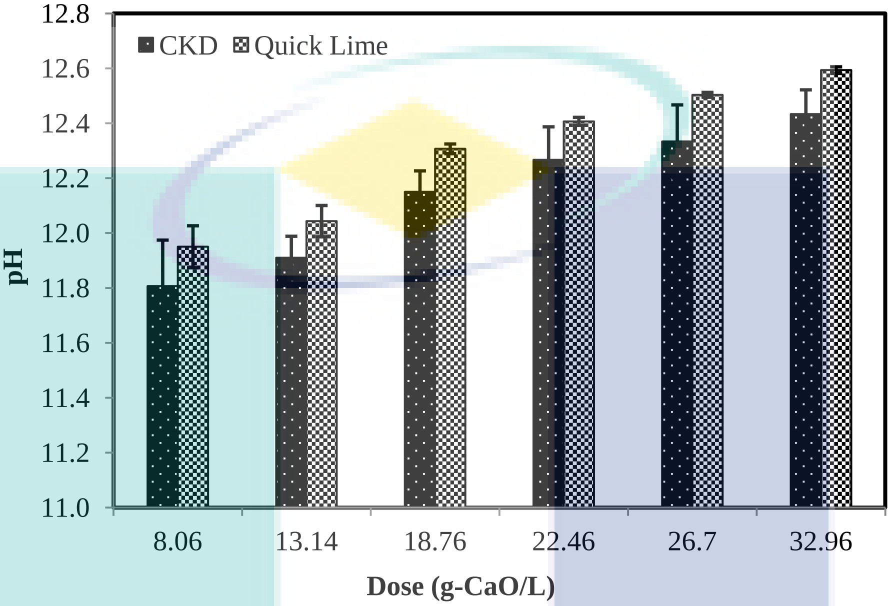

Figure 4.6: The pH results of filtered solution after slacking process, (values in mean ± SD)

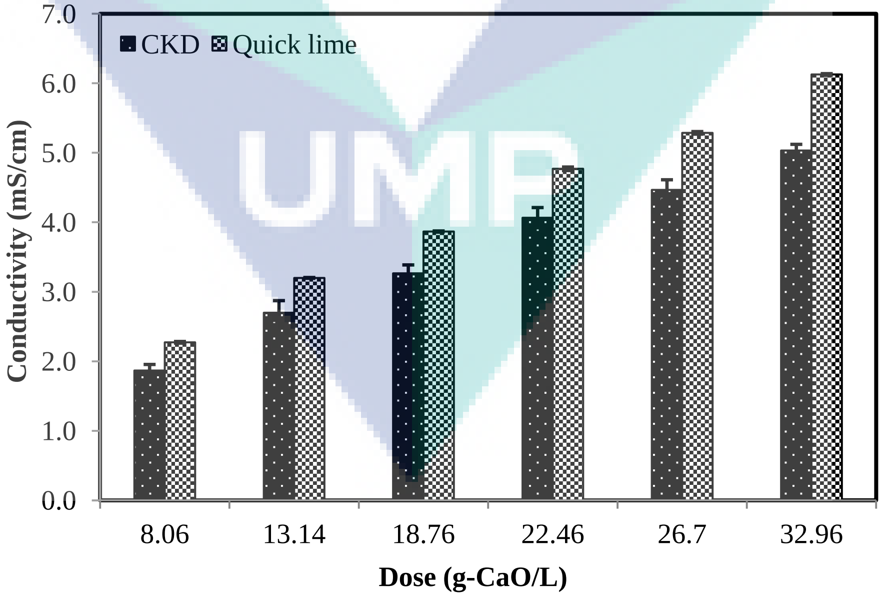

Figure 4.7: The conductivity results of filtered solution after slacking process, (values in mean ± SD)

4.3.2 Acid Neutralization Test

Acid neutralization tests were performed as part of the methods of this study by using each of the CKD samples to determine the feasibility of using CKD with different concentrations to treat acidic wastewater. Figure 4.8 shows the results of the acid neutralization trials, in g-CKD/L-H_2SO_4 versus pH achieved after 30 min. The results showed that CKD-4 and CKD-5, the two CKD samples with the highest concentration and those that responded best in the slacking process, achieved the highest pH. Specifically, CKD-4 required 36.7 g to achieve a pH of 9, while CKD-6 required 32.96 g. However, all CKD samples achieved pH values greater than 7, at varying doses of CKD for each. These results show potential application for even CKD samples to be effective at neutralizing acidic wastewater. The standard deviation and error values for pH, conductivity and neutralization test has been shown in Appendix B1-B3.

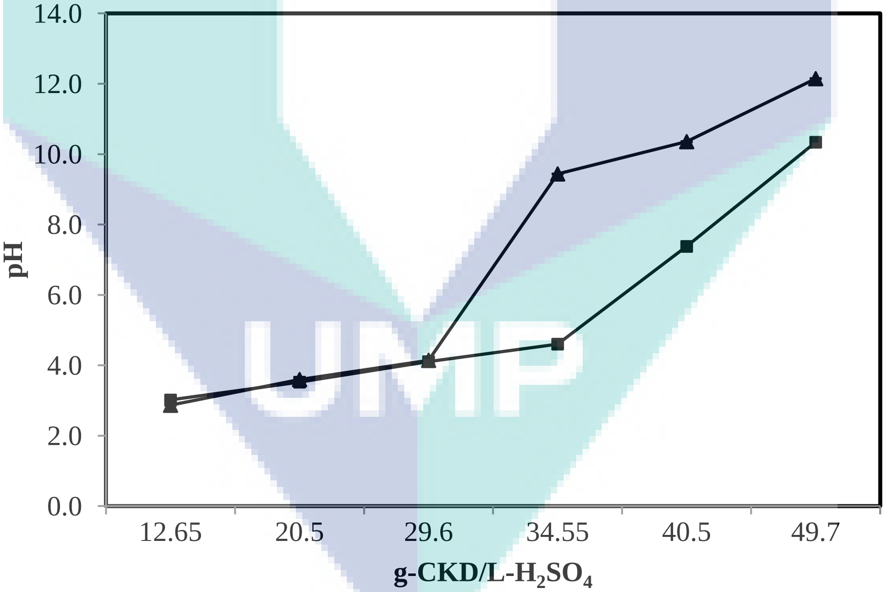

Figure 4.8: Acid neutralization effect with CKD dosage; ■ CKD, ▲ Quick lime (QL)

4.3.3 Effect of CKD on Reactor pH

The CKD proved a gradual effect of substrate pH as shown on Figure 4.9 having moderate effect in terms of neutralisation. Initially, reactor was start up with pH 6.2 but as CKD dosage increased, the pH also started increasing and maintain up to 7.1 till day 23. The decline from day 23 to 32 was observed where pH dropped to 6.21. This might because of fatty acids accumulation in which acids formation mainly have a stay in acidity of substrate and can stop the rapid methane production (Bouallagui et al., 2005). Moreover, it can also be considered that CKD dosage might not enough to overcome the acidity. Furthermore, due to non-recycling of the effluent, there is a possibility of reduction in substrate pH. Because, effluent pH is normally ranging from 7-7.5 and recycling can contribute in maintaining the pH to a great extent.

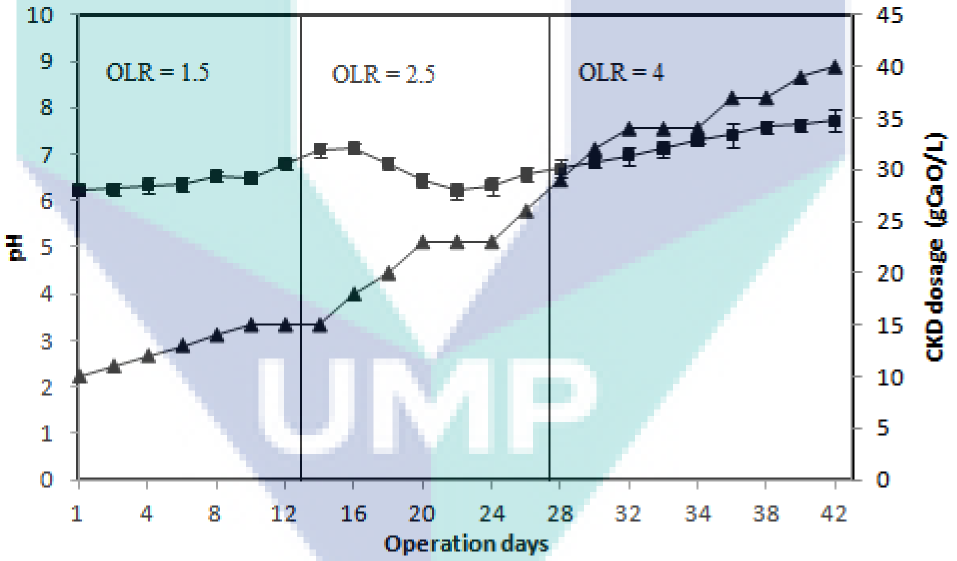

Figure 4.9: The change of pH with CKD dosages during experimental period with OLR (g-COD/L.day), ■ pH, ▲ CKD dose, (values in mean ± SD)

A study by Sandberg and Ahring (1992) proved that fish condensate can be treated well in a UASB reactor from pH 7.3 to 8.2. When the pH was increased slowly to 8.0 or higher then 15–17% drop in COD removal occurred. Acetate was the only carbon source in the condensate that accumulated upon increasing the pH. More than 99% of VFA in process wastewater were degraded up to pH 7.9. It was concluded that gradual pH increment was essential in order to achieve the necessary acclimatization of the granules and to prevent disintegration of the granules and that the pH should not exceed 8.2.

As CKD dosage increased from 23-40 g-CaO/L, the system showed an increasing trend of reactor pH till end of experimental period. This might has several reasons like acidogensis started converting to acetogenesis which contributes to overcome the acidity. Moreover, further CKD dose might mitigate in lowering the acidity along with acetogenesis. Furthermore, the methanogenesis process might be rapid which allow other acids to convert into acetates for methane formation. A correlation (r=0.78) was found thus showing a bit stronger effect of CKD on reactor pH. As pH has a dominating effect on substrate degradation (Liu et al., 2002) so CKD can be a potential neutrilising agent instead of other buffering solutions. Moreover, CKD is a waste material from cement industry and a freely available product which can easily be used as buffering agent in anaerobic treatment of POME. The results for the standard deviation and error in pH values have been shown in Appendix C1.

4.3.4 Effect of CKD on VFA and Alkalinity

The initial VFA was 322 mg-acetic acid/L at OLR 1.5 g-COD/L.day (Figure 4.10) could be attributed to as coming from the effluent after POME treatment. The VFA in the effluent increased with the increase in OLR. The higher VFA accumulation at fix HRT (4 days) is also reflected in the low COD removals. The high levels of VFA in the reactor contributed to the reduced methane level at low pH. The effluent VFA concentration was increased gradually whereby; the alkalinity also increased organic and CKD loadings.

The maximum VFA concentration was 723 mg-acetic acid/L at OLR 4 g-COD/L.day while alkalinity accounted between 1.6-2.4 g-$CaCO_3$/L. The VFA/alkalinity ratio ranged from 0.14-0.35 which shows the suitability of microbial growth. Similar findings were shown by Najafpour et al. (2006) where alkalinity was in the range of 1.57-

3.02 mg/L and maximum VFA/alkalinity ratio was 0.237. The results showed increase in effluent VFA with increase in OLR and CKD dosage.

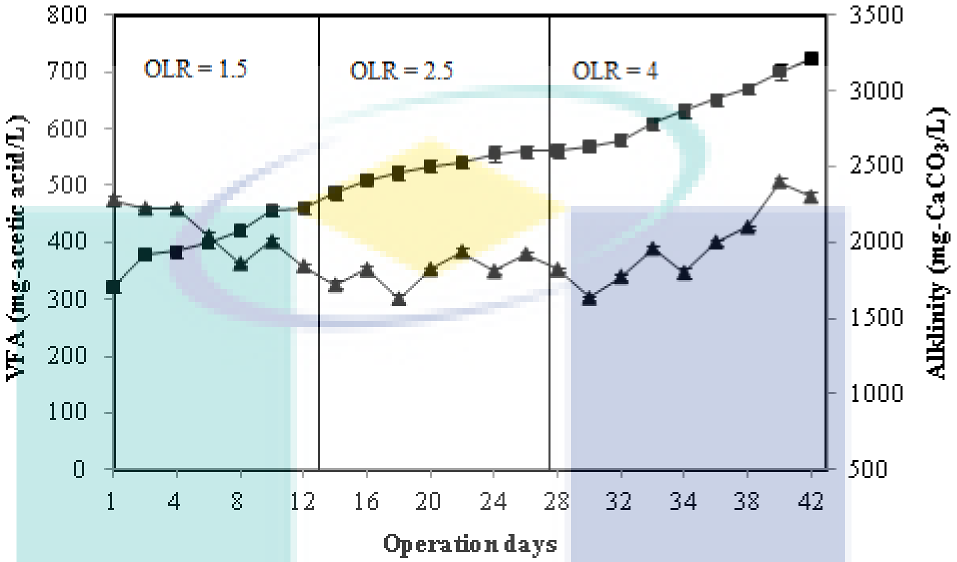

Figure 4.10: VFA and alkalinity during the experimental period with OLR (g-COD/L.day), ■ VFA, ▲ Alkalinity, (values in mean ± SD)

Sulaiman et al. (2009) recorded alkalinity between 2.16-2.79 mg/L throughout the study with VFA/alkalinity ratio ranging from 0.14 to 0.38. Whereas, Zinatizadeh et al. (2006) found effluent VFA between 953 mg-acetic acid/L and concluded that the drop in reactor pH from 7 to 6.6 caused high accumulation of VFA in the reactor. Choorit and Wisarnwan (2007) observed VFA 160 mg-acetic acid/L with relative alkalinity of 3.05 g/L. A negative correlation (r= -0.014) between VFA and alkalinity was found showing that an increase in VFA decreases the alkalinity. Thus system operated at stable conditions at maximum OLR of 4 g-COD/L.day with tolerable VFA/alkalinity ratio and pH values. VFA/alkalinity ratio has been expressed in Appendix C2 along with the standard deviation and error of three replicates for VFA and alkalinity.

4.3.5 Suspended Solids

Figure 4.11 illustrates the suspended solids concentration along four sampling ports with increasing CKD dosage. Initially, CKD and OLR loading were 10 g-CaO/L and 1.5 g-

COD/L.day respectively. The corresponding suspended solids concentration was also not prominent along four sampling ports. Suspended solids concentrations were 7.74 g/L and 0.80 g/L first and fourth sampling port respectively.

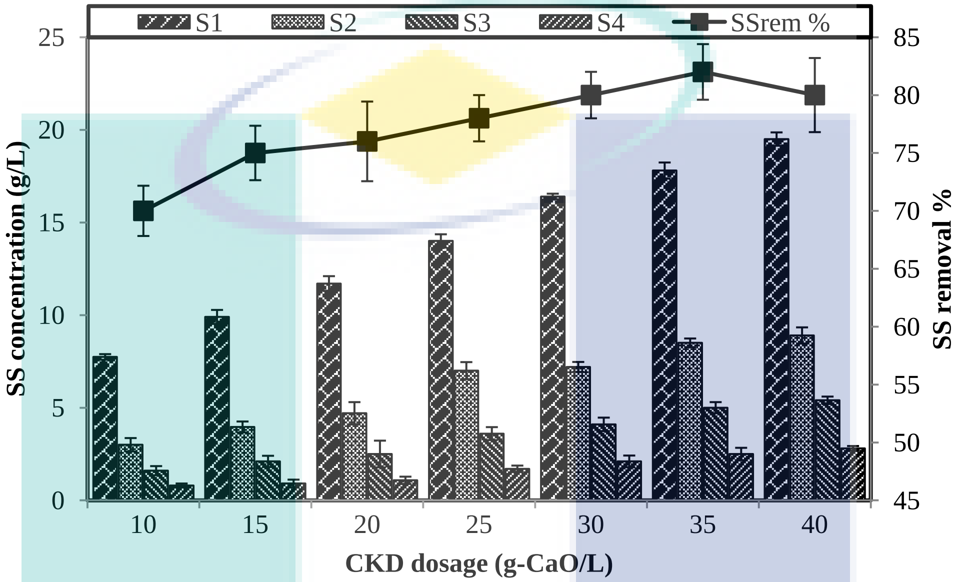

Figure 4.11: Suspended solids concentration along various sampling points with removal percentage by increasing CKD dosage, (values in mean ± SD)

Similarly first and second sampling ports showed higher concentration of suspended solids with increasing CKD dosage. This accumulation of suspended solids might due to heavy particles of CKD which contribute to faster settling of the solids. Moreover, CKD may affect the granule formation and allow excessive solids to settle down at lower flow rates. The influent flow rate was kept constant of 0.675 L/day. The higher feed flow rates contribute in sludge and solids wash out. Similarly high upflow velocities also take part in sludge and solids wash out. More over high gas turbulence may affect in blow the solids in upward direction causing a relatively excessive solids washout from the reactor.

The suspended solids removals are also embedded in Figure 4.8. Increased CKD dosage did not reduce the suspended solids removal. 70% of suspended solids have been observed at 10 g-CaO/L of CKD dose. At CKD 40 g-CaO/L, the suspended solids removal was 80% which indicates the CKD also support in suspended solids removal. This can be proved by Sapari et al. (1996) that influent contained a high proportion of suspended solids which can difficult to hydrolyze within the short HRT. More over these residues resulted in high concentration of loadings do not create any problem in the effluent as they settled easily and can be removed before final effluent discharge. The operation of the UASB reactor with increasing CKD dosage seemed not to cause excessive solids washout. The standard deviation and error of three replicates for solids concentration in the reactor has been shown in Appendix C3.

4.3.6 Flow Rate, HRT and F/M ratio

Flow rate is an important process control parameter which upholds the hydraulic retention time. The HRT was constant so flow rate was also constant during experiment (Table 4.3). The flow rate may vary according to the required volume of the reactor as in case if less volume in the reactor is required then flow rate and HRT both will change accordingly. The F/M ratio was calculated before each OLR and it ranged from 0.29-0.37 which was enough for microbial activity. The change in F/M was because of increase in VSS contents of the reactor. The VSS contents in the reactor were observed ranged from 4-12.6 g/L thus giving a way to keep the maximum OLR of 16 g-COD/L.day. Organic loading rate reveals balanced relationship between F/M ratios. It is a very important influential parameter in the anaerobic digestion. Organic loadings have a strong influence on organic matter degradation ratio during the process of anaerobic digestion. Higher organic loadings are critical factor leading to the acidification in the reactor.

Table 4.3: Process control parameters for reactors stability

COD (g/L)	OLR (g-COD/L)	HRT (days)	Q (L/d)	Vup (m/hr)	F/M (g-COD/ g-VSS.d)	VFA/Alka.	Reactor VSS (g/L)
6	1.5	4	0.675	0.2	0.37	0.25	4.2
10	2.5	4	0.675	0.2	0.27	0.31	8.65
16	4	4	0.675	0.2	0.31	0.29	12.4

Borja et al. (1996) worked on POME at a maximum flow rate of 11.76 L/day and 4.9 L/day at 24 h HRT for a two stage UASB reactor. The difference in flow rate at same HRT is due to change in volume of the reactor from 12 to 5 L. Experiment carried out by Siang (2006) for POME, where low HRT of 12.7 h was worked out at recycling mode UASB and maintained a flow rate of 383.26 L/day with effluent recycling. This low HRT might be due to less height of the reactor (52 cm) and bigger diameter (18.5 cm). Moawad et al. (2009) worked at two HRT of 4-3 h with flow rate of 270-353 L/day and used rectangular UASB rector. Ağdağ and Sponza (2005) studied degradation of municipal wastewater at two stage UASB reactor of same volume (2.5L) and designed flow rate was 2.0 L/day at 1.25 d HRT. The less flow rate accompanied due to reactor configurations (diameter 6 cm and height 100 cm).

4.3.7 Biogas Production

The biogas and methane production as per COD removed is shown in Figure 4.12. The average biogas production has been expressed with respective standard deviation. The addition of CKD caused low biogas production at early stages but it recovers later on as shown in figure. The biogas yield was 0.71 L/g-$COD_{removed}$.day during the first week of experiment which is comparatively low but the same time influent COD concentration was also low (6 g/L). It can be considered that the low biogas production might be because of CKD addition which instantly lowered down the conversion of acetate into biogas. Another reason is that, low influent COD concentrations also results less biogas production.

The biogas production from second week started increasing till end of the experimental period. The biogas yield during last week was 0.90 L/ g-$COD_{removed}$.day. So, it can be concluded that initially the addition of CKD may cause some inhibition in biogas production; it gets stable because of increase in buffer capacity of the substrate. The buffer capacity of substrate encourages the volatile fatty conversion into biogas production. The system worked on stable conditions with no major drawback in acids accumulation.

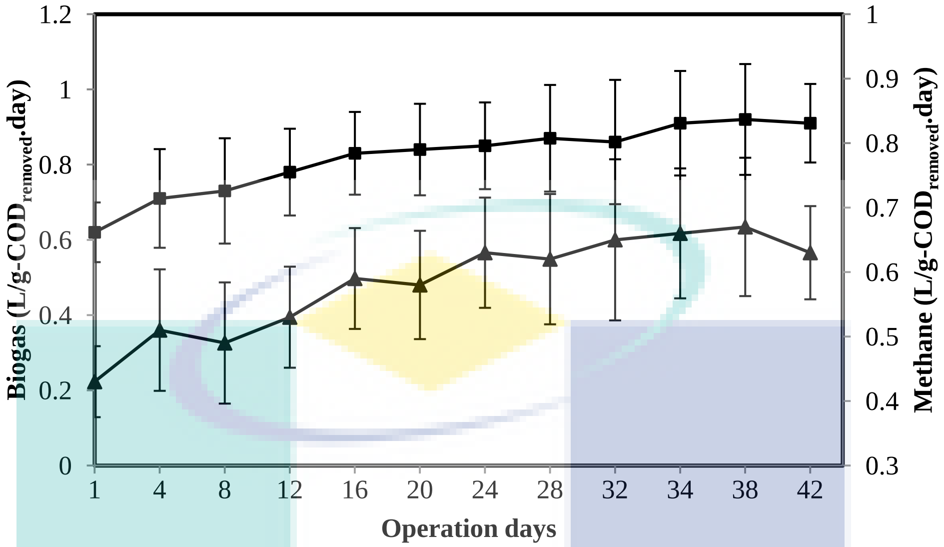

Figure 4.12: Biogas and methane yield along operation days; ■ biogas, ▲ methane, (values in mean ± SD)

In normal anaerobic process the acetate directly converted into methane and carbon dioxide but, by adding CKD which has a major proportion of calcium oxide, first react with acetate to make calcium acetate and then it further breakdown into methane and carbon dioxide by making calcium carbonate as a byproduct. So, the time consumed for the conversion of calcium acetate into methane and carbon dioxide is comparatively higher than that of a normal one. This statement is also similar with Verma (2002) that low biogas yield and organic matter removal because of the accumulation of inhibiting substances, such as volatile fatty acids. Chaisri et al. (2007) proposed similar findings that VFA decrease the biogas production initially and they found biogas production of 0.12 L/day. However, at the end of this period the maximum biogas production of 5.98 L/day was found. In another study, Zinatizadeh et al. (2007) recorded methane production rate ranging from 9.7-25.5 L/day (COD$_{in}$ 43 g/L) by using upflow anaerobic sludge blanket fix film reactor for POME treatment. Thus it can be concluded that biogas production at early stage of experiment is low and it increased by increasing loading rates, conversion of volatile fatty acids into methane and the buffering capacity of the substrate. Dague and Pidaparti (1992) concluded that operation of reactor with hydraulic retention time of 8.8 days and organic loading rate of 0.33 kg-BOD_5/m^3.day, yielded a BOD_5 removal efficiency of 85-90% and biogas production of 0.51 m^3-CH_4/ kg-$COD_{removed}$.

The trend of biogas production by using CKD as neutralizing agent was initially low. The low biogas production due to some co-substrate of buffering solution also supports the finding of Sulaiman et al. (2009) where they used refined glycerin wash water for COD removal and biogas production from POME. They found that high concentration of refined glycerin water has biogas production while low concentrations revealed comparatively higher biogas production. Appendix C4 shows the standard deviation and error of three replicate results for biogas and methane production.

4.3.8 COD Removal Efficiency

The COD removal efficiency was initially low but starting from almost 60% (Figure 4.13). This slightly high COD removal at early stages was because of the use of seed sludge where POME sample was activated and new COD feed supported for high removals. The COD removal efficiency was increased till day 13 and slight decrease was observed when new loading with higher COD concentration was done. After that it started improving and reached to a maximum of 95% and it was fluctuated within 3% range till the end of experiment.

The higher COD removal efficiency due to low OLR because instability had been seen by Borja and Banks (1994) where they applied 10.63 kg-COD/m^3.day OLR at 4 days HRT. In this study, COD removal efficiency up to 95% was achieved at HRT of 4 days with 4 g-COD/L.day of OLR.

A strong relationship (r=0.95) was found between COD removal efficiency and influent COD concentration. For this study, CKD used as stabilizing and neutrilising agent for POME treatment which contributed in high COD removal efficiencies even at 16 g/L influent COD concentrations. The system seemed to be stable and further increase of influent COD could be accomplished at same HRT of 4 days.

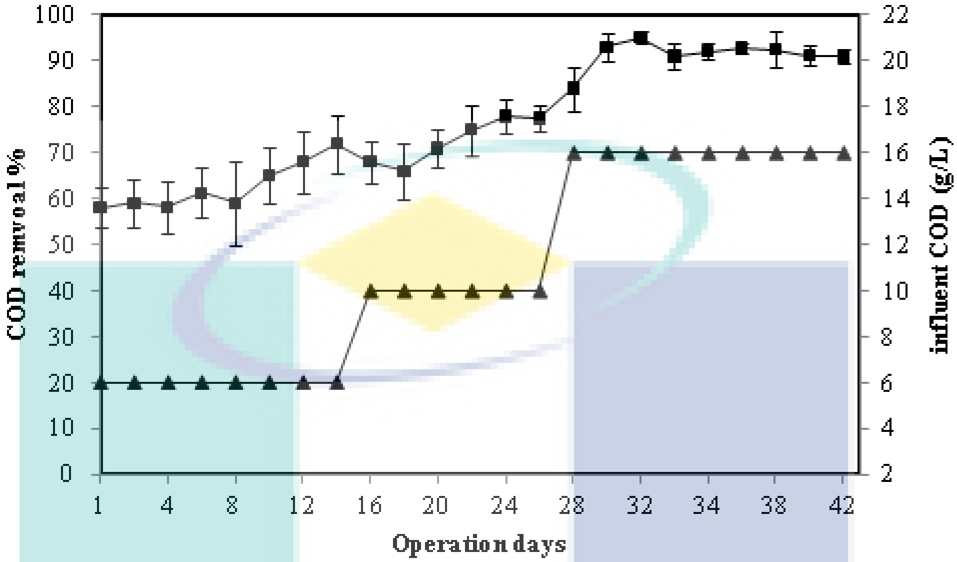

Figure 4.13: COD concentration and removal along time, ■ COD removal, ▲ Influent COD, (values in mean ± SD)

POME has been successful with UASB reactors, achieving COD removal efficiency up to 98.4% with the highest operating OLR of 10.63 kg-COD/m^3.day (Borja and Banks, 1994). However, reactor operated under overload conditions with high volatile fatty acid content became unstable after 15 days. Due to high amount of POME discharge daily from milling process, it is necessary to operate treatment system at higher OLR.

Borja et al. (1996) implemented a two-stage UASB system for POME treatment with the objective of preventing inhibition of granule formation at higher OLRs without having to remove solids from POME prior to treatment. This method is desirable since suspended solids in POME have high potential for gas production while extra costs from sludge disposal can be avoided. Results from this study showed the feasibility of separating anaerobic digestion into two-stages (acidogensis and methanogenesis) using a pair of UASB reactors. The methanogenic reactor was found to adapt quickly with the feed from the acidogenic reactor and also tolerate higher OLRs. It was suggested that OLR of 30 kg-COD/m^3.day could ensure an overall of 90% COD reduction and efficient methane conversion.

The treatment of POME with CKD may cause accumulation in shape of suspended solids during UASB treatment but it can be advantageous for its ability to treat wastewater with high suspended solid contents (Fang and Chui, 1994; Kalyuzhnyi et al., 1998) that may clog the packing materials in other reactors (Stronach et al., 1987; Kalyuzhnyi et al., 1996). Moreover, suspended and colloidal components of POME in the form of fat, protein, and cellulose have an adverse impact on UASB reactor performance and can cause deterioration of microbial activities and wash out of the active biomass (Borja and Banks, 1994; Torkian et al., 2003). However, this reactor might face long start-up periods if seeded sludge is not granulated. A study by Goodwin et al. (1992) has proved that reactors seeded with granulated sludge achieved high performance levels within a shorter start-up period. It could also acclimatize quickly to gradual increase of OLR (Kalyuzhnyi et al., 1996).

The use of two identical UASB reactors by Goodwin and Stuart (1994) operated in parallel for 327 days for the treatment of malt whisky pot ale, achieved COD reductions of up to 90% for influent concentrations of 3.5–5.2 g/L. When the OLRs of 15 kg-COD/m^3.day and above were used, the COD removal efficiency dropped to less than 20% in one of the reactor. A mesophilic two-stage system consisting of an anaerobic filter (AF) and an UASB reactor was found suitable for anaerobic digestion of distillery waste, enabling better conditions for the methanogenic phase (Blonskaja et al., 2003). A three-phased UASB reactor used by Caixeta et al. (2002) for slaughterhouse waste water treatment at an organic loading rate of 2.7-10.8 kg-COD/m^3.day and average COD removal efficiencies of 85, 84 and 80% at three different HRT of 22, 18 and 14 h, respectively. Syutsubo et al. (1997) reported a COD loading of 30 kg-COD/m^3.day with a COD removal efficiency of 85% at sludge loading rates (SLRs) of up to 3.7 g-COD/g-VSS.day for thermophilic reactors (Syutsubo et al., 1998). Organic loading rates (OLR) of up to 104 kg-COD/m^3.day have been reported for anaerobic digestion of sugar substrate under thermophilic conditions (Wiegant and Lettinga, 1985).

Torkian et al. (2003) concluded results under steady state condition where OLRs was between 13 and 39 kg-SCOD/m^3.day and HRT of 2-7 h. Removal efficiencies in the range of 75-90% were achieved at feed SCOD concentrations of 3-4.5 g/L. According to

Soto et al. (1997), excellent stability and high treatment efficiency can be achieved with hydraulic residence times as low as 2 h at an OLR of 6 kg-COD/m^3.day with the percent COD removals being 92-95%. Sayed et al. 1987 treated effluent from meat processing plant in a granular UASB reactor and achieved COD removal efficiency of 55-85% with HRT of 0.5-0.6 days at volumetric loading rate of 11 kg-COD/m^3.d. Two-phased UASB-septic tank used by Luostarinen and Rintala (2005) with high removal of organic matter for onsite treatment of synthetic black water (OLR 0.301 kg-COD/m^3.d) and dairy parlor wastewater (OLR 0.191 kg-COD/m^3.day) at low temperatures (10–20 ^{o}C). Moreover, COD$_{dissolve}$ removal was around 70% at 15 ^{o}C and 10 ^{o}C indicating good biological activity of the reactor sludge. Gavala et al. (1999) concluded that at an OLR of 6.2 g-COD/L.day may be safely used for treating dairy wastewater and could be increased up to 7.5 g-COD/L.day. Above that OLR, reduced performance is observed; while for non-diluted dairy wastewater, at HRT in excess of 30 days is required.

According to Palenzuela et al. (2002) the application of UASB system is a promising treatment option for fish processing wastewater. They determined the performance of USAB reactor for the treatment of mixed sardine and tuna canning effluent at varying lipid levels. They stated that at low lipid level (203-261 mg/L, 9% of total COD) approximately 78% COD removal and 61% COD conversion to methane can be achieved with an OLR of 2.3 g-COD/L.day and an HRT of 7.2 ± 2.8 h. In the case of high-lipid wastewater a two step UASB was recommended where the total COD removal and conversion to methane were 92% and 47%, respectively. Strydom et al. (1995) concluded that COD removal rate could reach a maximum and then start to decrease which is a sign of insufficient microbial biomass accumulation in the reactor to carry the additional organic load. In this study, the microbial biomass was sufficient to control further organic loads in terms of high COD concentrations at same HRT (4 days). The novelty of this study lies on the fact that despite the unavailability of other buffering agents, CKD can take place of these reagents without inhibiting the loading rates, removal efficiencies and biogas production. Appendix C5 shows the standard deviation and error of three replicates of COD removal percentage values.

CHAPTER 5

CONCLUSIONS AND RECOMMENDATIONS

5.1 CONCLUSIONS

A UASB reactor is efficient for COD removal and high methane production. A lab scale reactor was constructed to study the mesophilic to thermophilic anaerobic treatment of food waste. VFA accumulation was low and methane production was comparatively high at controlled temperature and pH. But a sudden change in temperature had adverse effect on biogas production and system stability. The high COD concentrations in the early experimental period at 55 ^{o}C provide an adequate substrate supply for the subsequent acidogenesis and methanogenesis steps. Thus, at 55^{o}C temperature and OLR of 12.5 g-COD/L.day with 4 d HRT support a maximum biogas production of 1.37 L/g-COD$_{removed}$.day.

Cement kiln dust can now proposed to be used in this study as an alkaline source to neutralize the POME and a degrade organic matter with aim to reduce chemical cost for pH adjustment and treatment. The reactor pH was successfully attained above 7.5 without any serious problem. The VFA/alkalinity ratio was remained below 0.40 which supported to bacterial growth. COD removal efficiency achieved was maximum 95% at an OLR of 4.5 g-COD/L.d. Further research can be done on granules morphology and microbial activity in biomass. Moreover, biogas production potential can be studied with CKD. Furthermore, effect of temperature on POME treatment with CKD can also be furnished in future studies.

5.2 RECOMMENDATIONS

For the treatment of liquidized food waste, the start up of UASB reactor was monitored with the effect of temperature and organic loading rate for biogas production. The successful experimental results can lead the food waste treatment for the following further research

- The liquidized food waste treatment should be carried out for steady state conditions in order to conclude the full experimental results based on reactor stability on organic shock load. (OLR higher than 12.5 g-COD/L.day)
- The microbial activity in terms of biogas should be monitored for maximizing biogas production.
- The granule formation during liquidized food waste with different temperature and loading rates can also be a part of future research.
- Feasibility of the energy recovery from pilot scale UASB reactor for liquidized food waste should be carried out.
- The significant kinetic of liquidized food waste treatment in UASB reactor should be investigated.

The feasibility of cement kiln dust as neutralizing source during palm oil mill treatment using UASB reactor has been proved successfully. The further research can be performed in terms of;

- The use of cement kiln dust as a potential neutralizing source for palm oil mill effluent should be continued for steady state conditions.
- The effect of cement kiln dust on palm oil mill effluent granule formation and structure should be carried out.
- The effect of cement kiln dust on microbiology of palm oil mill effluent should be investigated.
- The kinetics study of palm oil mill effluent with cement kiln dust should be performed for higher biogas production.

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APPENDIX A1

INFLUENT TOC AND ITS REMOVAL PERCENTAGE FROM LFW

Days	n	Influent TOC (g/L)			TOC removal (%)		
		Mean	±SD	±SE	Mean	±SD	±SE
1	3	21.13	0.61	0.35	33	2.08	1.20
3	3	18.30	0.36	0.21	39	2.05	1.18
5	3	13.47	0.21	0.12	43	1.00	0.58
7	3	9.33	0.12	0.07	53	2.31	1.33
9	3	6.30	0.36	0.21	59	1.15	0.67
11	3	3.60	0.08	0.05	72	2.00	1.15
13	3	4.04	0.11	0.06	76	3.21	1.86
15	3	4.23	0.11	0.06	77	1.53	0.88
17	3	3.99	0.07	0.04	74	2.08	1.20
19	3	3.39	0.08	0.04	79	1.73	1.00
21	3	3.69	0.07	0.04	75	2.00	1.15
23	3	3.59	0.12	0.07	79	2.00	1.15
25	3	3.71	0.14	0.08	79	2.31	1.33
27	3	4.08	0.08	0.05	77	3.06	1.76
29	3	4.48	0.15	0.09	79	1.53	0.88
31	3	4.13	0.08	0.05	79	2.00	1.15
33	3	4.44	0.10	0.06	79	1.53	0.88
35	3	4.65	0.11	0.06	77	2.52	1.45
37	3	3.26	0.12	0.07	76	1.53	0.88
39	3	4.70	0.08	0.04	78	1.53	0.88
41	3	4.65	0.08	0.04	66	2.08	1.20
43	3	3.88	0.18	0.10	68	3.06	1.76
45	3	3.44	0.15	0.09	66	3.06	1.76
47	3	5.37	0.17	0.10	64	1.00	0.58
49	3	5.71	0.23	0.13	66	2.08	1.20
51	3	5.21	0.16	0.09	68	2.52	1.45
53	3	7.21	0.16	0.09	56	3.06	1.76
55	3	8.98	0.18	0.10	53	1.73	1.00
57	3	13.78	0.33	0.19	48	2.08	1.20
59	3	17.75	0.32	0.18	44	2.00	1.15
61	3	18.11	0.34	0.19	42	2.08	1.20
63	3	17.32	0.24	0.14	42	1.53	0.88
65	3	19.43	0.25	0.15	41	2.52	1.45
67	3	19.28	0.38	0.22	42	1.53	0.88
69	3	19.60	0.47	0.27	43	2.00	1.15
71	3	18.21	0.44	0.25	44	2.00	1.15

APPENDIX A2

EFFECT OF OLR ON COD REMOVAL FROM LFW

Days	OLR (g-COD/L.day)	n	COD removal (%)		
			Mean	±SD	±SE
1	2	3	40	2.00	1.15
3	2	3	45	3.06	1.76
5	2	3	52	1.50	0.87
7	3	3	55	2.08	1.20
9	3	3	59	1.15	0.67
11	3	3	61	3.21	1.86
13	3	3	61	3.22	1.86
15	3	3	62	1.72	0.99
17	4	3	64	1.37	0.79
19	4	3	68	2.11	1.22
21	4	3	75	1.27	0.73
23	4	3	87	3.11	1.79
25	5	3	78	3.79	2.19
27	5	3	69	5.00	2.89
29	5	3	66	4.36	2.52
31	6.25	3	63	2.52	1.45
33	6.25	3	69	4.58	2.65
35	6.25	3	75	5.51	3.18
37	6.25	3	79	1.15	0.67
39	6.25	3	81	5.03	2.91
41	6.25	3	78	4.51	2.60
43	6.25	3	80	4.73	2.73
45	8.33	3	83	2.89	1.67
47	8.33	3	87	3.51	2.03
49	8.33	3	82	2.89	1.67
51	8.33	3	80	4.04	2.33
53	8.33	3	87	3.51	2.03
55	8.33	3	85	2.81	1.62
57	10.00	3	90	2.47	1.43
59	10.00	3	89	1.18	0.68
61	10.00	3	90	1.01	0.58
63	10.00	3	92	1.96	1.13
65	10.00	3	90	3.41	1.97
67	12.50	3	93	3.01	1.74
69	12.50	3	94	3.82	2.20
71	12.50	3	93	4.36	2.52

APPENDIX A3

BIOGAS AND METHANE PRODUCTION FROM LFW

Days	OLR (g-COD/L.day)	n	Biogas (L/g-COD$_{removed}$.day)			Mehtane (L/g-COD$_{removed}$.day)		
			Mean	±SD	±SE	Mean	±SD	±SE
1	2	3	0.43	0.06	0.04	0.29	0.04	0.02
3	2	3	0.62	0.05	0.03	0.39	0.05	0.03
5	2	3	0.72	0.04	0.02	0.47	0.05	0.03
7	3	3	0.78	0.04	0.02	0.48	0.03	0.02
9	3	3	0.83	0.04	0.02	0.50	0.04	0.02
11	3	3	0.84	0.03	0.02	0.54	0.04	0.02
13	3	3	0.85	0.04	0.02	0.52	0.03	0.02
15	3	3	0.87	0.06	0.04	0.54	0.04	0.02
17	4	3	0.85	0.06	0.03	0.53	0.04	0.02
19	4	3	0.91	0.03	0.02	0.57	0.06	0.03
21	4	3	0.92	0.03	0.02	0.58	0.05	0.03
23	4	3	0.94	0.04	0.02	0.58	0.06	0.03
25	5	3	0.87	0.05	0.03	0.50	0.05	0.03
27	5	3	0.64	0.04	0.02	0.36	0.05	0.03
29	5	3	0.54	0.05	0.03	0.32	0.06	0.03
31	6.25	3	0.63	0.04	0.02	0.39	0.05	0.03
33	6.25	3	0.99	0.03	0.02	0.61	0.05	0.03
35	6.25	3	1.28	0.04	0.02	0.80	0.07	0.04
37	6.25	3	1.21	0.04	0.02	0.76	0.09	0.05
39	6.25	3	1.22	0.03	0.02	0.77	0.07	0.04
41	6.25	3	1.25	0.05	0.03	0.80	0.05	0.03
43	6.25	3	1.23	0.03	0.02	0.74	0.05	0.03
45	8.33	3	1.32	0.04	0.02	0.82	0.06	0.03
47	8.33	3	1.29	0.05	0.03	0.80	0.05	0.03
49	8.33	3	1.28	0.07	0.04	0.81	0.04	0.02
51	8.33	3	1.27	0.04	0.02	0.82	0.05	0.03
53	8.33	3	1.26	0.06	0.03	0.83	0.05	0.03
55	8.33	3	1.27	0.04	0.02	0.82	0.03	0.02
57	10.00	3	1.28	0.04	0.02	0.84	0.03	0.01
59	10.00	3	1.26	0.06	0.04	0.83	0.07	0.04
61	10.00	3	1.19	0.02	0.01	0.78	0.04	0.02
63	10.00	3	1.22	0.07	0.04	0.79	0.04	0.02
65	10.00	3	1.29	0.05	0.03	0.83	0.06	0.04
67	12.50	3	1.34	0.06	0.04	0.88	0.05	0.03
69	12.50	3	1.36	0.05	0.03	0.91	0.04	0.02
71	12.50	3	1.36	0.06	0.03	0.91	0.05	0.03

APPENDIX A4

EFFECT TEMPERATURE ON REACTOR VFA FOR LFW

Days	Temperature (°C)	n	VFA (g-acetic acid/L)		
			Mean	±SD	±SE
1	30	3	3.60	0.12	0.07
3	30	3	3.80	0.17	0.10
5	30	3	4.60	0.13	0.08
7	33	3	3.60	0.16	0.09
9	33	3	4.05	0.20	0.11
11	33	3	4.13	0.12	0.07
13	33	3	4.29	0.18	0.10
15	33	3	4.93	0.15	0.09
17	37	3	5.94	0.21	0.12
19	37	3	5.60	0.08	0.05
21	37	3	4.40	0.04	0.02
23	37	3	4.00	0.22	0.12
25	50	3	7.00	0.15	0.09
27	50	3	7.50	0.09	0.05
29	50	3	8.00	0.11	0.06
31	50	3	8.50	0.17	0.10
33	50	3	9.00	0.19	0.11
35	50	3	8.25	0.15	0.09
37	50	3	7.78	0.16	0.09
39	50	3	7.50	0.16	0.09
41	50	3	7.25	0.16	0.09
43	50	3	7.09	0.21	0.12
45	55	3	7.50	0.07	0.04
47	55	3	7.75	0.10	0.06
49	55	3	7.99	0.12	0.07
51	55	3	8.11	0.19	0.11
53	55	3	8.43	0.35	0.20
55	55	3	8.28	0.16	0.09
57	55	3	8.36	0.12	0.07
59	55	3	8.60	0.09	0.05
61	55	3	8.57	0.10	0.06
63	55	3	8.42	0.07	0.04
65	55	3	8.20	0.11	0.06
67	55	3	8.11	0.12	0.07
69	55	3	8.01	0.20	0.11
71	55	3	7.95	0.17	0.10

APPENDIX A5

EFFECT OF TEMPERATURE ON REACTOR PH FOR LFW

Days	Temperature (^oC)	n	Reactor pH		
			Mean	±SD	±SE
1	30	3	7.2	0.10	0.06
3	30	3	6.8	0.17	0.10
5	30	3	6.3	0.13	0.08
7	33	3	6.7	0.12	0.07
9	33	3	6.8	0.26	0.15
11	33	3	6.9	0.09	0.05
13	33	3	7.4	0.13	0.08
15	33	3	7.5	0.09	0.05
17	37	3	7.3	0.12	0.07
19	37	3	7.0	0.16	0.09
21	37	3	7.1	0.13	0.07
23	37	3	7.2	0.24	0.14
25	50	3	6.9	0.23	0.13
27	50	3	6.3	0.19	0.11
29	50	3	6.0	0.19	0.11
31	50	3	6.2	0.12	0.07
33	50	3	7.8	0.20	0.12
35	50	3	7.9	0.13	0.08
37	50	3	8.2	0.23	0.13
39	50	3	8.3	0.19	0.11
41	50	3	8.7	0.14	0.08
43	50	3	8.2	0.09	0.05
45	55	3	8.7	0.16	0.09
47	55	3	8.8	0.10	0.06
49	55	3	8.8	0.07	0.04
51	55	3	8.8	0.24	0.14
53	55	3	8.9	0.04	0.02
55	55	3	8.3	0.08	0.05
57	55	3	8.6	0.07	0.04
59	55	3	8.7	0.22	0.13
61	55	3	8.4	0.15	0.09
63	55	3	8.2	0.12	0.07
65	55	3	8.7	0.09	0.05
67	55	3	8.7	0.12	0.07
69	55	3	8.8	0.11	0.07
71	55	3	8.9	0.15	0.09

APPENDIX B1

pH RESULTS OF FILTERED CKD SOLUTION

Dose (g-CaO)/L)	**n**	**pH with CKD**			**pH with Quick lime**		
		Mean	**±SD**	**±SE**	**Mean**	**±SD**	**±SE**
8.06	3	11.8	0.29	0.17	11.95	0.13	0.08
13.14	3	11.9	0.14	0.08	12.04	0.10	0.06
18.76	3	12.2	0.13	0.08	12.31	0.03	0.02
22.46	3	12.3	0.21	0.12	12.41	0.03	0.01
26.7	3	12.3	0.23	0.13	12.50	0.02	0.01
32.96	3	12.4	0.15	0.09	12.59	0.02	0.01

APPENDIX B2

CONDUCTIVITY RESULTS OF FILTERED CKD SOLUTION

Dose (g-CaO)/L)	**n**	**Conductivity (mS/cm) with CKD**			**Conductivity (mS/cm) with Quick lime**		
		Mean	**±SD**	**±SE**	**Mean**	**±SD**	**±SE**
8.06	3	1.87	0.15	0.09	2.27	0.02	0.01
13.14	3	2.70	0.30	0.17	3.20	0.01	0.01
18.76	3	3.27	0.21	0.12	3.87	0.02	0.01
22.46	3	4.07	0.25	0.15	4.77	0.04	0.03
26.7	3	4.47	0.25	0.15	5.29	0.03	0.02
32.96	3	5.03	0.15	0.09	6.13	0.02	0.01

APPENDIX B3

ACID NEUTRALIZING EFFECT OF CKD DOSAGE

g-CKD/L-H_2SO_4	n	pH with CKD			pH with Quick lime		
		Mean	±SD	±SE	Mean	±SD	±SE
12.65	3	3.02	0.10	0.06	2.87	0.03	0.01
20.5	3	3.52	0.06	0.04	3.60	0.04	0.02
29.6	3	4.11	0.04	0.02	4.14	0.02	0.01
34.55	3	4.60	0.02	0.01	9.44	0.02	0.01
40.5	3	7.38	0.03	0.02	10.36	0.03	0.01
49.7	3	10.34	0.05	0.03	12.15	0.03	0.01

APPENDIX C1

EFFECT OF CKD DOSE ON REACTOR pH

Days	CKD Dosage (g-CaO/L)	n	Reactor pH		
			Mean	±SD	±SE
1	10	3	6.23	0.11	0.06
2	11	3	6.26	0.12	0.07
4	12	3	6.32	0.18	0.10
6	13	3	6.36	0.16	0.09
8	14	3	6.52	0.11	0.06
10	15	3	6.49	0.10	0.06
12	15	3	6.8	0.10	0.06
14	15	3	7.1	0.15	0.09
16	18	3	7.13	0.12	0.07
18	20	3	6.8	0.13	0.08
20	23	3	6.44	0.15	0.08
22	23	3	6.21	0.17	0.10
24	23	3	6.32	0.20	0.12
26	26	3	6.57	0.14	0.08
28	29	3	6.69	0.20	0.12
30	32	3	6.82	0.10	0.06
32	34	3	6.97	0.19	0.11
33	34	3	7.12	0.18	0.10
34	34	3	7.29	0.07	0.04
36	37	3	7.41	0.26	0.15
38	37	3	7.59	0.12	0.07
40	39	3	7.63	0.12	0.07
41	40	3	7.72	0.24	0.14

APPENDIX C2

EFFECT OF CKD DOSE ON REACTOR VFA AND ALKALINITY FOR POME

Days	**CKD Dosage (g-CaO/L)**	**n**	**VFA (mg-acetic acid/L)**			**Alklinity (mg-$CaCO_3$/L)**			**VFA/Alk.**
			Mean	**±SD**	**±SE**	**Mean**	**±SD**	**±SE**	
1	10	3	322	10.54	6.08	2278	24.56	14.18	0.14
2	11	3	378	8.54	4.93	2224	13.53	7.81	0.17
4	12	3	384	10.58	6.11	2220	15.87	9.17	0.17
6	13	3	398	7.55	4.36	2040	25.06	14.47	0.20
8	14	3	421	9.85	5.69	1858	17.06	9.85	0.23
10	15	3	456	8.89	5.13	2006	18.36	10.60	0.23
12	15	3	460	10.54	6.08	1843	19.16	11.06	0.25
14	15	3	488	12.00	6.93	1722	27.87	16.09	0.28
16	18	3	510	9.54	5.51	1821	25.63	14.80	0.28
18	20	3	523	12.00	6.93	1630	26.21	15.13	0.32
20	23	3	534	11.14	6.43	1822	15.52	8.96	0.29
22	23	3	540	7.00	4.04	1938	25.87	14.93	0.28
24	23	3	555	13.86	8.00	1810	21.00	12.12	0.31
26	26	3	560	11.27	6.51	1921	23.52	13.58	0.29
28	29	3	562	12.12	7.00	1820	16.46	9.50	0.31
30	32	3	568	8.72	5.03	1635	17.06	9.85	0.35
32	34	3	580	10.15	5.86	1775	17.35	10.02	0.33
33	34	3	610	9.85	5.69	1957	19.08	11.02	0.31
34	34	3	632	12.12	7.00	1800	28.93	16.70	0.35
36	37	3	652	9.54	5.51	2000.00	19.97	11.53	0.33
38	37	3	670	8.00	4.62	2100.00	10.54	6.08	0.32
40	39	3	700	14.00	8.08	2400.00	30.05	17.35	0.29
41	40	3	723	9.00	5.20	2300.00	29.51	17.04	0.31

APPENDIX C3

SUSPENDED SOLIDS CONCENTRATION ALONG FOUR SAMPLING PORTS FOR POME

CKD Dosage (g-CaO/L)	n	Sampling port 1			Sampling port 2			Sampling port 3			Sampling port 4		
		Mean	±SD	±SE	Mean	±SD	±SE	Mean	±SD	±SE	Mean	±SD	±SE
10	3	7.74	0.15	0.09	3.00	0.36	0.21	1.60	0.25	0.14	0.80	0.11	0.06
15	3	9.90	0.38	0.22	3.96	0.30	0.17	2.10	0.30	0.18	0.91	0.21	0.12
20	3	11.70	0.40	0.23	4.70	0.60	0.35	2.50	0.72	0.42	1.08	0.20	0.11
25	3	14.00	0.36	0.21	7.00	0.46	0.27	3.60	0.35	0.20	1.70	0.18	0.10
30	3	16.40	0.15	0.09	7.20	0.27	0.16	4.10	0.36	0.21	2.10	0.32	0.18
35	3	17.80	0.44	0.25	8.50	0.24	0.14	5.00	0.30	0.17	2.50	0.33	0.19
40	3	19.50	0.36	0.21	8.90	0.44	0.25	5.40	0.20	0.12	2.80	0.13	0.08

APPENDIX C4

BIOGAS AND METHANE PRODUCTION FROM POME

Days	n	Biogas (L/g-COD$_{removed}$.day)			Methane (L/g-COD$_{removed}$.day)			Biogas production (L/day)	Methane production (L/day)
		Mean	±SD	±SE	Mean	±SD	±SE		
1	3	0.62	0.08	0.05	0.43	0.06	0.03	2.16	1.50
4	3	0.71	0.13	0.08	0.51	0.09	0.05	2.48	1.78
8	3	0.73	0.14	0.08	0.49	0.09	0.05	2.58	1.73
12	3	0.78	0.12	0.07	0.53	0.08	0.05	3.18	2.16
16	3	0.83	0.11	0.06	0.59	0.08	0.05	5.64	4.01
20	3	0.84	0.12	0.07	0.58	0.08	0.05	5.96	4.12
24	3	0.85	0.12	0.07	0.63	0.09	0.05	6.63	4.91
28	3	0.87	0.14	0.08	0.62	0.10	0.06	11.69	8.33
32	3	0.86	0.17	0.10	0.65	0.12	0.07	13.07	9.88
34	3	0.91	0.14	0.08	0.66	0.10	0.06	13.40	9.72
38	3	0.92	0.15	0.09	0.67	0.11	0.06	13.60	9.91
42	3	0.91	0.10	0.06	0.63	0.07	0.04	13.25	9.17

APPENDIX C5

INFLUENT COD AND ITS REMOVAL PERCENTAGE FROM POME

Days	Influent COD (g/L)	n	COD removal (%)		
			Mean	±SD	±SE
1	6	3	58.00	4.36	2.52
2	6	3	59.00	5.29	3.06
4	6	3	58.20	5.52	3.19
6	6	3	61.30	5.56	3.21
8	6	3	59.00	9.17	5.29
10	6	3	65.00	6.00	3.46
12	6	3	68.00	6.76	3.91
14	6	3	72.00	6.35	3.67
16	10	3	68.00	4.58	2.65
18	10	3	66.00	6.24	3.61
20	10	3	71.00	4.00	2.31
22	10	3	75.00	5.44	3.14
24	10	3	78.00	3.78	2.18
26	10	3	77.50	2.67	1.54
28	16	3	84.00	4.79	2.76
30	16	3	93.00	2.88	1.67
32	16	3	95.00	1.35	0.78
33	16	3	91.00	2.89	1.67
34	16	3	92.00	1.73	1.00
36	16	3	92.70	1.01	0.59
38	16	3	92.40	3.82	2.20
40	16	3	91.20	2.02	1.17
41	16	3	91.00	1.61	0.93

APPENDIX D1

LIST OF PUBLICATIONS

Published Papers:

Muhammad Asif Latif, Anwar Ahmad, Rumana Ghufran, Zularisam Abdul Wahid, (2011). Integrated Application of Upflow Anaerobic Sludge Blanket Reactor for the Treatment of Wastewaters; A Review, "*Water Research*". DOI:10.1016/j.watres.2011.05.049

Muhammad Asif Latif, Anwar Ahmad, Rumana Ghufran, Zularisam Abdul Wahid, (2011). Effect of Temperature and Organic Loading Rate on Upflow Anaerobic Sludge Blanket Reactor (UASBR) and CH_4 production by Treating Liquidized Food Waste, "*Environmental Progress and Sustainable Energy*", Wiley. DOI 10.1002/ep.10540.

Muhammad Asif Latif, Anwar Ahmad, Rumana Ghufran, Zularisam Abdul Wahid, (2010). Thermophilic Anaerobic UASB Reactor to Remove COD and Waste Reduction from Food Waste. Proceedings of *Malaysian Technical Universities Conference on Engineering and Technology (MUCET)*. June 28-29, Melaka, Malaysia.