Anaerobic treatment of nightsoil and toilet sludge from on-site sanitation systems in Ghana

By

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The candidate confirms that the work submitted is his own and that appropriate credit has been given where reference has been made to the work of others.

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Abstract

The feasibility of faecal sludge treatment by an upflow anaerobic sludge blanket (UASB) reactor was studied using first, untreated primary sludge from a sewage treatment works treating only domestic sewage, and then actual faecal sludge. The primary sludge was diluted in the ratio 1:20 – 1:10 while the faecal sludge was diluted in to the ratio 1:6. The UASB reactor treating the primary sludge had a volume of 15 litres and was operated at a mean hydraulic retention time (HRT) of 9.8 h, at a temperature of 37 °C, and at an organic loading rate (OLR) in the range of 5.6 – 15.0 kg COD/m^3.d. The UASB reactor treating the faecal sludges had a volume of 50 litres and was operated at a mean HRT of 12.1 h, at ambient temperatures in the range of 23.0 – 31.2 °C, and at OLR in the range of 12.5 – 21.5 kg COD/m^3.d. The first experiment involving the untreated primary sludge was run for 114 days while the second was run for 119 days. The results from both experiments indicate that it is feasible to treat faecal sludges using the UASB reactor. The average removal efficiencies obtained for the first experiment were: 78% for COD, 62% for total solids (TS), 75% for total volatile solids (TVS) and 91% for total suspended solids (TSS). The pH was in the range of 6.9 – 7.4. With regards to faecal sludges, the average removal efficiencies were: 71% for COD, 61% for TS, 74% for TVS and 73% for TSS. The removal efficiencies are comparable to those obtained for a UASB reactor treating for domestic sewage. High removal efficiencies were obtained in a much shorter time compared to UASB reactors treating domestic sewage. The COD concentration in the effluents is too high for direct discharge and hence a form of post-treatment would be necessary. The calculated volume of methane in the biogas collected ranged from 4 – 8 l/kg COD, not accounting for practical losses.

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Abbreviations

BOD	Biochemical Oxygen Demand
COD	Chemical Oxygen Demand
EAWAG	Swiss Federal Institute for Environmental Science & Technology
EGSB	Expanded Granular Sludge Blanket
FSTP	Faecal Sludge Treatment Plant
HRT	Hydraulic Retention Time
ISE	Ion Selective Electrode
KNUST	Kwame Nkrumah University of Science and Technology
KVIP	Kumasi Ventilated Improved Pit
OLR	Organic Loading Rate
SANDEC	Department of Water and Sanitation in Developing Countries
TKN	Total Kjeldahl Nitrogen
TP	Total Phosphorus
TS	Total Solids
TSS	Total Suspended Solids
TVS	Total Volatile Solids
UASB	Upflow Anaerobic Sludge Blanket
VIP	Ventilated Improved Pit
VSS	Volatile Suspended Solids
WC	Water Closet
WRI	Water Research Institute
WRRI	Water Resources Research Institute

Chapter One

1 Introduction

1.1 Background

In Ghana, one of the major environmental pollution problems faced is the indiscriminate disposal of a major percentage of nightsoil, toilet sludge and septage collected from urban cities and towns by vacuum tankers on to land and into water bodies (Plates 1 and 2). The nightsoil arises mainly from household and communal bucket latrines (Plates 3 and 4), whilst the toilet sludge is from household and public toilets, which may be non-flush aqua-privies (Plates 5, 6 and 7), single/multiple pit VIPs (Plates 8 and 9). The septage is from household and communal septic tanks connected to water closets (Plate 10).

Improper disposal of human excreta results in the contamination of water bodies, soil and food crops. This practice poses a serious health hazard because human excreta is the principal source of pathogenic organisms, which may be transmitted by direct contact, contaminated water and food, insects and other vectors. Human excreta ***must*** therefore be treated before its ultimate disposal into the environment or its use in agriculture in order to:

a. reduce the spread of communicable diseases caused by excreted pathogenic organisms; and
b. prevent the pollution of the environment, water sources and soil.

Recognising the environmental deficiencies that result from the indiscriminate disposal of nightsoil and faecal sludges, concerted efforts were put in to find treatment systems with technology suited to the socio-economic conditions of Ghana to treat and safely dispose of the faecal wastes. Research activities to find appropriate treatment options thus began in earnest in Accra, the capital city of Ghana, from 1986 under a waste management improvement project for the city.

In 1989/90, while these research activities were ongoing, a prototype faecal sludge treatment plant (FSTP) consisting of a solids-liquid separation step in settling/

Plate 1 A vacuum tanker discharging faecal sludge onto the ground to flow into a nearby stream

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Plate 7 **Aqua privy – squat hole**

Plate 8 **A public toilet: multiple pit VIP latrine locally called the KVIP**

Plate 9 Squat hole arrangement in the KVIP

Plate 10 A typical household septic tank at the end of the drive way with vent pipes showing

thickening tanks, followed by a series of four waste stabilisation ponds, a trickling stack, a "maturation pond" and a series of evaporation beds was built at Achimota, north of Accra to:

- reduce the incidence of unauthorised dumping and the cost of haulage from the north of Accra to the beach, and
- obtain adequate operational data for system components of a prototype plant.

The solids separated in the settling/thickening tanks are used for composting while the liquid fraction is treated in the ponds following the settling/thickening tanks. A schematic diagram of the prototype FSTP is shown in Figure 1.1.

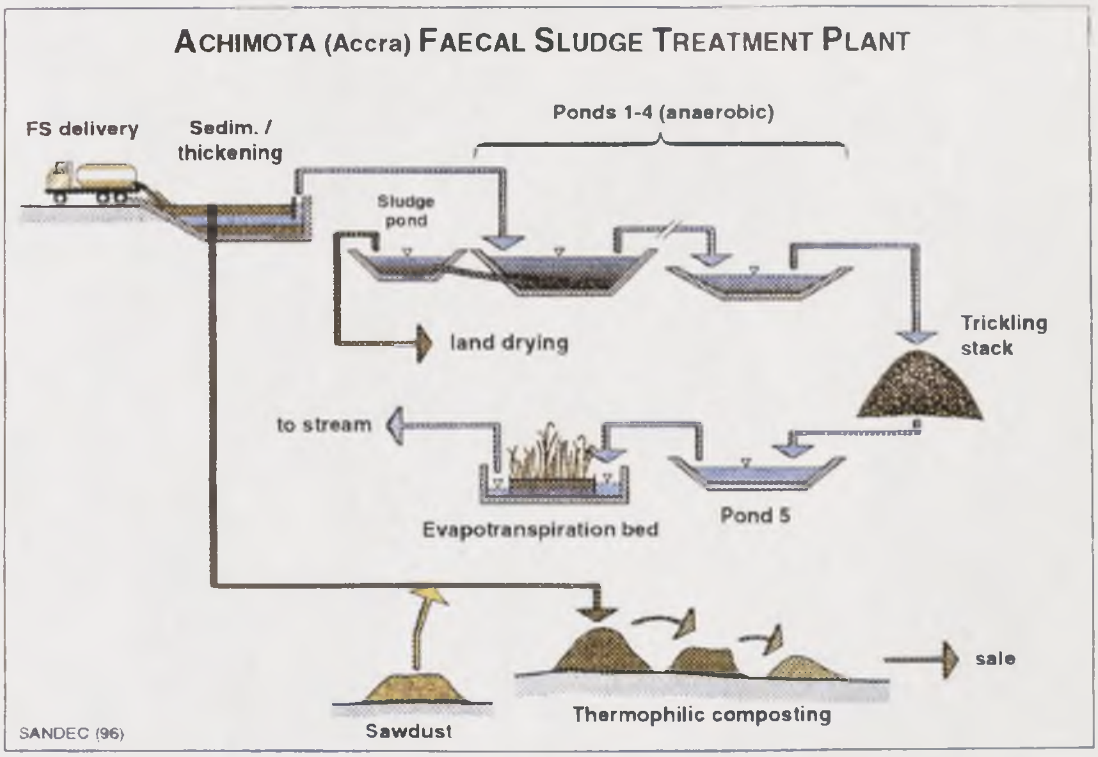


Figure 1-1 Schematic diagram of the Achimota Faecal Sludge Treatment Plant

Source: Heinss et al. (1998)

The building of this plant made Ghana "one of the first countries to set up and operate plants for the separate treatment of sludges from septic tanks, bucket latrines and public toilets" (Heinss *et al.* 1998). Since then three other FSTPs have been constructed to test new conceptual designs in an attempt to improve upon operational performance. Under an ongoing urban environmental sanitation project, which is partly aimed at improving excreta management by providing facilities for the treatment of nightsoil and faecal sludge, new FSTPs are being planned for the country's five major cities.

Between 1993 and 1997, the Achimota plant was monitored and evaluated under a collaborative research programme between the Water Research Institute (WRI) [formerly Water Resources Research Institute (WRRI)] of Ghana, and the Department of Water and Sanitation in Developing Countries (SANDEC) of the Swiss Federal Institute for Environmental Science & Technology (EAWAG). The purpose of the research programme was to provide additional and adequate data to enable the preparation of design and operational management guidelines for faecal sludge treatment. The field research results from the monitoring and evaluation of FSTPs in the tropics (Heinss *et al.* 1998) including the Achimota plant, while proposing preliminary deign guidelines, also indicate that there are still some teething problems associated with the existing plant. Thus there is the need for more applied research to consolidate the results obtained to date and to explore other sustainable treatment options.

1.2 Problem Statement

The environmental deficiencies and health hazards that result from the indiscriminate disposal of human excreta or their use in agriculture are well documented. The practice, however, continues in Ghana and many other developing countries. This could be attributed to a number of factors, namely:

- the absence of tested and suitable technologies to treat faecal sludges in developing countries;
- a lack of treatment plants and that fact that only a few have been built to date resulting in an overall shortage of treatment capacity to handle the volume of faecal wastes generated; and
- poor operation and maintenance of existing treatment facilities leading to malfunctioning and eventually their breakdown.

In a developing country like Ghana, treatment facilities should:

a. be technologically appropriate, i.e. low-cost both in capital and operating costs, simple to construct, operate and maintain (compatible with available expertise);
b. need little or no imported equipment;
c. not be energy-intensive; and
d. be able to treat the wastes to at least secondary level, with emphasis on the removal of pathogens and helminth eggs.

As already stated previously, in Ghana, there are four FSTPs in operation and these satisfy the above criteria for treatment facilities in a developing country. However, as shown by the collaborative research results (Strauss *et al.*. 1997; Heinss *et al.* 1998), the nature of the wastes (Table 1.1) is such that current FSTPs alone cannot be relied upon for their effective treatment.

Table 1.1 Characteristics of Faecal Sludges from On-site Sanitation Systems in Accra, Ghana

Parameter	Public Toilet Sludge	Septage
BOD (mg/l)	8,800 (3,800 - 15,000)	630 (360 - 1,300)
COD (mg/l)	47,600 (10,400 - 97,000)	8,500 (820 - 52,000)
TS (%)	-	1.4 (0.3 - 11.4)
TVS (% of TS)	62	63
TSS (%)	6.4 (2 - 19)	0.7 (0.07 - 3.4)
VSS (% of TSS)	58	70
Helminth eggs (no./l)	29,000 (3,600 - 62,000)	4,300 (200 - 13,000)

Source: WRRI/SANDEC, 1994

The results from the monitoring and evaluation of the Achimota FSTP by WRRI/SANDEC indicate that the current system could be effective in the treatment of the septage because the solids easily separate from the liquid in the settling/thickening tanks ensuring that only the liquid fraction flows into the pond system for treatment. The FSTP can also, to some extent, treat mixtures of nightsoil/toilet sludge and septage (the mixtures containing higher proportions of septage) if operational and design guidelines are adhered to. With regards to the nightsoil/toilet sludges the results from the monitoring and evaluation of the

Achimota FSTP indicate that the nightsoil/toilet sludges are hardly conducive to solids-liquid separation, the first step in the present FSTP design. This results in the faecal sludge flowing into the pond system and causing the system to fail from overloading due to high organic strength and high concentrations of ammonia. Heinss et al. (1998) attribute the lack of solids-liquid separation to the fact that the nightsoil/toilet sludges are mostly fresh i.e. undigested and highly concentrated compared with the septage. The solids-liquid separation can be improved by digesting the nightsoil/toilet sludges. The digestion of the high-strength nightsoil/toilet sludges using anaerobic treatment processes with the intent of reducing its organic strength and improving the solids-liquid separation is the major purpose of this research.

1.3 Objective of Study

1.3.1 Overall Objective

The problems previously enumerated are happening because at the moment there is very little experience with regards to the use of technologies considered suitable to the socio-economic conditions of developing countries in the treatment of nightsoil/toilet sludge. The overall objective of this research was to investigate how nightsoil/toilet sludge could be treated anaerobically using the upflow anaerobic sludge blanket (UASB) reactor.

1.3.2 Specific Objectives

The specific objectives of the research were to:

a. establish some physical and chemical characteristics of nightsoil/toilet sludge in Kumasi, the second largest city in Ghana, and thereby contribute towards the improvement of the statistical significance of existing data;

b. build and operate, on a pilot-scale, an UASB treatment system, determining optimum loading rates and other essential operating parameters necessary for the anaerobic digestion of nightsoil/toilet sludge; and

c. identify realistic gas utilisation potentials for the gas produced.

1.4 Choice of the UASB Reactor

The Upflow Anaerobic Sludge Blanket (UASB) reactor was selected for this research after initial consultation on, and examination of the nature of the problem to be addressed. The selection was also influenced by the fact that at the start of the research, an UASB reactor was being built as part of a new sewage treatment works (the Korle Lagoon Sewage Treatment Works) for Accra by Taylor Woodrow International under the Accra Waste Project financed by the Department for International Development (DFID). At present the construction of this UASB reactor has been completed and the reactor commissioned.

Although the UASB reactor has been designed primarily to treat raw sewage, adequate provision has been made in the design for 40 m^3/d (about five tanker loads) of faecal sludge to be added to the raw sewage. The findings of this research would have a direct impact on the amount of faecal sludge that can be added to the raw sewage without the risk of overloading the treatment plant. Positive findings from this research could lead to increasing the daily inflows of faecal sludge and thereby reducing the amount of faecal sludge discharged untreated and also raises the possibility of having additional UASB plants for faecal sludge treatment.

As mentioned previously, the effectiveness of the current FSTP could be improved if the faecal sludge is pre-treated to improve the solids-liquid separation and reduce its organic strength. With the increased possibility of building new FSTPs for the other five major cities, including Kumasi, it is intended that the UASB reactor would be used for the pre-treatment prior to the FSTP.

1.5 Organisation of the report

The report is divided into seven sections. In the first chapter the background and problem statements are highlighted. Also presented in Chapter 1 are the objectives of the research and the choice of the UASB reactor for the study is presented.

The second chapter deals with the literature review on the research subject of faecal sludge treatment. The process of anaerobic digestion and the applicability of the UASB reactor for the treatment of faecal sludges are also presented.

In the third section, the materials and methods used for the experimental study are presented. Included in this section are the various methods used in the laboratory analysis and the experimental set-up for the two experiments carried out in Leeds and Kumasi.

The fourth chapter covers the results obtained from the various experiments and laboratory analysis. In Chapter 5, the results obtained from the experiments are discussed and compared to results obtained in similar experiments using the UASB reactor.

The conclusions for the research work are presented in Chapter 6. In the last section, Chapter 7, recommendations for future work are presented and Appendix 1 presents a design example for the city of Kumasi, Ghana.

Chapter Two

2 Literature Review

2.1 Introduction

In Ghana and most developing/newly industrialised countries, disposal of human excreta is through on-site sanitation systems and not by water-flush toilets connected to centralised sewerage systems. Whilst the treatment of wastewater from centralised sewerage systems by conventional treatment systems are well developed in the western and industrialised countries and hence the abundance of literature on them, the same cannot be said of the treatment of nightsoil and toilet sludge from on-site sanitation systems. Relatively, very little has been published to date on the treatment of nightsoil and toilet sludge, especially utilising low-cost technology appropriate to the needs of developing countries (Pescod, 1971; Heinss *et al.* 1998). It is thus desirable to carry out a review of the literature to gather the information available to date on the treatment of nightsoil and toilet sludge with the view of learning from what has been and is being done, and to serve as a basis for the present research.

2.2 Definitions: nightsoil, toilet sludge, septage and faecal sludge

2.2.1 Introduction

It is essential to define the terms "nightsoil", "toilet sludge", "septage" and "faecal sludges" as used in this literature review since they are the focus of the research. The definitions are also essential to establish consistency in the use of the terms throughout this literature review, eliminating any ambiguity in their use as it appears in the literature, especially with "nightsoil". Furthermore, the definitions will be necessary to distinguish between the various terms as they are used in reference to human body wastes.

2.2.2 Nightsoil

In the literature, both terms "nightsoil" as one word and "night soil" as two words are used interchangeably. In this review, "nightsoil" will be used. This term is mostly used to represent, in general, a mixture of human faeces and urine. In certain instances, the term is also used to represent a mixture of human faeces and urine that has undergone some considerable putrefaction. Pradt (1971), Mara (1976), Satyanarayan *et al.* (1987) and Choi *et al.* (1997) all use the term for a mixture of human faeces and urine. Mara and Caincross (1989) use the term for "a mixture of human faeces and urine transported without flushing water". Caincross and Feachem (1993) state that "nightsoil comprises only faeces and urine plus small volumes of water if it is used for anal cleansing and pour-flushing". Stoll and Parameswaran (1996) refer to the contents from septic tanks/leaching pits which are connected to pour flush latrines in Bangkok as nightsoil. Choi *et al.* (1996) use the term for contents of cesspool and holding tanks/storage pits in Korea where it is stored for more than three months before being collected.

Some of the listed references indicate how loosely the term nightsoil is used by some authors. It is desirable to use the term consistently and in a specific manner in standard technical texts and literature. This would help in the search for and sharing of information from research findings. In this review, nightsoil is used for "a mixture of human faeces and urine that is mostly fresh".

2.2.3 Toilet Sludge

In conventional wastewater treatment, the objectives are accomplished by concentrating the impurities into solid and semisolid residuals and then separating them from the bulk liquid. The concentration of the solid and semisolid residuals is referred to as sludge (Peavy *et al.* 1985; Metcalf & Eddy, Inc., 1991). The solid and semisolids residuals are removed from the bulk liquid after primary and/or secondary treatment. The sludge has thus undergone some measure of treatment, although it may be inadequate for its ultimate disposal.

In Ghana and developing countries, where conventional sewerage and treatment systems are mostly absent, various forms of on-site sanitation facilities are utilised. These on-site sanitation facilities could be either water dependent, e.g. pour flush, water closet and aqua privies, or non-water dependent e.g. bucket, ventilated improved pit (VIP) and vault latrines. The human body wastes may be stored in these on-site sanitation facilities for a couple of days to several years depending on the type of facility, its storage capacity, emptying frequency and the collection/ transportation system is use.

For the non-water dependent systems, when the mixture of faeces and urine is stored for a couple of days (up to 3 days) as in the case of bucket latrines, the contents may still be fresh and hence would be referred to nightsoil. However, when nightsoil is stored for longer periods as in the case of "dry" aqua-privies, watertight vented and non-vented toilets with no or low flush water, and holding tanks, the nightsoil undergoes some digestion (little or partial). This category of little or partially digested nightsoil, i.e., nightsoil that has undergone some measure of treatment, collected from the non-water dependent systems is referred to as "toilet sludge" throughout this document.

2.2.4 Septage

As mentioned in the previous section, some of the on-site sanitation facilities at homes, offices, commercial houses and institutions are water dependent. In the water dependent on-site sanitation facilities, the human excreta is flushed out using water. The resulting wastewater (mixture of flush water, faeces and urine) is discharged into septic tanks, where the solid fraction settles out and undergoes anaerobic digestion. The effluent from the tank is usually discharged into a subsurface-soil absorption system for final treatment and disposal. The combination of the sludge produced in the septic tank as a result of the anaerobic digestion of the settled solids, scum and liquid pumped from as septic tank is known as septage.

2.2.5 Faecal Sludge

The collection and transportation of nightsoil, toilet sludge and septage from their various sources to the final treatment/disposal sites is done by vacuum trucks. Depending on their capacities, haulage distances and socio-economic conditions, the vacuum trucks may carry loads of only toilet sludge, septage or mixtures of both. Most often, the contents of the trucks are mixtures of both toilet sludge and septage and hence difficult to distinguish between the wastes that arrive at the treatment/disposal sites. It is thus appropriate to have a term that would include all types of faecal wastes coming from the on-site sanitation systems that have undergone some measure of digestion. The term faecal sludge, as used by Strauss *et al.* (1997) and Heinss *et al.* (1998), is used in this review for "all sludges (little or partially digested) collected and transported from on-site sanitation systems by vacuum trucks".

2.3 Quantities, Characteristics and Classification of Faecal Sludges

2.3.1 Per Capita Quantities of Faecal Sludges

As with wastewater, determining the quantities of faecal sludges is an essential and fundamental step in the planning and design of collection, treatment and disposal facilities. Reliable data on faecal sludge quantities are needed if the facilities are to be designed properly with adequate capacities. As reported by Martin *et al.* (1997), the per capita quantities reported in the literature vary widely for both nightsoil and faecal sludges. Table 2.1 shows the overall averages of the daily per capita volumes and constituent contributions in nightsoil and faecal sludges in Accra, Ghana.

Table 2.1 Daily Per Capita BOD, TS, TKN Quantities of Different Types of Faecal Sludges

Parameter	Septage [1]	Public toilet and bucket latrine sludge [1]	Fresh excreta
BOD g/cap. day	1	16	45
TS g/cap. day	14	100	110
TKN g/cap. day	0.8	8	10
l/cap. day	1	2 (includes water for toilet cleansing)	1.5 (faeces and urine)

Source: Heinss *et al.* (1998).

[1] Estimates are based on faecal sludge collection survey conducted in Accra, Ghana.

TS (total solids = residue after evaporation at 103 °C)

The daily per capita volume for septage in Table 2.1 (1 l/cap. day) is rather low considering that, on a per capita basis, more water is used in flushing water closets than in public toilets and bucket latrines. Heinss et al. (1998) do not offer any explanation for this rather low figure.

2.3.2 Characteristics of Faecal Sludges

An understanding of the nature of faecal sludges is essential in the design and operation of collection, treatment and disposal facilities. Characterisation of the physical, chemical and biological composition of faecal sludges is a major step in understanding their nature. However, compared with wastewater and sludges generated in the western and industrialised countries, very little has been done to date to characterise faecal sludges from on-site sanitation systems in developing countries. This is mainly due to the fact that in developing countries, analytical techniques for the assessment of the characteristics of wastes and waste treatment plants are not routinely applied as pertains in western and industrialised countries. There are no selected variables on which to base the assessment and also a lack of standard methods of analysis suited for developing countries.

As noted by Pescod (1971), Pradt (1971), Um and Kim (1986), Guo *et al.* (1991) and Strauss *et al.* (1997), the characteristics of collected faecal sludges vary greatly. The characteristics depend, among others, on the season, type of the on-site sanitation system (e.g. water closet/septic tank system, "dry" aqua privy, watertight vented pit latrines), the emptying frequency (i.e. is the retention time in the facility), the extent of stormwater or groundwater infiltration into the sanitation facility, and on user habits. This section presents and discusses the characteristics of faecal sludges as contained in the literature to date.

2.3.2.1 Septage

As with all faecal sludges, and noted by Metcalf & Eddy, Inc. (1991), the actual quantities and constituents of septage vary widely, the greatest variation being found in communities that do not regulate the collection and disposal of septage. Undoubtedly, this is the situation which occurs in a developing country like Ghana. In the industrialised countries, where the collection and disposal of septage is well regulated, septage is well characterised compared to developing countries. Furthermore very few data exist on the characteristics of septage in developing countries. Typical constituents found in septage in industrialised countries and Accra, Ghana, are given Tables 2-2.

Table 2.2 Typical characteristics of septage

Constituent	Concentration, mg/l			
	Metcalf & Eddy, Inc. (1991)		Strauss and Heinss (1995) (Accra, Ghana)	
	Range	Typical	Range	Typical
TS	5,000 - 100,000	40,000	3,000 - 114,000	14,000
TSS	4,000 - 100,000	15,000	700 - 34,000	7,000
VSS	1,200 - 14,000	7,000	490 - 23,800	4,900
BOD_5	2,000 - 30,000	6,000	360 - 1,300	630
COD	5,000 - 80,000	30,000	820 - 52,000	8,500
TKN as N	100 - 1,600	700	-	-
NH_3 as N	100 - 800	400	-	-
TP as P	50 - 800	250	-	-
Heavy metals [a]	100 - 1,000	300	-	-
Helminth eggs, (no./l)	-	-	200 - 13,000	4,300

[a] Primarily iron (Fe), zinc (Zn), and aluminium (Al);

Table 2.2 shows a wide range of variation of the constituents. The typical values for Accra with tropical climate are, however, much lower than the typical values reported by Metcalf & Eddy, Inc. (1991), mostly for industrialised countries located in temperate climates. For example the BOD_5 reported by Metcalf & Eddy, Inc. ranges from 2,000 – 30,0000 mg/l with a mean value of 6,000 mg/l while that reported by Strauss and Heinss (1995) range from 360 – 1,300 mg/l with a mean value of 630 mg/l. The lower values for Accra could be partly attributed to:

1. the higher degree of mineralisation in the tropical septage due to higher temperature conditions associated with tropical climates among others.
2. when desludging septic tanks, vacuum tanker operators have the tendency of pumping only the top clarified liquid portion.

It is established that the rate of anaerobic digestion, which could be measured by the rate methane fermentation, is higher at high temperatures (Table 2.3). Unheated anaerobic treatment systems would therefore perform better in tropical climates than temperate climates because ambient temperatures are much higher. Table 2.3 shows that the rate of methane fermentation at 25°C or more is at least twice the rate which occurs when the temperature 15°C or less, the temperature being the temperature at which the fermentation is taking place.

Table 2.3 Estimated effect of temperature on anaerobic treatment

Temperature (°C)	Rate of methane fermentation relative to that at 35°C
5	0.1
15	0.4
25	0.8
35	1.0

Source: McCarthy (1966) [Cited by Mara and Sinnatamby, 1986]

The high degree of mineralisation of tropical septage influences the characteristics of septage and ultimately affect the type of treatment/disposal facilities that could be utilised. Mara *et al.* (1992) for instance argue that "anaerobic ponds are of no purpose as septage is already highly mineralised", while Strauss *et al.* (1997) are of the view that "a first anaerobic stage might offer advantages even at moderate BOD and TVS reduction efficiencies, and may result in smaller land requirements than by directly feeding the septage into a facultative pond".

2.3.2.2 Nightsoil and Toilet Sludge

The characteristics of both nightsoil and toilet sludge (mainly from unsewered toilets) are presented in this section. The two have been merged because the literature hardly distinguishes between the characteristics of nightsoil and toilet sludge.

Generally, nightsoil and toilet sludge are more concentrated than septage because in majority of cases no flushing water is used. Even in cases where water is used (e.g. pour-flush latrines) the quantities are small compared to conventional water closet toilets. Table 2-4 shows reported analysis of collected nightsoil and toilet sludges in the literature. It is evident from the table there is no uniformity with respect to the constituents reported. Various authors report on parameters relevant to their particular research. There is no minimum set of variables reported by each author. A minimum set of variables for the assessment of faecal sludges has been proposed by

Strauss *et al.* (1997). Table 2.5 contains the minimum set of variables proposed by Strauss et al. (1997) to assess untreated faecal sludges as well as liquids and sludges formed during faecal sludge treatment. Variables for FSTP monitoring and control are also listed in Table 2.5. Depending on the specific requirements, additional variables would have to be added.

Table 2.4 Characteristics of nightsoil/toilet sludge

Country	Japan	Japan	Korea	Korea	Korea	Korea
	Concentration (mg/l except pH and Helminth eggs)					
Parameter	1	2	3	4	5	6 (average of 10 samples)
pH	8.5	-	7.2	7.2 - 7.9	8.2 - 8.8	-
BOD_5	10,190	12,900	22,100	12,600 - 19,200	16,800 - 22,900	13,400 - 19,000 (16,000)
$COD_{(Mn)}$	-	-	-	10,600 - 15,400	-	-
$COD_{(Cr)}$	-	36,700	64,700	-	38,600 - 44,600	34,700 - 63,900 (50,200)
TS	30,100	31,400	45,100	32,000 - 44,600	20,500 - 38,400	-
TVS	17,600	20,400	-	-	-	-
TSS	12,000	-	35,400	14,000 - 26,700	7,100 - 10,600	7,000 - 15,300 (11,300)
VSS	-	-	-	-	-	6,500 - 13,000 (9,800)
TKN	-	-	4,300	-	-	-
NH_4-N	3,471	-	-	-	-	-
Cl^-	4,671	-	4,100	3,800 - 5,600	-	-
PO_4^{3-} - P	-	-	650	1,050 - 1,600	-	-

1. Iwai *et al.* (1962). The figures represent an average of analyses of ten samples of collected nightsoil
2. Ikeda (1965, 1966, 1968). Figures represent the average daily analyses for a seven day run at a test nightsoil processing plant at Yokohama.
3. Choi (1985).
4. Um and Lee (1982). Values measured from nightsoil comminuted into 5 mm particles and filtered.
5. Um and Choi (1984). Values measured from nightsoil comminuted into 5 mm particles and filtered.
6. Kim and Lee (1986). Values calculated from values of diluted samples used in experiment.

Table 2.4 Characteristics of nightsoil/toilet sludge (continued)

Country	Japan	Japan	Japan	Japan	Ghana	China
	Concentration (mg/l except pH and Helm. eggs)					
Parameter	7	8	9 (average)	10	11	12
pH (units)	7.61	8.48	-	-	-	-
BOD$_5$	8,034	8,551	10,500 - 12,900 (11,600)	12,000	3,800 - 15,000 (8,800)	15,000 - 18,000
COD$_{(Mn)}$	4,966	6,356	3,340 - 8,200 (5,080)	5,900	-	-
COD$_{(Cr)}$	25, 118	-	-	26,000	10,400 - 97,000 (47,600)	26,000 - 33,000
TS	18,150	26,574	-	-	-	12,000 - 30,000
TVS	10,107	-	-	-	62% of TS	-
TSS	7,140	14,417	-	13,000	20,000 - 190,000 (64,000)	-
VSS	5,812	-	-	-	11,600 - 110,000 (37,000)	-
TKN	-	4,413	-	3,700	-	5,000 - 6,000
NH$_4$-N	2,430	2,979	2,100 - 3,140 (2,550)	3,010	-	-
NO$_2$-N	-	500	-	-	-	-
NO$_3$-N	-	132	-	-	-	-
Cl$^-$	-	4,386	2,300 - 3,800 (2,870)	-	-	-
PO$_4^{3-}$ - P	-	810	-	-	-	-
Hem. Eggs (no./l)	-	-	-	-	3,600 - 62,000 (29,000)	18,000 - 360,000

7. Noike and Matsumoto (1986)
8. Murata *et al.* (1986)
9. Suzuki and Tohya (1986)
10. Murakami *et al.* (1986)
11. Strauss and Heinss (1995)
12. Shiru and Bo (1991)

Table 2.4 Characteristics of nightsoil/toilet sludge (continued)

Country	Korea	Korea	Japan	Japan	Tropical Countries
	Concentration (mg/l except pH and Helm. eggs)				
Parameter	13	14	15	16	17
pH (units)	-	7.9	8.0 – 8.6	8.2	-
BOD_5	-	19,000	11,000 – 14,000	12,130	COD:BOD 2:1 - 5:1
$COD_{(Mn)}$	-	-	-	4,305	-
$COD_{(Cr)}$	45,800	48,000	-	-	20,000 - 50,000
TS	-	-	-	26,120	$\geq$35,000
TVS	-	-	-	11,880	-
TSS	33,500	25,000	14,000 – 20,000	-	-
VSS	26,700	21,000	-	-	-
TKN	4,480	6,000	4,200 – 5,200	4,335	-
NH_4-N	3,260	4,800	-	3,626	2,000 - 5,000
Cl^-	-	-	3,200 – 4,200	-	-
T-P	810	1,000	480 – 680	-	-
PO_4^{3-} - P	630	600	-	-	-
Hem. Eggs (no./l)	-	-	-	-	20,000 - 60,000

13. Choi *et al.* (1996).
14. Choi *et al.* (1997).
15. Misaki and Matsui (1996) – the characteristic is dependent on the quality of human waste, the higher the percentage of collected human waste, the higher the characteristic.
16. Iwai *et al.* (1964).
17. Heinss *et al.* (1998)

Table 2.5 Minimum Set of Variables for FS and FSTP Assessment

Variables to be Assessed by Laboratory Analyses (Raw sludge and performance assessment)	**Variables to be Assessed by Field Measurements or Observations** (Process and operational control)
• TS (total solids = residue after evaporation at 103 °C) • Volume of settleable and floatable solids • Dewaterability and filterability tests (suitable tests still to be defined) • COD (chemical oxygen demand) (non-filtered and filtered) • BOD (biochemical oxygen demand)* • NH_4N • Helminth Eggs • Faecal coliforms	• Volume of settleable and floatable solids in 1 or 2 litre cylinders DO (dissolved oxygen) • pH • Colour check for algal growth • Microscopic examination (e.g. for pond organisms) • Temperature (in thermophilic composting) • Settled sludge and scum thickness • Sludge thickness in drying beds • Weather data

*Only if samples can be properly treated and standard analytical techniques adhered to.
Source: Strauss et al. (1997)

Almost all the data reported in the literature originate from Asia and tropical developing countries. The majority of the reported data are from Japan and South Korea. Values reported in Table 2.4 show nightsoil/toilet sludge contains very high concentrations of solids, with over 30% being suspended solids. The total volatile solids and volatile suspended solids constitute over 50% of total solids and over 60% of suspended solids respectively. This indicates a high organic content of the nightsoil/toilet sludge. The COD to BOD ratios range from 1.5:1 to 5:1. Thus of the organic content a considerable portion is biodegradable. The reported pH values in Table 2.4 ranges from 7.2 to 8.8 and the ammonium (NH_4-N) concentrations are high ($\geq$ 2,000 mg/l). The number of helminth eggs is also high in the three reported cases (ranges from 3,600 – 360,000 eggs./L) indicating the unsafe nature of untreated nightsoil/toilet sludge.

2.3.3 Classification of Faecal Sludges

Faecal sludges, just as wastewater, can be classified depending on the concentrations of the constituents. However, throughout the literature, very little has been done on the classification. Most authors rather stress the fact that characteristics of faecal sludges differ greatly from municipal wastewater collected in centralised sewerage systems. As is evident from Tables 2.2 and Table 2.4, faecal sludges could be 10 to over 100 times more concentrated than municipal wastewater. Strauss *et al.* (1997) have classified faecal sludges into two broad categories: high-strength and low-strength (Table 2.6).

Table 2.6 Important Characteristics and Classification of Faecal Sludges

Item	**High-strength**	**Low-strength**	**Sewage** (for comparison)
Example	Public toilet or bucket latrine sludge	Septage	Tropical Sewage
Characterisation	Highly concentrated, mostly fresh FS, stored for days or weeks only	FS of low concentration; usually stored for several years; more stabilised	
COD mg/l	20,000 - 50,000	< 10,000	500 - 2,500
COD:BOD	2:1 - 5:1	5:1 - 10:1	2:1
NH_4-N mg/l	2,000 - 5,000	< 1,000	30 - 70
TS	≥ 3.5%	< 3%	< 1%
TSS mg/l	≥ 30,000	≅7,000	200 - 700
Helminth eggs /l	20,000 - 60,000	≤ 4,000	300 - 2,000

Source: Strauss *et al.* (1997)

The classification is based on the concentrations of organics, ammonium, solids, and the degree of putrefaction. Thus the high-strength faecal sludges, as described in Table 2.6, are rather fresh and exhibit high concentrations of organics, ammonium and solids. These originate from on-site sanitation systems consisting mainly of bucket latrines, aqua privies, KVIPs and pour flush toilets. The low-strength faecal sludges are relatively weak, older and have undergone considerable digestion. Septage falls into this category.

2.4 Collection and Treatment of Nightsoil and Faecal sludge

2.4.1 Collection

2.4.1.1 Nightsoil

Traditionally, nightsoil is collected from buckets or nightsoil vaults situated immediately below the toilet by conservancy labourers. In Ghana, nightsoil is collected from both household or communal bucket latrines. The nightsoil collected by conservancy labourers is dumped either at authorised sites which may be holding tanks at designated sanitary sites or at unauthorised sites such as open drains, open refuse dumps and nearby bushes. This mode of collection has become increasing unpopular because of the health hazards posed to both the conservancy labourers and the general public, and nuisance resulting from such dreadful collection practices. In Ghana, bucket latrines are now being converted into other low-cost on-site sanitation systems which eliminates the need for manual collection as described above. Nightsoil holding tanks, depending on their storage capacity and whether they are permanent or temporary structures, are emptied daily, after several weeks or even months by vacuum trucks and taken to final treatment/disposal sites.

2.4.1.2 Toilet Sludge and Septage

Both toilet sludge and septage from on-site sanitation facilities are collected by truck tankers fitted with vacuum pumps. In Japan, Pradt (1971) and Misaki and Matsui (1996) report that nightsoil accumulates in concrete vaults in individual houses in unsewered parts of the country. The toilet sludge is then collected every 3 - 4 weeks by a “vacuum car”, which is small tank truck with a capacity of 2-4 m^3 and equipped with a long 100 mm hose and a vacuum pump. Similar systems are used in most countries where toilet sludge and septage have to be collected from either individual houses or public on-site sanitation facilities. This system of collection, which is more hygienic, avoids the hazards from spillage and direct contact with human excreta. Figure 2-1 shows the flow of nightsoil, toilet sludge and septage in Ghanaian towns and cities.

2.4.2 Treatment

Various methods/processes used for the treatment faecal sludges from on-site sanitation facilities have been mentioned in the literature. Countries in Asia appear to be at the forefront in the development of treatment methods for faecal sludges, with Japan taking a leading role (Pradt, (1971); Misaki and Matsui, (1996)). The treatment methods that have been and are being used have been dictated by the treatment goals and objectives. Examples of the treatment methods in use are presented in Table 2.7 and Figures 2.2 – 2.9.

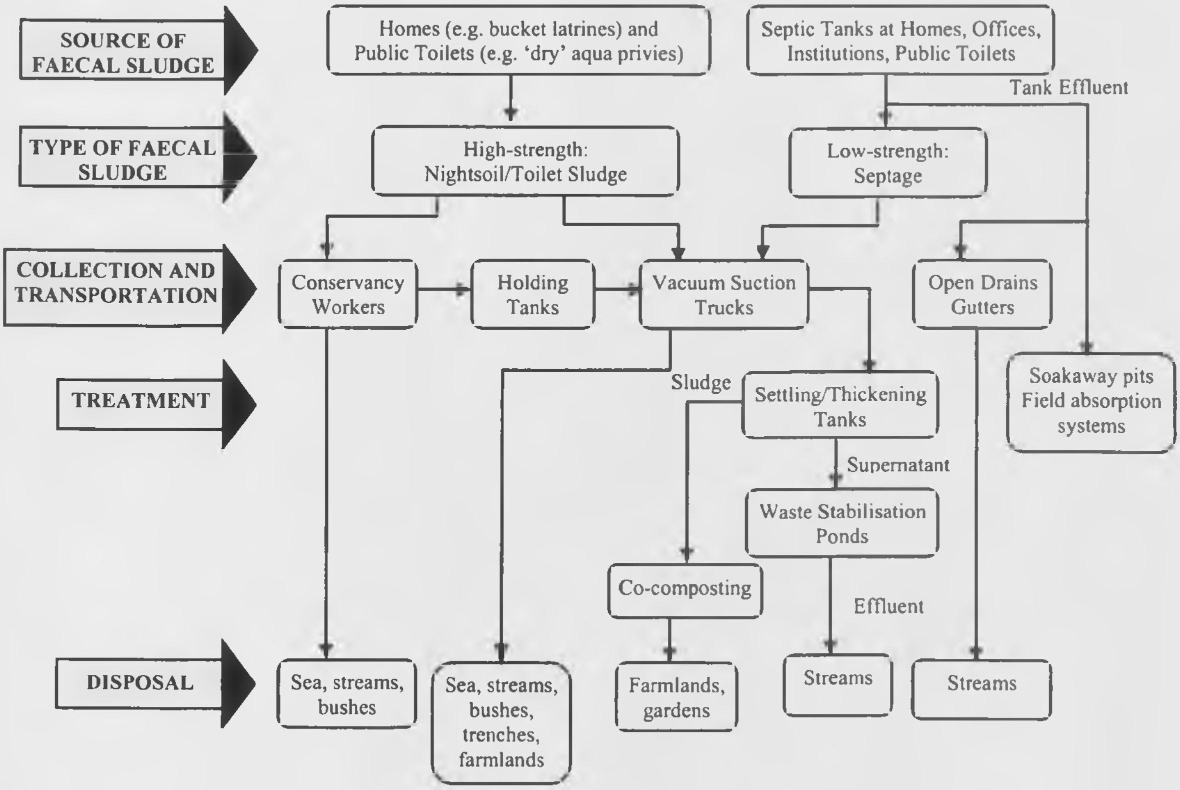


Figure 2-1 **Schematic diagram of current situation showing sources, types, collection, transportation, treatment and disposal of nightsoil and faecal sludges in Ghana.**

Table 2.7 shows that technologies for the treatment of sewage sludge in industrialised countries have been modified and further developed, and used successfully in Southeast Asia, with Japan being at the forefront. As evident from Table 2.7, information from the Western World with regards to nightsoil treatment is lacking compared to information on conventional treatment methods. This has been attributed to the influence of the conquering Roman Legions who brought with them the technology of waterborne sewerage to the far reaches of their Empire (Pradt, 1971). In fact Pradt (1971) notes that "the Japanese are exclusively responsible for the development of a sophisticated body of night soil treatment technology." Both Japan and Korea have design manuals for nightsoil treatment systems. Most of the treatment plants are equipped with conventional anaerobic or aerobic digestion for primary treatment. These involve high capital investment for equipment, use of energy intensive mechanical equipment, high costs of operation, and the need for skilled operators.

Table 2.7 Examples of Faecal Sludge Treatment Systems

Country: Japan		
Type of Process	**Reference**	**Remarks**
Conventional digestion process **(Figure 2-2)** • Conventional digesters used for sewage sludge digestion, however many of the digesters being unheated with holding periods of up to 90 days. • Aerobic treatment, usually trickling filters and the activated sludge process, are provided for the digester supernatant, after diluting it 20-40 times with fresh water. • The digested residue is dewatered by filter press, vacuum filter, centrifuge or sand bed • Cake is hauled away for fill or fertiliser, or sometimes incinerated	Pradt (1971)	High installation cost
Chemical treatment process **(Figure 2-3)** • Chemical, including calcium hydroxide, alum, ferric chloride and ferrous sulphate, are added to precipitate and coagulate the solubilised solids • Supernatant and sludge from a thickener are treated as described for the conventional digestion process. • Considerable sludge is produced in this method compared to conventional digestion process.	Pradt (1971)	Supernatant is more easily treated by aerobic methods than digester supernatant. Cost less to install, but highest operating costs. Used only as temporary facilities.
Bio-oxidation treatment process **(Figure 2-4)** • Pre-treatment by screening, settling or centrifugation • Dilution water is added to the centrate and then treated by conventional activated sludge process. • The solids slurry is first treated by chemical flocculation and dewatered by sand bed or mechanical filter	Pradt (1971)	Installation cost less than that of a digester; operating cost about same level as a digester.
Wet-air oxidation heat treatment system **(Figure 2-5)** • Involves heating of faecal sludge for short periods of time under pressure • Water scrubber is used to clean gas released before being discharged into the atmosphere • Gravity settling is used to remove the residual solids. • Supernatant liquor is diluted with fresh water and treated by conventional activated sludge process.	Pradt (1971)	Plant more expensive to install and operate. Discontinued in South Korea due to high power consumption, poor durability of equipment and facilities, and operational difficulties (Um and Kim, 1986)

Table 2.7 Examples of Faecal Sludge Treatment Systems (continued).

Country: Japan continued		
Type of Process	**Reference**	**Remarks**
Two-stage anaerobic digestion processes	Noike and Matsumoto (1986)	Gas production is almost completed in the first reactor; main role of second reactor is to separate supernatant from digested sludge by physical sedimentation.
Nitrified Liquor Recycling Process	Suzuki and Tohya (1986)	Over 100 plants; High removal rate for BOD and Nitrogen; some of the plants have been re-constructed from two stage anaerobic digesters
Low dilution: two-stage biological denitrification treatment + advanced treatment **(Figure 2.6)**	Misaki and Matsui (1996)	Low dilution, two-stage biological denitrification involving a 10-times dilution. Used from latter half the 1970s
Advanced treatment: **(Figure 2.7)** This is a combination of coagulation, sedimentation, ozone oxidation, sand filtration, and activated carbon absorption	Misaki and Matsui (1996)	Used from latter half the 1970s.
High-load denitrification treatment + advanced treatment **(Figure 2.8)**	Misaki and Matsui (1996)	System does not require additional water for dilution. Used from the early 1980s
High-load denitrification membrane separation treatment + advanced treatment **(Figure 2.9)**	Misaki and Matsui (1996)	System used ultrafiltration (UF) membranes and was put in operation in 1987. Over 50 constructed by 1994

Table 2.7 Examples of Faecal Sludge Treatment Systems (continued).

Country: Korea		
Type of Process	**Reference**	**Remarks**
Extended aeration process (similar to aerobic digestion process) • Pre-treatment (receiving, screening & holding) • Non-diluted aeration (15 days) • Supernatant treated by activated sludge process	Um and Kim (1986)	Rural[1] type process. The aerobic process is effective in achieving a good level of effluent quality and has simpler operation and maintenance
Anaerobic digestion process[2] • Pre-treatment (receiving, screening & holding) • Conventional anaerobic digestion (30 days) • Diluted 10 times and treated by oxidation ditch process	Um and Kim (1986)	Rural type process.
Unheated digestion process (non-diluted) • Pre-treatment (receiving, screening & holding) • anaerobic digestion (90 days) followed by oxidation pond process	Um and Kim (1986)	Rural type process
Anaerobic digestion process • Pre-treatment (receiving, screening & holding) • Primary treatment of anaerobic digestion • Secondary treatment: 20 times dilution with water and then treated by activated sludge process • Effluent is disinfected and discharged	Um and Kim (1986)	Urban[2] type process
Aerobic digestion process • Pre-treatment (receiving, screening & holding) • Primary treatment: Non-diluted aeration process • Secondary treatment: 20 times dilution with water and then treated by activated sludge process	Um and Kim (1986)	Urban type process. The aerobic process is effective in achieving a good level of effluent quality.

[1] Two main treatment systems are defined in the Design Manual for Korea - urban and rural types. In recent years, lagoons and RBC processes have also been constructed to treat supernatant as additional processes in rural type plants. Urban type plant include sludge treatment and deodorization processes

[2] Number of plants using anaerobic digestion process is continuously increasing, especially for the urban type; increase is attributed to the possibility of energy recycling and lower running costs

Table 2.7 Examples of Faecal Sludge Treatment Systems (continued).

Country: Korea		
Type of Process	**Reference**	**Remarks**
Two-stage activated sludge process • Pre-treatment (receiving, screening & holding) • Primary treatment: diluted aeration process (20 times dilution) • Secondary treatment: 20 times dilution with water and then treated by activated sludge process	Um and Kim (1986)	Urban type process. No new plants since 1981 and existing plants are being modified; situation seems to be the result of excessive power costs and technical problems in operation and maintenance.
Single Stage Sequencing Batch Reactor	Choi *et al.* (1997)	Converted from an existing 2 stage ASP nightsoil treatment plant for nutrient removal
Country: China		
Anaerobic digestion for sludge treatment; Trickling filter for supernatant; Drying beds used in dewatering the digested sludge, which is used as fertiliser in the countryside	Guo *et al.* (1991)	Plant had to be modified after commissioning because the primary investigation on the characteristics of nightsoil was insufficient.
Country: India		
Either burial in the ground alone or with town refuse	Satyanarayan *et al.* (1987)	Creates fly and odour nuisance along with contamination of ground water by percolation and leaching
Composting of nightsoil along with town refuse	Satyanarayan *et al.* (1987)	Practice not well designed and hence results in breeding of flies, odour nuisance and incidence of helminth infections
Anaerobic digestion	Mara (1976)	Digested sludge is used as fertiliser; applied in liquid form

Table 2.7 Examples of Faecal Sludge Treatment Systems (continued)

Type of Process	Reference	Remarks
Country: Thailand		
Extended aeration followed by ponds, drying beds for separated sludge	Strauss and Heinss (1995)	
Country: Indonesia		
Type of Process	**Reference**	**Remarks**
Extended aeration followed by unplanned dumping of sludge generated at disposal sites	Stoll and Parameswaran (1996)	Priority options have been identified for sludge management and these are: 1. Mono-incineration after on-site dewatering 2. Direct agricultural use after on-site dewatering 3. Composting after on-site dewatering for agricultural use or land reclamation
Country: Ghana		
Batch-operated settling/thickening tanks followed by ponds, separated solids are windrow-composted with sawdust	Strauss and Heinss (1995)	Ponds operate in anaerobic regime, inhibition of algal growth from ammonia toxicity
Country: Nigeria		
Collected nightsoil is screened, diluted with make-up water, macerated and treated in aerated lagoons	Mara (1976)	
Country: Tanzania		
Co-treatment at central sewerage treatment facilities using waste stabilisation ponds	Strauss and Heinss (1995) Mgana (1997)	Faecal sludge mostly from urban areas; pre-treated in anaerobic ponds

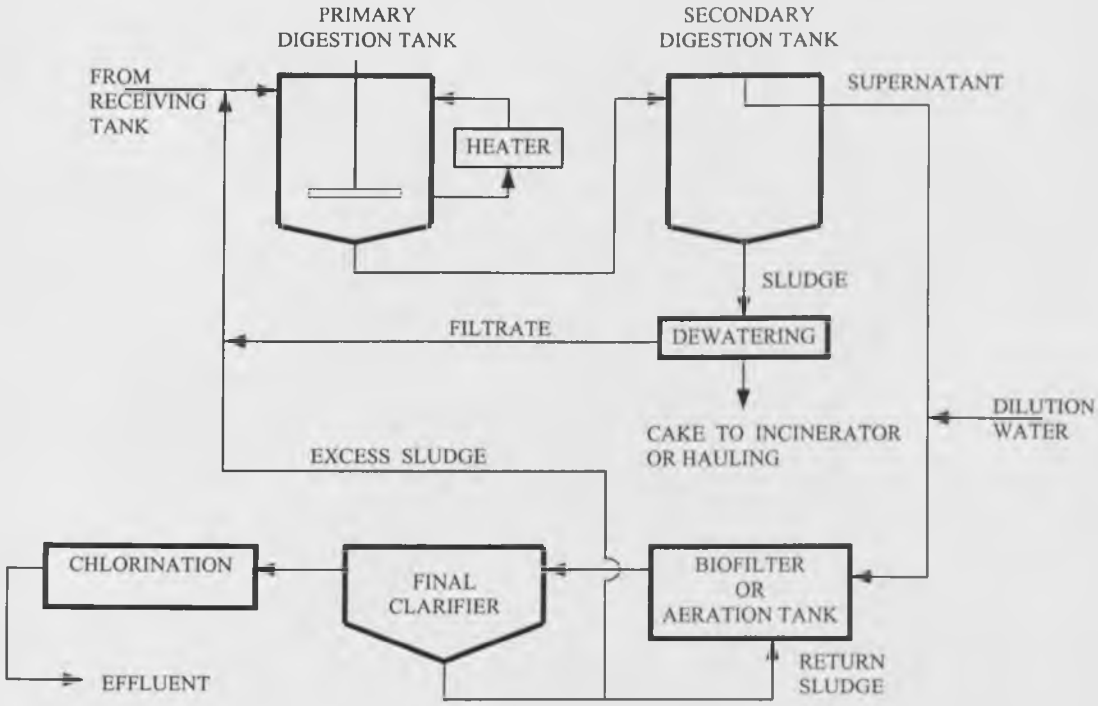


Figure 2-2 Nightsoil digestion process

Source: Pradt (1971)

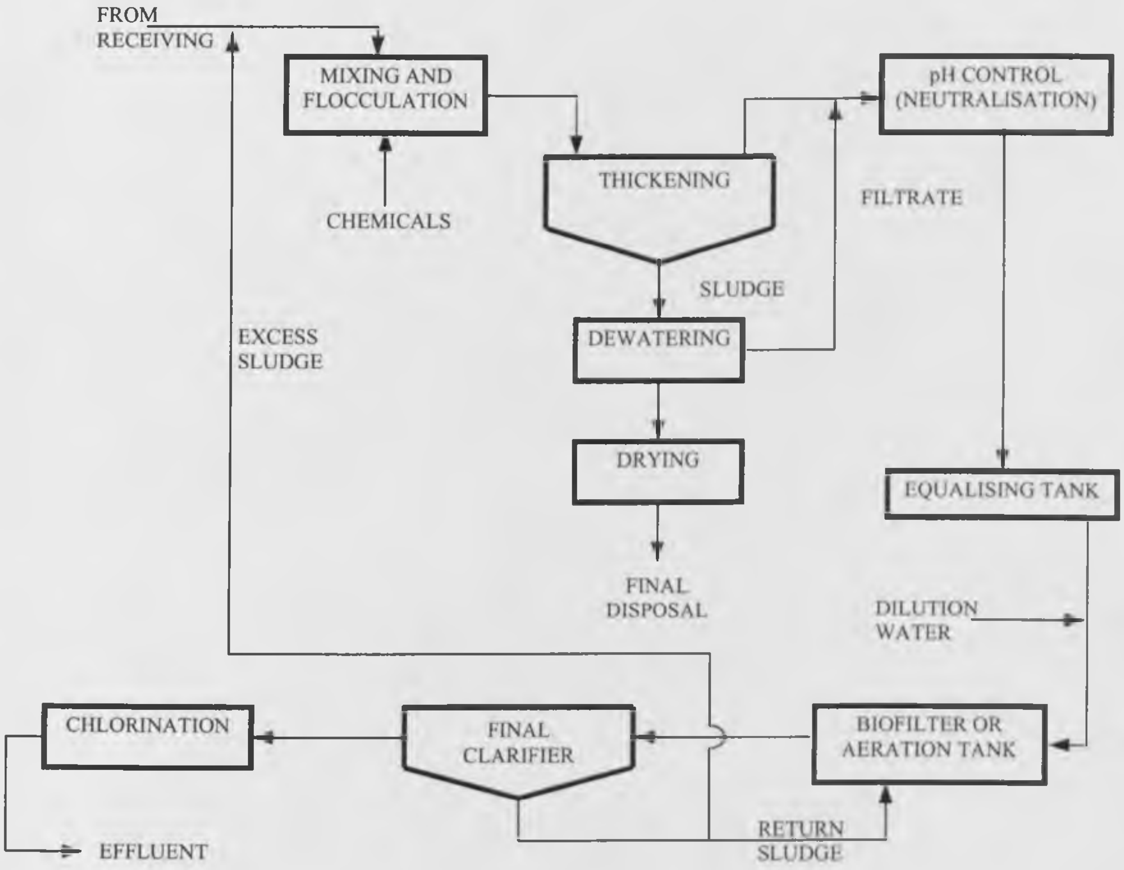


Figure 2-3 Nightsoil chemical treatment process

Source: Pradt (1971)

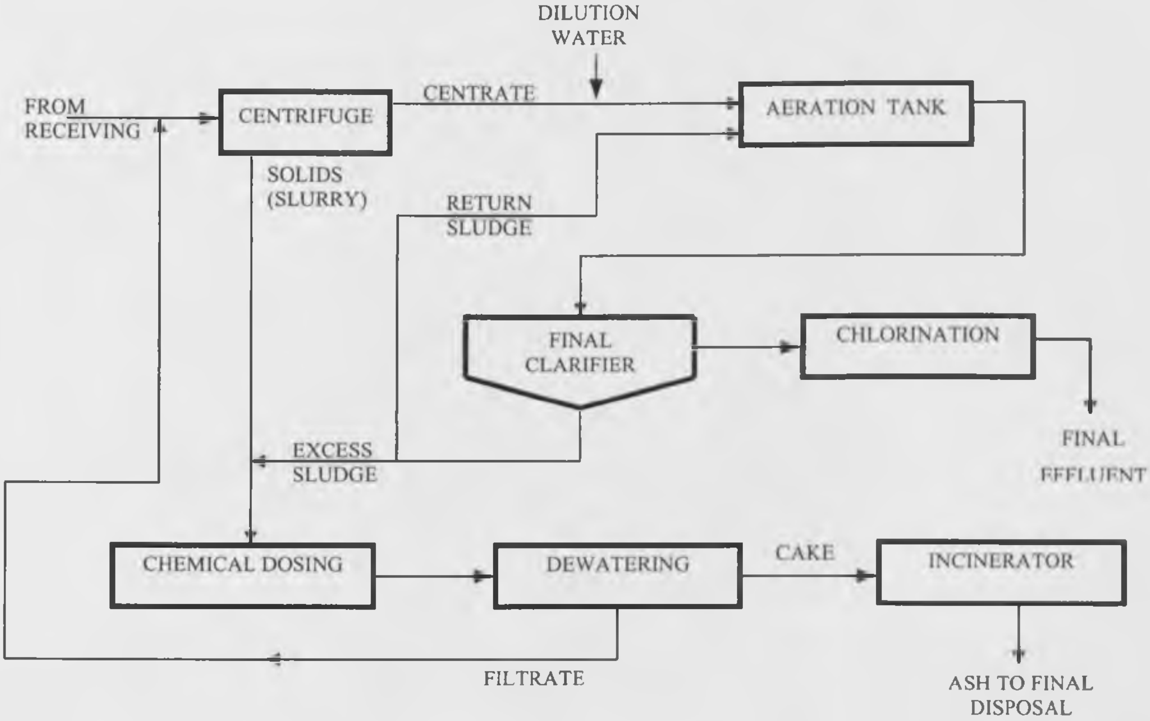


Figure 2-4 Nightsoil bio-oxidation process

Source: Pradt (1971)

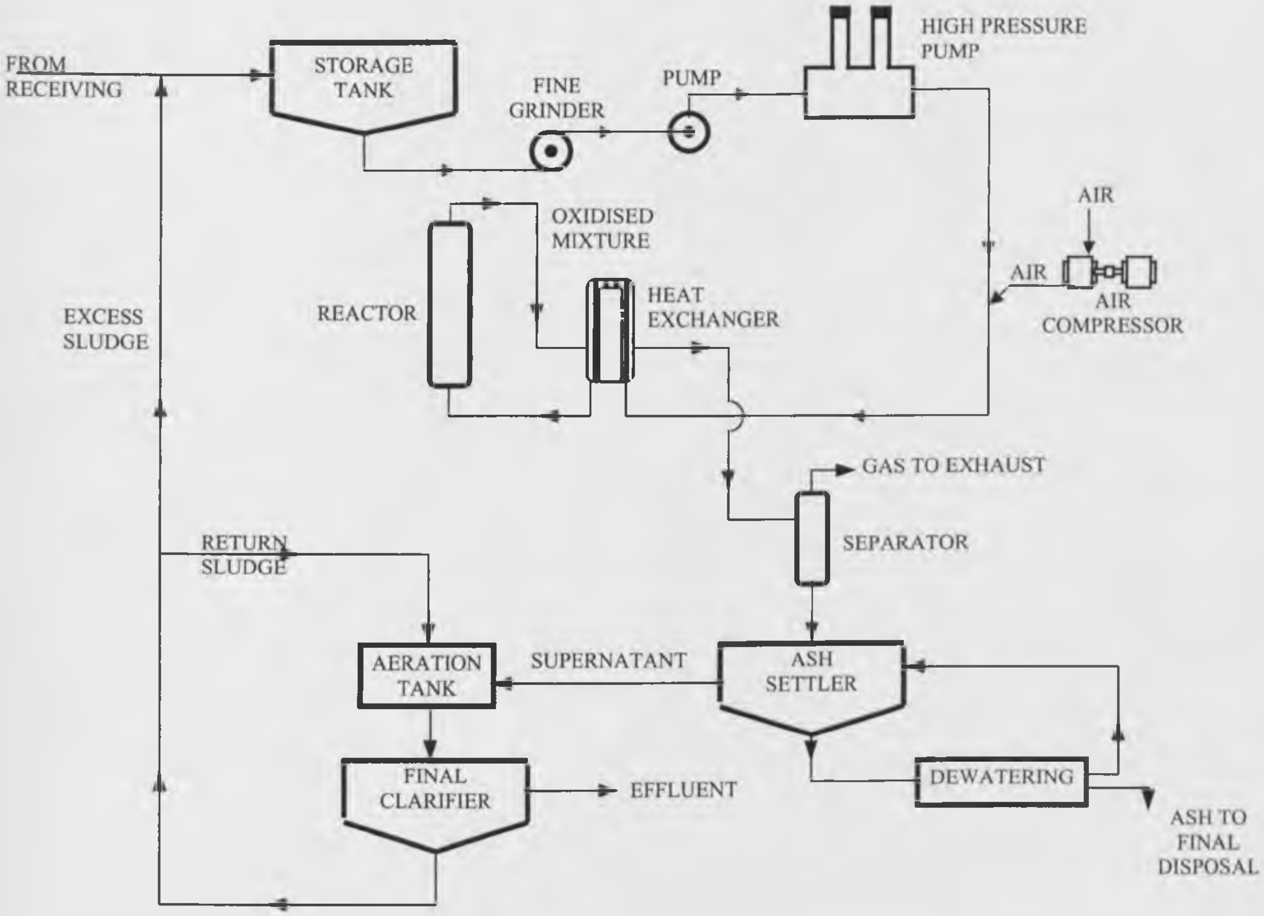


Figure 2-5 Nightsoil wet air oxidation process.

Source: Pradt (1971)

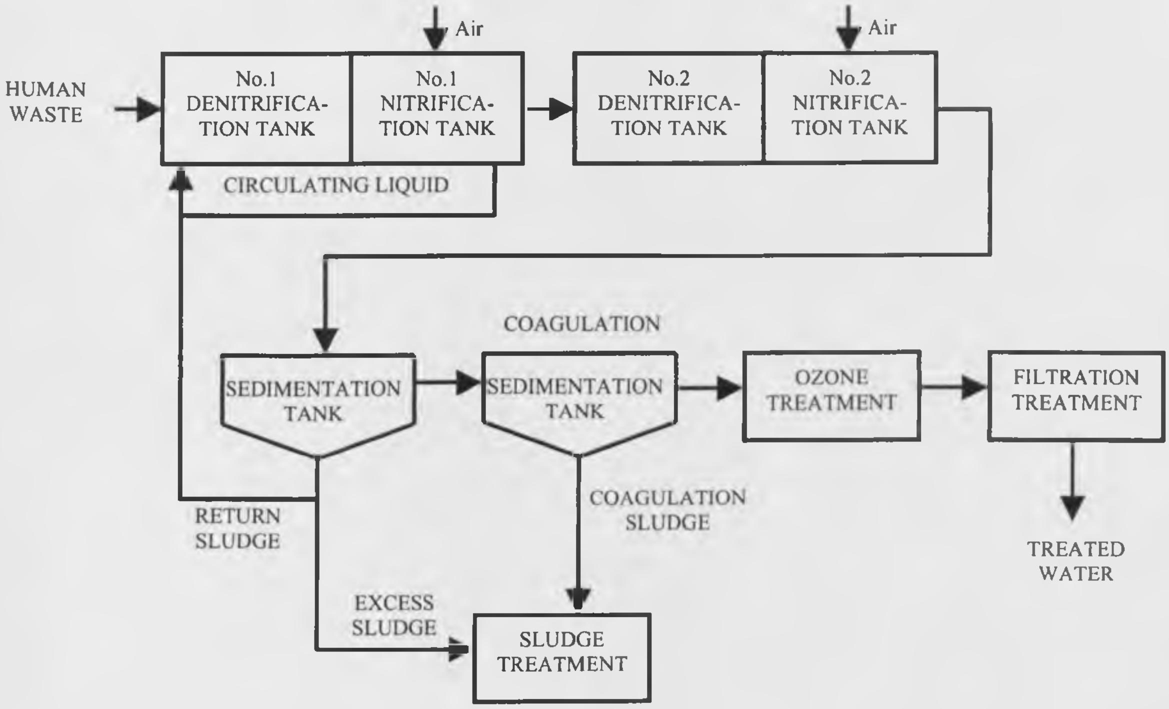


Figure 2-6 **Low dilution: two-stage biological denitrification treatment + advanced treatment**

Source: Misaki and Matsui (1996)

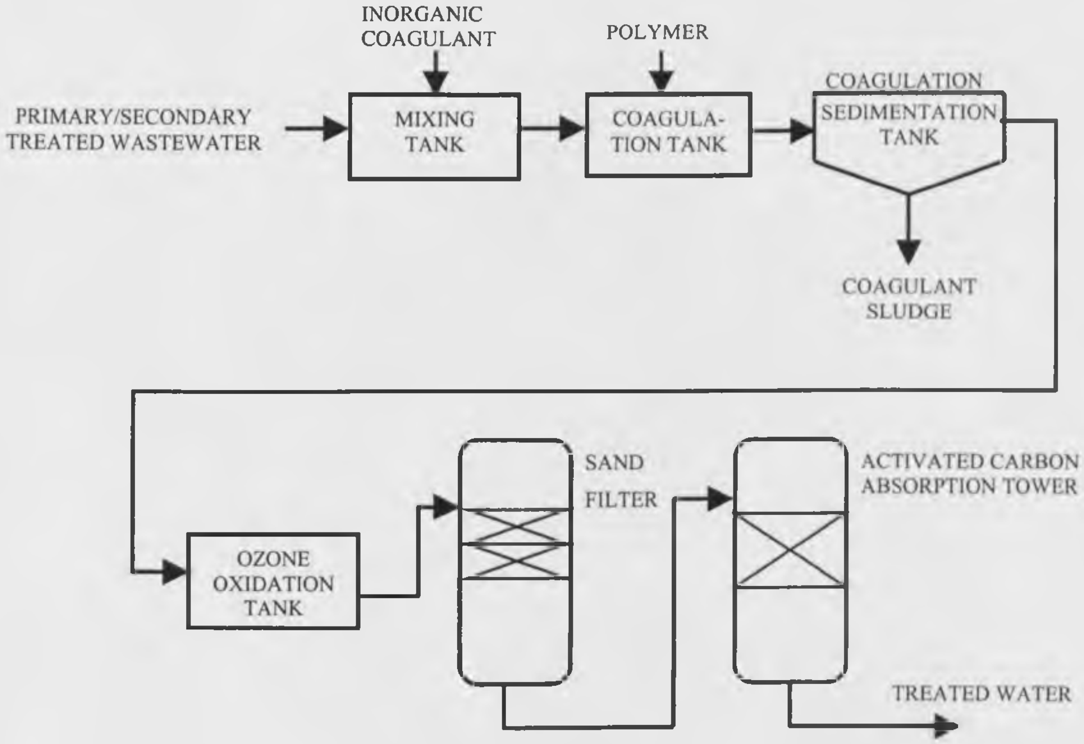


Figure 2-7 Advanced Treatment

Source: Misaki and Matsui (1996)

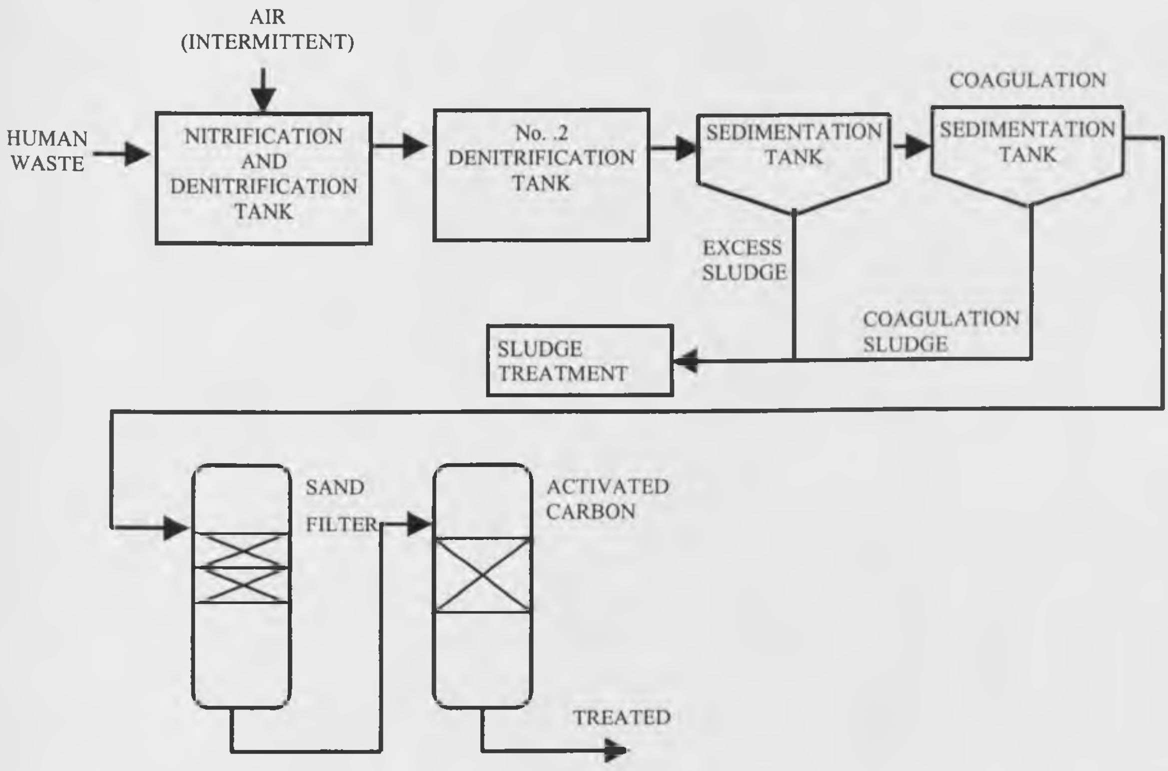


Figure 2-8 High-load denitrification treatment + advanced treatment

Source: Misaki and Matsui (1996)

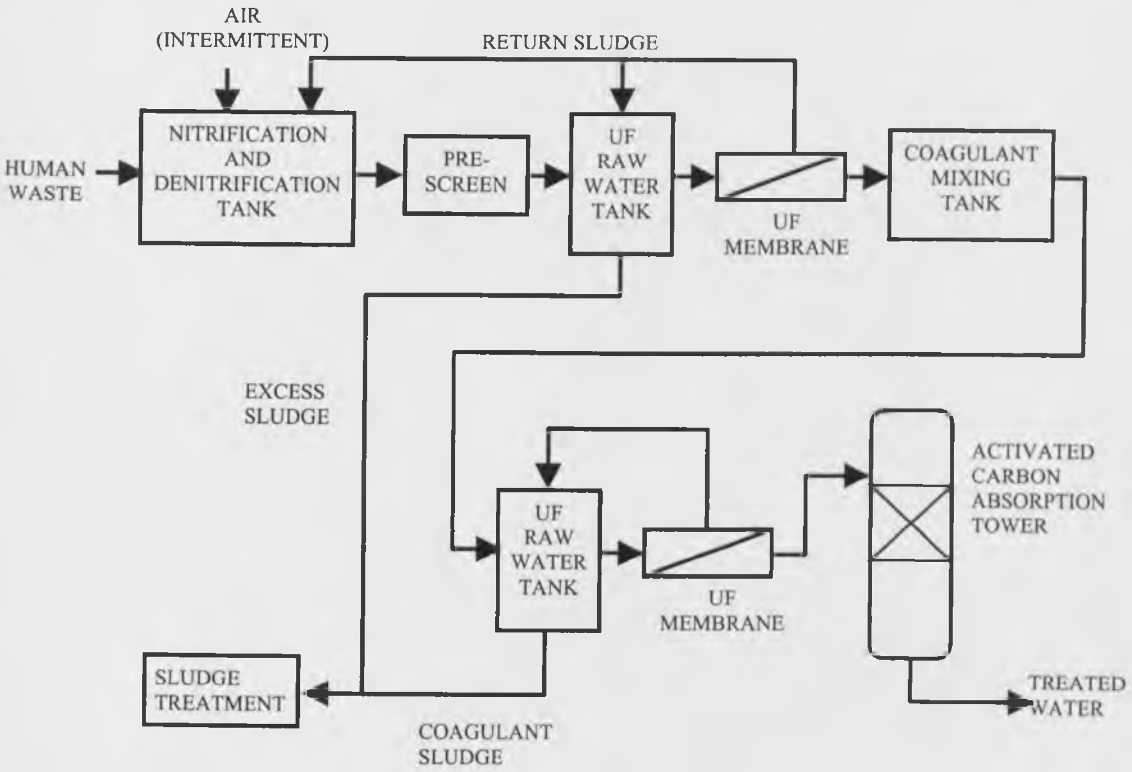


Figure 2-9 High-load denitrification membrane separation treatment + advanced treatment

Source: Misaki and Matsui (1996)

Despite this remarkable progress, the costs and sophistication of the technologies precludes their use in developing countries like Ghana, unless with further modifications to make them simpler in the application. Heinss *et al.* (1998) stress the fact that methods for treatment of faecal sludges in developing and newly industrialising countries should be relatively low-cost, i.e., low in capital and operating costs, and also compatible with the expertise available in the particular country. This cannot be said about the conventional anaerobic or aerobic digestion, and the nitrification and denitrification treatment coupled with advanced treatment systems as used in Japan and elsewhere.

Development of low-cost technology for the treatment of nightsoil and faecal sludge really lacks behind that of the conventional technologies. As evident from Table 2-7, the simple and low-cost technologies in use are waste stabilisation ponds or lagoons for the treatment of the supernatant after primary treatment. When lagoons are used solely they have their shortfalls as outlined by Heinss *et al.* (1998) which must be addressed to make them more effective. One major area that has to be addressed is the need for a pre-treatment system to reduce significantly the strength of the waste before using the pond system. As seen from Table 2-7, in most of the easier and simpler systems that can be adapted to suit conditions in developing countries, the reduction is achieved by using either conventional aerobic or anaerobic digestion followed by the addition of dilution water. In an integrated treatment system, the dilution water could be provided by recycling the final effluent from the treatment system.

In rural Korea, the reduction in the strength of the waste is achieved using unheated anaerobic digestion process for primary treatment of the undiluted nightsoil and lagoons or ponds to treat the supernatant from the digestion process (Um and Kim, 1986). Judging from the socio-economic and climatic conditions of Ghana and most developing countries, this approach may be the most favourably provided the unheated anaerobic digestion process could be adapted. The unheated anaerobic digestion system could fit well into the present faecal sludge treatment plants (FSTP) in operation in Accra and other cities/towns in Ghana. Figure 2-10 shows a treatment flowpath if the unheated anaerobic digestion process is incorporated into the current FSTPs.

Unheated anaerobic digestion should pose no problems in developing countries since ambient temperatures in tropical countries will be near the upper end of the optimal temperature range (16°C to 38°C) required by the mesophilic bacteria. Unheated anaerobic digestion studies carried out by Suchint (1967) [cited by Pescod (1971)] on sludge settled from fresh nightsoil indicated that anaerobic digestion was a suitable method for treating nightsoil sludge. Lagoons or ponds systems are well developed and are already in use in Ghana and other developing countries. They could be used in the treatment of the supernatant resulting from the unheated digestion process, after separation of the digested sludge.

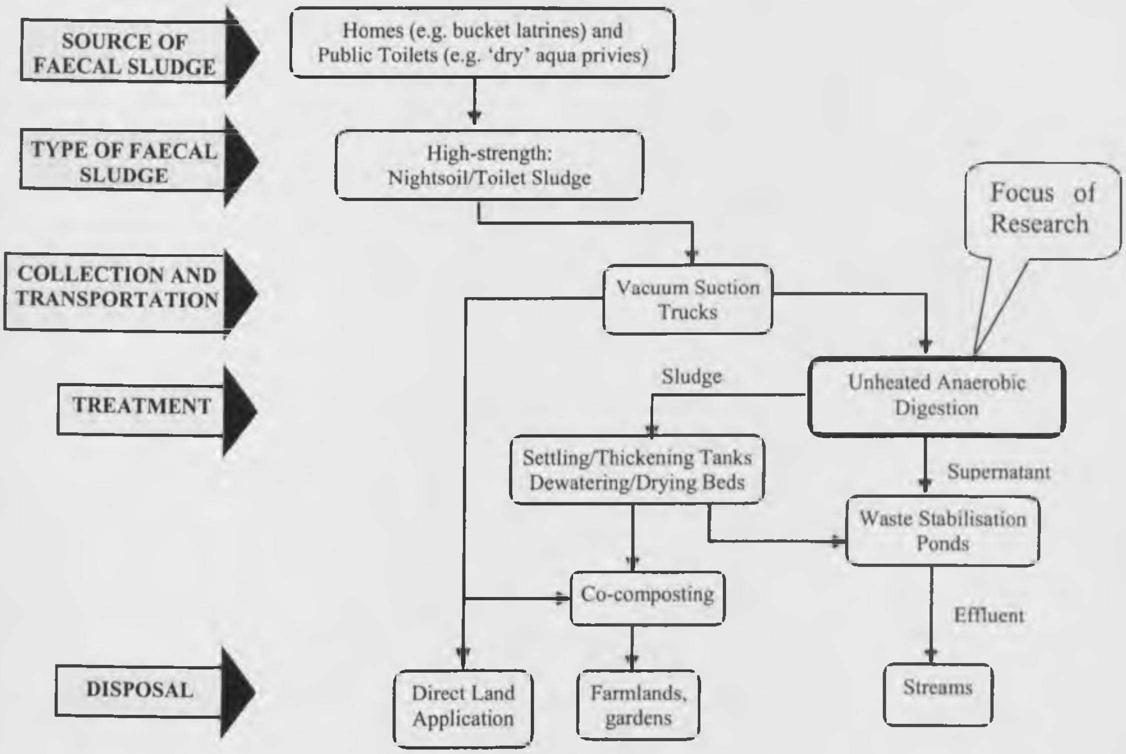


Figure 2-10 **Treatment Flowpath for high-strength faecal sludge in Ghana incorporating unheated anaerobic digestion process**

2.5 Anaerobic Digestion Process

2.5.1 Introduction

In today's world, the role played by treatment of wastes in the abatement of environmental pollution is well recognised by all countries, both developed and developing. In the field of human wastes treatment, conventional aerobic treatment systems have been well developed and used in the industrialised countries while their application in developing countries have been dismal. These aerobic treatment systems are energy intensive, use a lot of mechanical equipment and require trained and skilled personnel for smooth operation. With the steep rise in the costs of energy from the 1970's it has become imperative to develop alternate processes with lower energy consumption compared to energy required for aerobic processes. This is especially true for developing countries if waste treatment systems are to play any meaningful role in the fight against environmental pollution resulting from human wastes. The result of years of intensified research efforts is the development of alternate treatment systems that utilise anaerobic digestion instead of aerobic metabolism for the removal of organic material from human wastes. Anaerobic digestion has become increasing popular because of the following advantages relative to aerobic methods (Pretorius, 1983; Mudrack and Kunst, 1986; Sterritt and Lester, 1988; Malina, 1992; van Haandel *et al.*, 1996):

- anaerobic digestion process requires considerably less energy and dispenses with the need for mechanical aeration, an essential requirement for aerobic processes;
- useful energy may be recovered from methane, one of the end products;
- relatively less sludge is produced and hence a resulting lower costs of disposal of the organic residues;
- well-designed anaerobic processes have far greater treatment capacity than aerobic processes and therefore require a much smaller reactor volume; and
- many substances which are not degradable under aerobic conditions can be decomposed anaerobically.

2.5.2 Basic Concepts

Basically, anaerobic digestion is a bacterial fermentation process by which organic material is broken down in the absence of dissolved oxygen to produce stable end-products, mainly methane and carbon dioxide (McCarty, 1982; van Haandel *et al.*, 1996). Haandel and Lettinga (1994) referred to the process as the ultimate fermentative process because it is characterised by the production of methane, which is the most reduced organic compound. For an organic matter $C_xH_yO_z$, the process of anaerobic digestion can be written as:

$$C_xH_yO_z + \tfrac{1}{4}(4x - y - 2z)H_2O \rightarrow \tfrac{1}{8}(4x - y - 2z)CO_2 + \tfrac{1}{8}(4x + y - 2z)\ CH_4 \qquad (2.1)$$

During the process of anaerobic digestion, only a minor fraction of the chemical energy in the organic matter is released, the major part remaining as chemical energy in the methane produced. McInerney *et al.* (1979) calculate that about 90% of the available energy in the organic matter is retained in the methane produced as illustrated in the following example using acetic acid (van Haandel and Lettinga, 1994): during the oxidation of acetic acid with oxygen, the free energy released is equal to 207 kcal mol^{-1} in the reaction

$$C_2H_4O_2 + 2O_2 \rightarrow 2CO_2 + 2H_2O + 207\,\text{kcal} \qquad (2.2)$$

Similarly, during the oxidation of methane with oxygen, the free energy released is equal to 191 kcal mol^{-1} in the reaction

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O + 191\,\text{kcal} \qquad (2.3)$$

Acetic acid is digested to methane and carbon dioxide according to the following reaction

$$C_2H_4O_2 \rightarrow CH_4 + CO_2 + E_{dig} \qquad (2.4)$$

where E_{dig} is the free energy released.

The combination of equations 2.3 and 2.4 results in equation 2.2. Hence the free energy released from equations 2.3 and 2.4 must be equal to that released in equation 2.2.

$E_{dig} + 191 = 207$, hence $E_{dig} = 16$ kcal mol^{-1}

This shows that free energy released during the anaerobic digestion of acetic acid is only a fraction of 16/207 = 8 percent of the free energy released during the aerobic oxidation of the same compound, the rest of the energy (92%) being retained in the methane. Furthermore, with the relatively large release of chemical energy very little energy is locked up in the new microbial cells produced during the process, and thus the relative amount of new microbial cells formed as surplus sludge is also small (Mosey, 1981).

The fact that dissolved oxygen is not needed for the process, that methane as a combustible gas has commercial value, and that the biomass production is relatively small makes the anaerobic digestion process ideal for the stabilisation of organic sludges, the treatment of concentrated organic industrial wastes and the production of methane gas from agricultural and domestic wastes (Pretorius, 1983).

Despite the advantages highlighted previously, the general acceptance and applicability of anaerobic digestion is only now beginning to rise within the last decade. The lack of general acceptance and applicability in the past (over a decade and half ago), have been attributed the following disadvantages (Pretorius, 1983; Mudrack and Kunst, 1986; Sterritt and Lester, 1988; Malina, 1992) which may no longer hold due to the advances in the knowledge of anaerobic digestion process and treatment technology:

- the lower rates of treatment attainable in anaerobic systems result in the treated effluent being fairly heavily polluted, and hence requiring further treatment before it can be discharged into receiving environment;
- bacteriologically speaking it was a very complex system which was not fully understood making it difficult to rectify problems cropping up during the process;

- the growth rate of certain members of the 'key' physiological groups responsible for the production of methane is very low at ambient temperatures so that the process becomes economically attractive only for wastes with temperatures at 25°C or above; and
- long hydraulic detention times are required to develop and maintain a population of methane-producing bacteria.

Accepting that anaerobic digestion generally cannot provide a complete treatment, Lettinga (1995) dismisses the previously mentioned drawbacks based on the present state of knowledge of the anaerobic digestion process:

- with regard to the bacteriological complexity of the anaerobic systems, significantly more is known about the system today and gradually a better insight is being gained in the countermeasures that can be taken if problems arise in the operation of the system;
- growth rates for the 'key' physiological groups responsible for the production of methane will be optimal when the anaerobic digestion process is applied in tropical climates where the ambient temperatures are within the mesophilic range.
- the presumed low stability could be attributed to a lack of knowledge about the basic principles of the anaerobic treatment process. As a matter of fact, the anaerobic digestion process is highly stable provided the system is operated in the proper way;
- much more is understood of the growth conditions of anaerobic organisms, and gradually large quantities of highly active anaerobic sludge from existing full-scale installations are becoming available, so that start-up of new systems can be made within a few weeks, sometimes even a few days.

2.5.3 Conversion Processes in Anaerobic Digestion

The process of anaerobic digestion of organic matter involves a number of transformations of the macromolecules present by several micro-organisms. Six distinct conversion processes have been identified in the degradation of particulate organic material to methane by Gujer and Zehnder (1983). These six processes are:

1. Hydrolysis of particulate organic material (biopolymers).
    a. Hydrolysis of proteins
    b. Hydrolysis of carbohydrates
    c. Hydrolysis of lipids
2. Fermentation of amino acids and sugars.
3. Anaerobic oxidation of long chain fatty acids and alcohols.
4. Anaerobic oxidation of intermediary products such as volatile acids (with the exception of acetate).
5. Conversion of acetate to methane
6. Conversion of hydrogen to methane.

Gujer and Zehnder (1983), Sterritt and Lester (1988) and van Haandel and Lettinga (1994) place these six distinct reactions in the conversion processes into four phases which are illustrated in Figure 2.11.

2.5.3.1 Hydrolysis

In general bacteria are not able to take up particulate organic material since it first has to be broken down into soluble polymers or monomers. Thus hydrolysis or liquefaction is the first step required for microbial utilisation of the complex biopolymers. In the hydrolysis process the particulate organic matter is converted into dissolved compounds of a lower molecular weight. Proteins are degraded to amino acids, carbohydrates are transformed into soluble sugars (mono- and disaccharides) and lipids are converted to long chain fatty acids and glycerine. Exo-enzymes, excreted by fermentative bacteria, are required for this process (van Haandel and Lettinga (1994) and Sterritt and Lester (1988)).

2.5.3.2 Acidogenesis

Once the particulate organic material has been converted into soluble compounds as a result of hydrolysis, the fermentative bacteria takes up the dissolved compounds. In this acid-forming phase, various short-chain organic acids (e.g. butyric acid, propionic acid, acetic acid) are formed, together with alcohols, and mineral compounds such as carbon dioxide, hydrogen, ammonia and hydrogen sulphide gas.

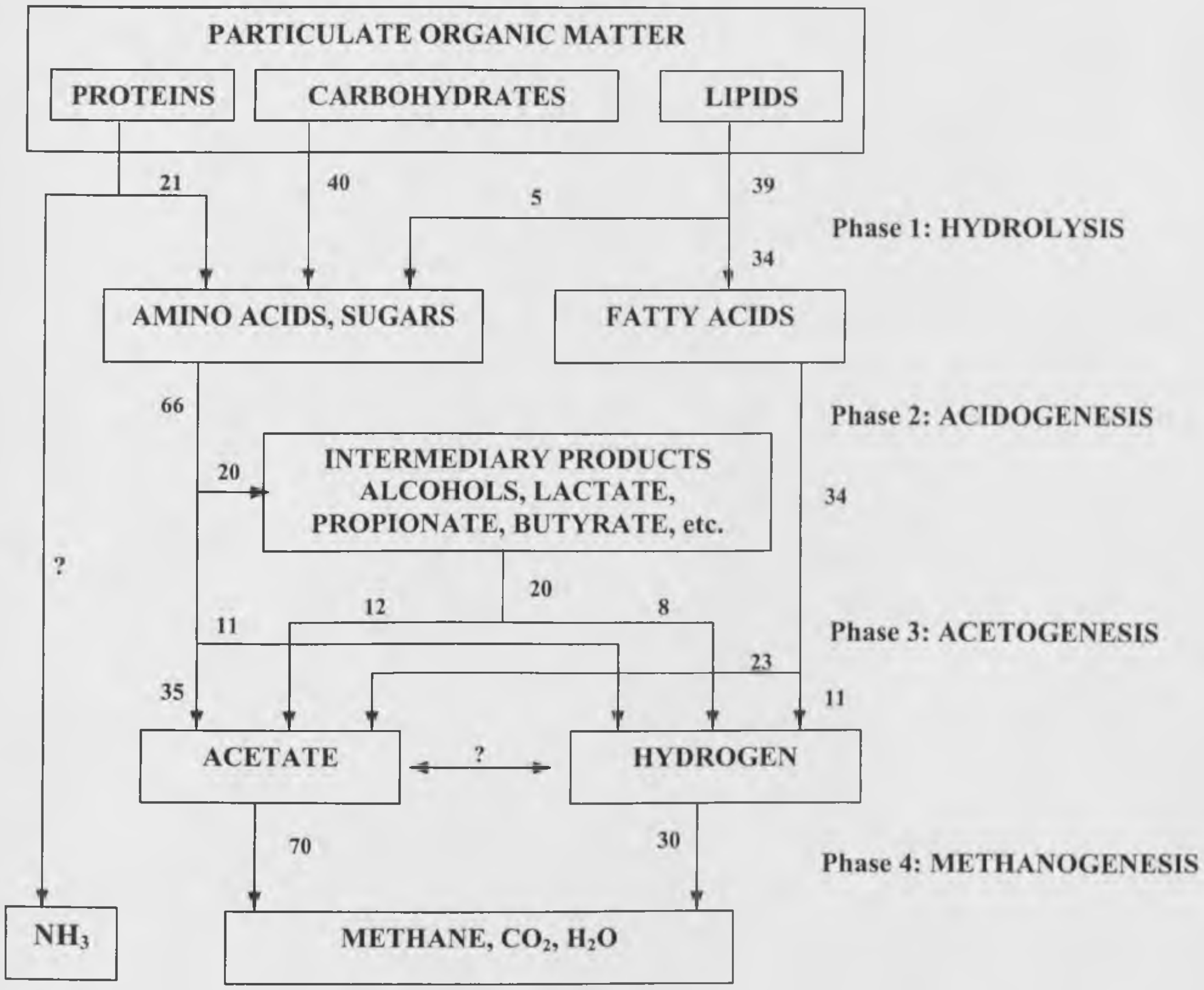


Figure 2-11 Reaction sequence for the anaerobic digestion of particulate organic material

(Numbers refer to percentages expressed as COD)

Source: Adapted from Gujer and Zehnder (1983), Sterritt and Lester (1988) and van Haandel and Lettinga (1994).

This phase is a carried out by a diverse group of bacteria, most of which are obligate anaerobes. However, Mudrack and Kunst (1986) and van Haandel and Lettinga (1994) state that some facultative bacteria are also involved and metabolise organic matter via the oxidative pathway. In this way, the little dissolved oxygen that might otherwise become toxic to the obligate anaerobic organisms is utilised.

2.5.3.3 Acetogenesis

Among the products formed in the acidogenesis stage, only acetic acid (acetate), hydrogen and carbon dioxide can be converted by the methane bacteria directly into methane. Hence in this acetogenic phase, the other products of acidogenesis are transformed into acetate, hydrogen and carbon dioxide, the final products for methane production. As indicated in Figure 2.11, a larger percentage (approximately 70%) of the COD originally present is converted into methane via the acetate route. The production of acetate is generally accompanied by the formation of both hydrogen and carbon dioxide, with hydrogen being more than the carbon dioxide. Collectively, the related organisms responsible for the production of acetate and hydrogen in this phase are knows as the obligatory hydrogen-producing acetogenic (OHPA) bacteria.

2.5.3.4 Methanogenesis

In this last and final phase, methane is produced from acetate or from the reduction of carbon dioxide by hydrogen using acetotrophic (Equation 2.5) and hydrogenotrophic (Equation 2.6)bacteria respectively.

Acetotrophic methanogenesis:

$$CH_3COOH \rightarrow CH_4 + CO_2 \qquad (2.5)$$

Hydrogenotrophic methanogenesis:

$$4H_2 + CO_2 \rightarrow CH_4 + 2HO_2 \qquad (2.6)$$

Methanogenesis is the final stage of the anaerobic digestion process and is of vital importance for the whole process. In particular, Henze and Harremoes (1983) state that the bacteria responsible for hydrogenotrophic methanogenesis grow faster than the bacteria which utilise acetate, and so the acetotrophic methanogenesis is the rate limiting step with respect to the transformation of particulate organic material to methane.

2.5.4 Organic nitrogen transformations

Total nitrogen is comprised of organic nitrogen, ammonia nitrogen, nitrite and nitrate. In faecal sludge the nitrogen is present principally as organic nitrogen and ammonia nitrogen and with only small amounts of nitrite and nitrate. In aqueous solution, ammonia nitrogen exits as either the ammonium ion or ammonia depending on the pH of the solution (Figure 2.15) in accordance with the following equilibrium reaction:

$$NH_3 + H_2O \leftrightarrow NH_4^+ + OH^- \tag{2.7}$$

Organic nitrogen is found in complex nitrogenous compounds such as proteins and urea. As shown in Figure 2.11, during the first phase of anaerobic digestion, the hydrolysis or liquefaction process degrades the proteins to amino acids and the subsequent deamination of these amino acids results in the release of ammonia nitrogen. The conversion of soluble organic nitrogen into ammonia nitrogen as bacteria consume soluble organic matter containing nitrogen is referred to as ammonification. Leslie Grady Jr. (1999) states that it is difficult to measure the true rate of ammonification because the ammonia nitrogen is being consumed by the bacteria as they grow, and the only measurable event is the net accumulation or loss of ammonia in the medium. There is accumulation of ammonia in the medium if the amount of nitrogen available exceeds the need. On the other hand there is a decrease in the concentration of ammonia in the medium if the amount of nitrogen is available is less than the need.

2.5.5 Stability of the Conversion Processes

Since the conversion processes in an anaerobic digestion depend on the interactions of several bacteria, maintaining the ecological balance and favourable environmental conditions are vital in preventing failure of the process. Important environmental factors of primary importance to the anaerobic digestion processes are temperature, pH, the presence of essential nutrients, and toxic or inhibitory substances. In nightsoil and toilet sludges, nutrients (both macronutrients -nitrogen and phosphorus, and micronutrients) are abundantly available and is therefore not considered.

2.5.5.1 Temperature

Anaerobic digestion processes, like other biological processes, strongly depend on temperature. The influence of temperature on the rate and extent of anaerobic digestion has been the subject of many investigations. With respect to the conversion rate of digestion processes, van Haandel and Lettinga (1994) state that there are maxima between 35 and 40°C for mesophilic range and at about 55°C for the thermophilic range. For an unheated anaerobic treatment, only the mesophilic digestion range is considered. Henze and Harremoes (1983) evaluated available data and Figure 2.12 shows a graphical representation of their analysis and of some recent data as reported by van Haandel and Lettinga (1994).

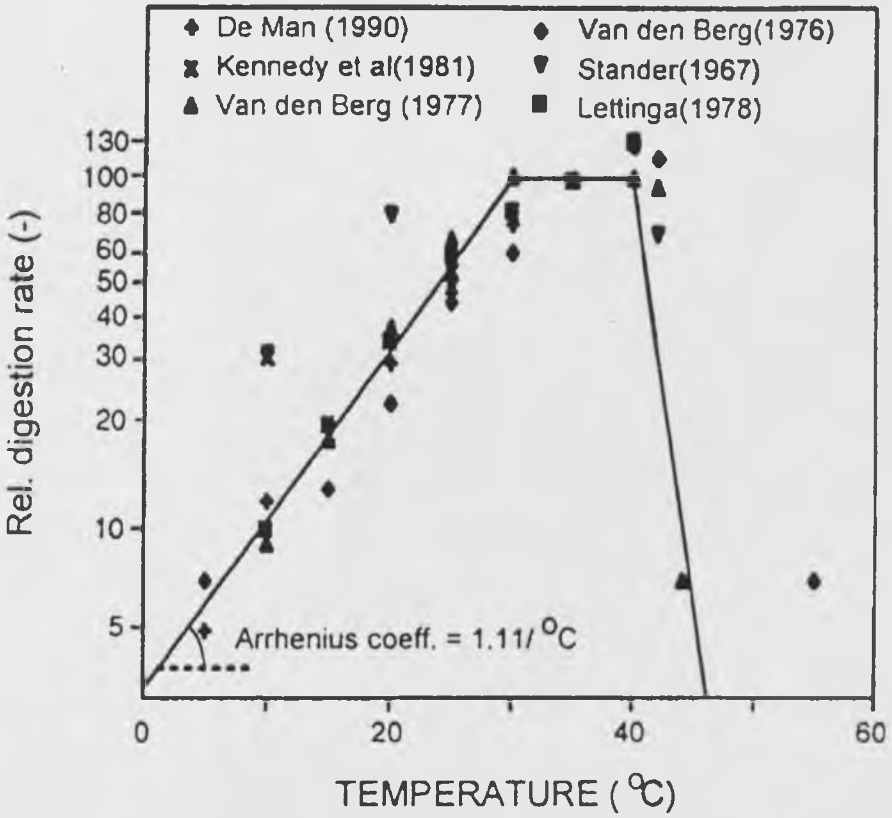


Figure 2-12 Influence of temperature on the rate of anaerobic digestion in the mesophilic range.

Source: van Haandel and Lettinga (1994)

From Figure 2.12, van Haandel and Lettinga (1994) drew the following conclusions:

1. that the optimum range is between 30 and 40°C, and
2. that for temperatures below the optimum range the digestion rate decreases by about 11 per cent for each °C temperature decrease, or according to the Arrhenius expression:

$$r_t = r_{30}(1.11)^{(t-30)} \tag{2.8}$$

where t = temperature in °C and r_t, r_{30} = digestion rate at temperature t and 30°C, respectively.

From their evaluation of available data, Henze and Harremoes (1983) reported that the anaerobic digestion processes can operate in the temperature range of 10-45°C without major changes to microbial ecosystem. For the various conversion processes described, Mudrack and Kunst (1986) state that the optimal conversion rates for the acid-forming bacteria (phases 1 to 3 in Figure 2.11) and the methanogenic organisms occur respectively at 30°C and 35-37°C. For temperatures below the optimum range, and in particular below 20°C, the conversion rate of lipids becomes very slow and hence the hydrolysis rate can be limiting for the overall rate of anaerobic digestion (van Haandel and Lettinga, 1994).

In addition to the influence of temperature on the rate of the anaerobic digestion process, the extent of the digestion is also affected as found by O'Rourke (1968) [cited by van Haandel and Lettinga (1994)] and Van der Last (1991). Figure 2.13 shows that for the same incubation time, the organic material removal efficiency decreases as the temperature decreases. The decrease in the fraction of organic matter degraded is attributed to the low rate of hydrolysis at low temperatures.

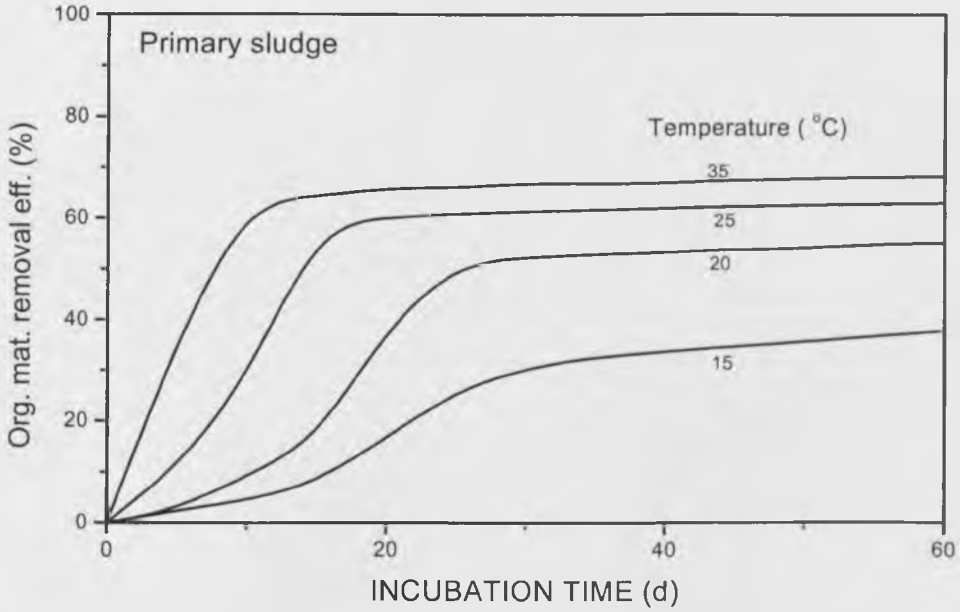


Figure 2-13 Influence of temperature on the extent and rate of anaerobic digestion of primary sludge.

Source: van Haandel and Lettinga (1994)

2.5.5.2 pH

Anaerobic digestion processes are dependent on pH. Hence it is very important that the value and stability of the optimal pH in an anaerobic rector are maintained. Most anaerobic conversion processes operate best near neutral pH. The acid forming bacteria, acidogens, (responsible for phases 1 to 3 in Figure 2.11) have an optimum pH between 5 and 6 while the methane bacteria, methanogens, have pH-optimum in the range 6 and 8 (Zehnder *et al.*, 1981) for uninhibited methane formation. The methane forming bacteria are very sensitive to pH values outside the optimal range whereas the acid forming bacteria are significantly less sensitive to low or high pH values. The optimal pH conducive for the growth of bacterial populations in an anaerobic digester is in the range pH 6.4 to 7.6 beyond which a state of inhibition may occur resulting from the toxic effects of the hydrogen ions (Anderson and Yang, 1992). Specifically, at pH values lower than 6.3 or higher than 7.8 the rate of methanogenesis decreases (van Haandel and Lettinga, 1994). Because the formation of methane is the rate limiting step it is important to maintain the pH in the neutral range. However, in an anaerobic reactor, deviations from these optima do occur.

These deviations, if not introduced with the influent, are usually the result of excess production and accumulation of acidic conversion products such as organic fatty acids (acidogens grow considerably faster than methanogens: Anderson and Yang, 1992), or basic conversion products such as ammonia.

2.5.5.3 Toxic Substances

Anaerobic digestion processes, like all other biological processes, can be affected by the presence of toxic substances. The toxicity or inhibition of the processes can be due to either introduction of the toxic substances with the influent or consequenced by the generation of intermediary products such as the volatile fatty acids. With respect to nightsoil and toilet sludge, the potentially toxic compounds that might be present apart from the hydrogen ion concentration are oxygen, sulphide (van Haandel and Lettinga, 1994) and ammonia (Henze and Harremoes, 1983).

Oxygen Toxicity

Some oxygen may be introduced in the influent, but this is used for oxidative metabolism in the acidogenesis process by facultative bacteria. Thus, practically, no dissolved oxygen will be present in the anaerobic reactor unless air is entrained together with the influent. In such circumstances, its introduction will be of no consequence for the performance of the reactor since it will be used for oxidative metabolism.

Sulphide Toxicity

Under anaerobic conditions, sulphite and sulphate are converted to sulphide, which has been implicated in exerting toxic effects on methanogenesis (Henze and Harremoes, 1983; Pohland, 1992; van Haandel and Lettinga, 1994). Thus of particular interest in the anaerobic digestion processes is the non-toxic sulphite and sulphate.

Sulphate is reduced to sulphide by sulphate-reducing bacteria (equation 2.9)

$$C_xH_yO_z + \tfrac{1}{8}(4x + y - 2z)SO_4^{2-} \rightarrow xCO_2 + \tfrac{1}{8}(4x + y - 2z)S^{2-} + \tfrac{1}{2}yH_2O \tag{2.9}$$

The toxicity of sulphide, which is normally present in solution as a weak acid, is closely related to free hydrogen sulphide concentration, which is pH dependent. At low pH (<6.5) the toxicity increases (Henze and Harremoes, 1983) since the percentage of free hydrogen sulphide is higher (Figure 2.14). Free hydrogen sulphide will inhibit at concentrations of approximately 100 mg S/l (Henze and Harremoes, 1983). Sulphide toxicity has been observed at concentrations ranging from 200 to 1500 mg/l (Stronach et al., 1986). However, according to results of Rinzema (1989) the sulphide concentration to be expected in anaerobic sewage treatment systems (up to 50 mg/l) is far lower than the minimum concentration of noticeable toxicity. Therefore sulphide toxicity may not be a problem in anaerobic sewage treatment systems. At higher pH (>8) and in the presence of other cations, especially iron, ferrous sulphide is precipitated and this reduces the toxicity. In general, Henze and Harremoes (1983) state that sulphate-sulphur concentrations in the influent below 300 - 600 mg/l should be regarded as harmless, unless the pH and precipitation change the picture radically.

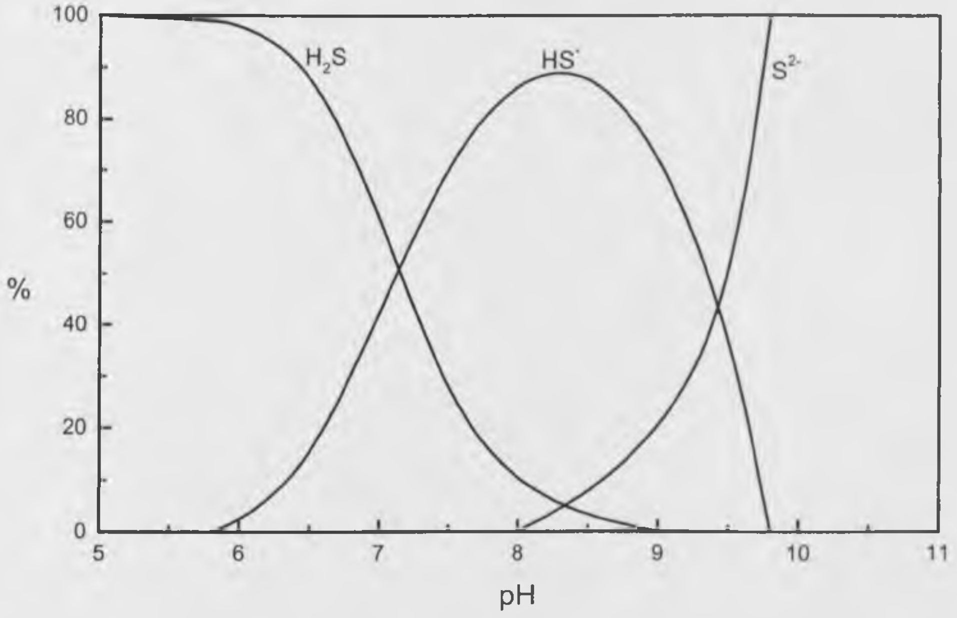


Figure 2-14 Effect of pH on hydrogen sulphide-sulphide equilibrium (10^{-3} molar solution, 32 mg H_2S/l)

Source: Sawyer and McCarty (1978)

Ammonia Toxicity

Although ammonia nitrogen is a necessary nutrient for the growth of bacteria involved in the anaerobic process (McCarty, 1964; Mah et al., 1977; Angelidaki and Ahring, 1994), it becomes inhibitory to methanogenesis (Koster and Lettinga, 1984; McCarty, 1964) when the concentration exceeds a certain limit. This toxicity has been suggested to be due to free molecular ammonia (NH_3) (Sawyer and McCarty, 1978; Henze and Harremoes, 1983; Sprott et al., 1984; Zeeman et al., 1985; Koster and Koomen, 1988). Molecular ammonia exists in equilibrium with ammonium, the relationship being dependent on pH and temperature (Figure 2.15), the free ammonia concentration increasing with increasing pH and temperature. Methane fermentation of high ammonia containing wastes are more easily inhibited at thermophilic temperatures than at mesophilic temperatures (Braun et al., 1981; Parkin and Miller, 1983; Angelidaki and Ahring, 1994). Hashimoto et al. (1981) and Angelidaki and Ahring (1994) showed that temperature had no significant effect on methane fermentation for temperatures in the mesophilic range. For unadapted methanogenic cultures, free ammonia inhibition has been observed to commence at concentrations of 100-200 mg N/l (Braun et al., 1981; Henze and Harremoes, 1983; De Baere et al., 1984) and total ammonia + ammonium inhibition at concentrations of 1500-2500 mg N/l (Van Velsen, 1979; Hashimoto, 1986). However, by adaptation of the methane fermentation process to ammonia, tolerance to 4000 mg N/l of total ammonia + ammonium (Hashimoto, 1986; Angelidaki and Ahring, 1993) and tolerance of up to 700 mg N/l of free ammonia (Angelidaki and Ahring, 1993) has been demonstrated if the pH of the reaction medium is low. As seen from Figure 2.15, at pH values close to the neutral range, the percentage of molecular ammonia is almost zero.

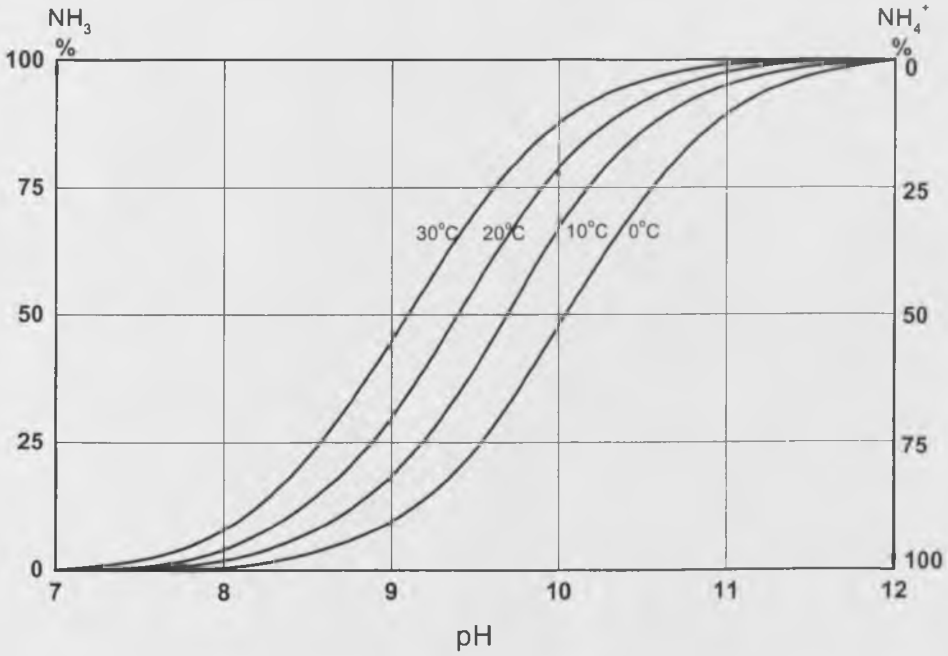


Figure 2-15 **NH_4-N and NH_3-N as a function of pH and temperature**
Source: Heinss *et al.*, 1998

2.5.6 Applicability of Anaerobic Digestion Process for the Treatment of Nightsoil and Toilet Sludges

Although literature on the direct use of anaerobic digestions processes for the treatment of nightsoil and toilet sludges is sparse, published work indicates that it is feasible to use anaerobic digestion process for the treatment of nightsoil and toilet sludge. The main task is getting the right environmental conditions and applying the treatment process suited to the socio-economic needs of a developing country like Ghana.

As described in the previous sections, anaerobic digestion processes can proceed under optimal conditions for all the bacterial groups involved if the following environmental conditions can be realised:

- For *temperature*, the most common range for conversion rate of the anaerobic digestion processes is reported to be 30-40°C, with the acid-forming bacteria and methanogenic organisms having optimal conversion rates at 30 °C and 35-37 °C respectively.

- For *pH*, the more sensitive methanogenic organisms have an optimal conversion rate in the range 6.3 - 7.8, with values closer to the lower end of the range being favourably also to the acid forming bacteria.
- With regards to toxic substances
    - sulphide toxicity can be prevented provided the sulphate-sulphur concentrations in the influent is below 300-600 mg/l and the pH is kept well above 6.5 but within the optimal range for methanogenic organisms
    - ammonia toxicity can also be prevented if pH values are in the neutral range, the influent total ammonia + ammonium concentration is up to 4000 mg N/l, and the concentration of free ammonia kept well below 100 mg/l.

As discussed in section 2.3, the main characteristics of nightsoil and toilet sludge from the literature are:

- high organic concentration with a considerable portion being biodegradable (COD>20,000);
- pH values slightly above the neutral value;
- the ammonium-nitrogen concentrations ranging from 2,000 - 5,000 mg/l; and
- high solids concentration with considerable volatile portions.

Anaerobic processes are specially suited to the treatment of heavily contaminated organic wastes with COD greater than 5000 mg/l (Mudrack and Kunst, 1986). Thus with the characteristics of nightsoil and toilet sludges being within or close to the ranges optimal for the anaerobic digestion conversion process, it should be possible to stabilise nightsoil and toilet sludges by anaerobic treatment process as evidenced by the use of the process technology throughout the available literature.

2.6 Anaerobic Treatment Systems

Anaerobic treatment systems are based on the anaerobic digestion processes. Basically, there are two types of anaerobic treatment systems based on the rate of the digestion process:

1. **Low rate systems** in which the removal of organic matter is based on the settling of suspended organic solids with anaerobic digestion processes taking place in the accumulated sludge at the bottom. To this category belong the Imhoff tank, septic tank and anaerobic ponds.
2. **High rate systems** in which the removal of organic is based on intense contact between the influent organic matter and the large and active bacterial mass retained in the reactor. In this category belong the 'modern' anaerobic treatment systems e.g. upflow or downflow anaerobic filter, sludge bed reactors, conventional contact process, upflow anaerobic sludge blanket (UASB) and expanded granular sludge blanket (EGSB).

The high rate systems are further characterised by the mechanism of sludge retention in the anaerobic reactor. Basically two methods of sludge retention have found wide application in practice:

1. **Systems based on immobilisation of the sludge**. These are fixed film processes in which the anaerobic organisms form a thin sludge layer on a solid carried material which may be composed of granular material like sand which acts as a fluid bed (Jeris, 1982); or expanded bed reactors of macroscopic bodies like stones (Jewell, 1982); or an artificial medium forming a packed bed known as an anaerobic filter (Young and McCarty, 1969).
2. **Systems based on mobilisation of the sludge**. These are suspended growth processes which employ gravity settling to retain the anaerobic material mass in the treatment system. The gravity settler may be external, as for instance in the Contact Process (Coulter et al, 1957), or internal, as in the UASB (van Haandel *et al.*, 1996)

Figure 2.16 summarises the main anaerobic treatment systems and Figure 2.17 shows a schematic representation of the different high rate anaerobic reactors. Although low rate systems have been in use for the treatment of mainly domestic sewage for over a century, they were less popular than aerobic sewage treatment systems. This decreased application of earlier anaerobic treatment systems was principally due to the higher removal efficiency of organic matter achieved in aerobic systems (van

Haandel and Lettinga, 1994). However, lately, anaerobic treatment systems are becoming popular because of the breakthrough in the design of 'modern' or high rate systems and their associated advantages.

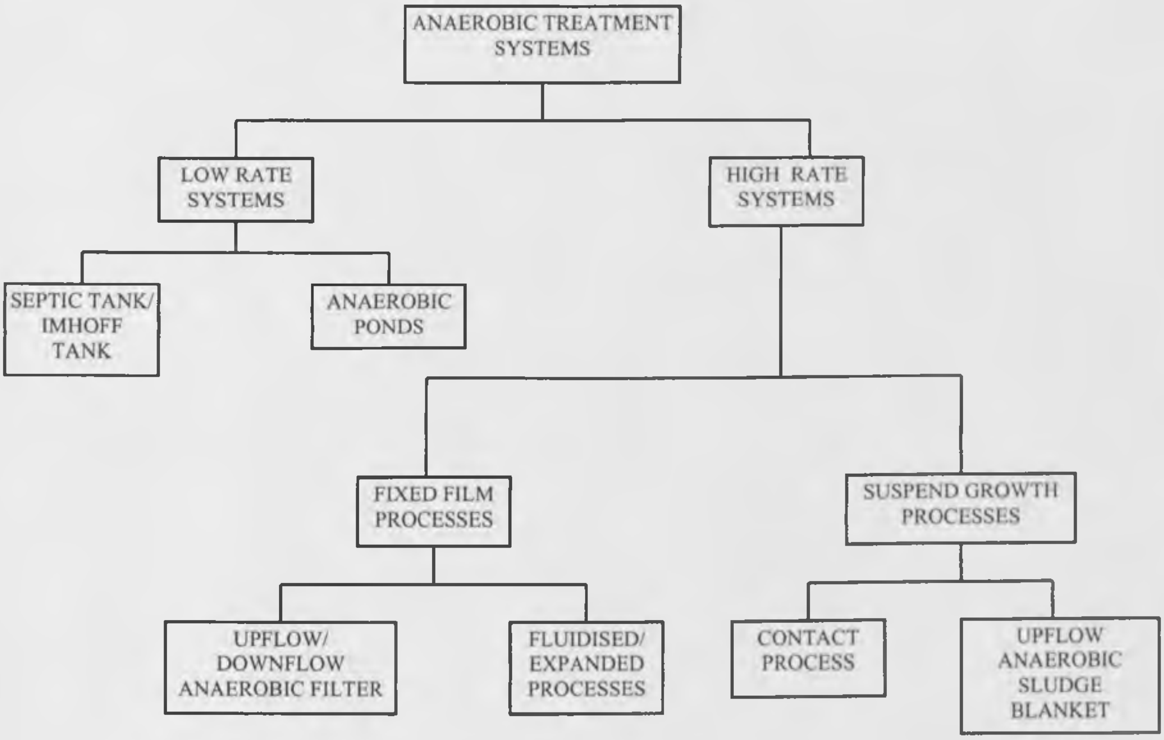


Figure 2-16 Anaerobic treatment systems

Although exceptions exist, Hall (1992) stated that in general suspended growth processes are advantageous for the treatment of sludges or wastewaters containing high proportions of particulate biodegradable organic material, while fixed film processes on the other hand are well suited to wastewaters that contain high proportions of soluble organic substrates. Thus suspended growth processes would be more suited to the treatment of nightsoil and toilet sludges since the wastes contain high proportions of solids. Of the suspended growth processes, van Haandel and Lettinga (1994) report that the UASB has been applied far more than other modern anaerobic treatment systems due to its high efficiency of organic material removal for many kinds of wastewaters, its low construction cost and land requirements, and its extremely simple operation. For these reasons and others mentioned in section 1.4, only the UASB digester was considered for this research.

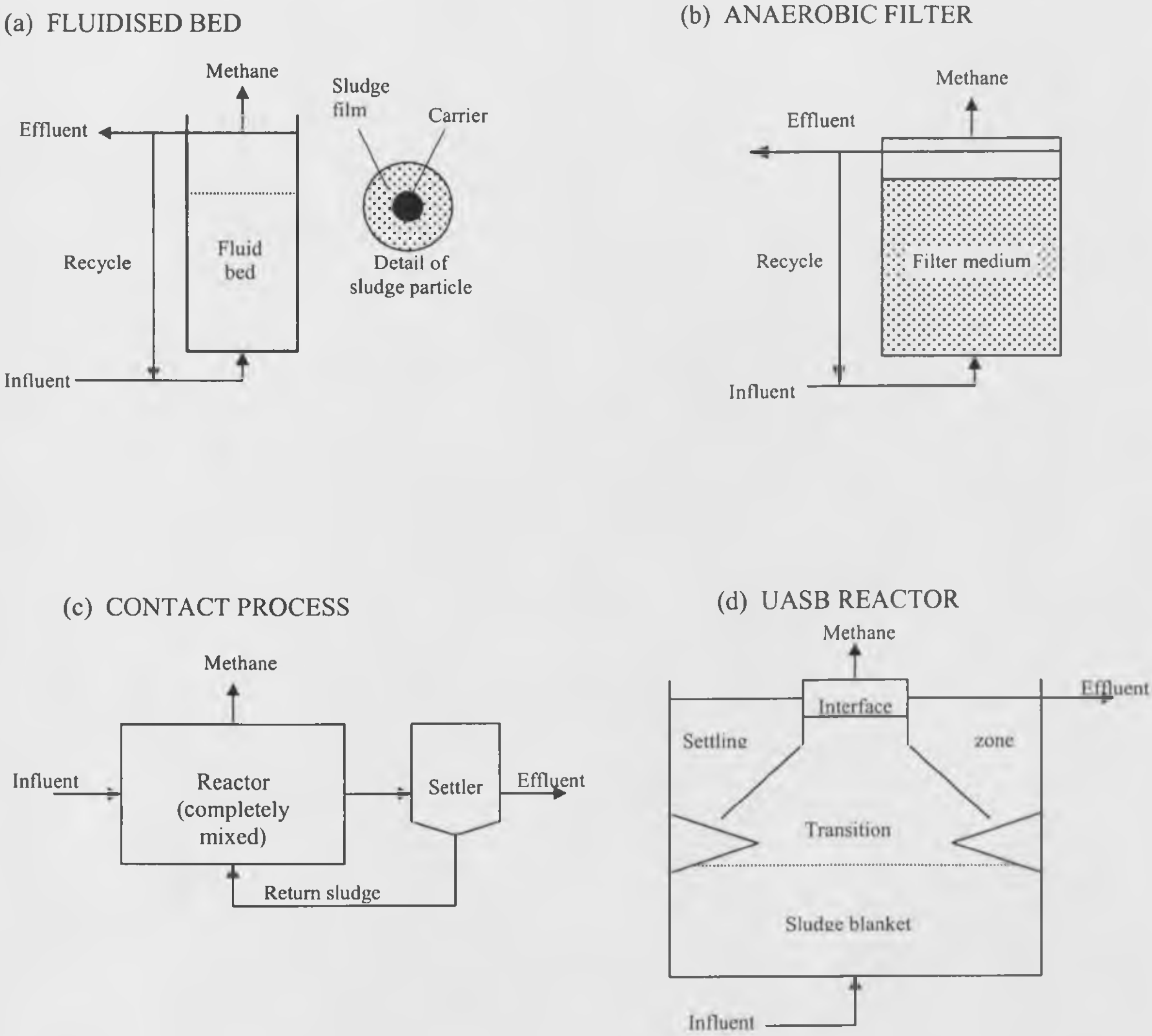


Figure 2-17 Schematic representation of different high rate anaerobic processes

Source: van Haandel *et al.* (1996)

In the following sections, descriptions of the high rate anaerobic reactors shown in Figure 2.17 are presented. The UASB reactor section is more detailed as it was the basis for this research.

2.7 The Anaerobic Contact Process

The first generation of anaerobic reactors for anaerobic biotechnology consisted of a flow-through tank with no solids recycle. The contents of these first generation or conventional anaerobic digesters were unmixed, continuously mixed, or intermittently mixed. The mixing devices used are either mechanical impeller-type systems or gas recirculation mixers. Being a flow-through tank the solids retention time (SRT), that is the average retention time of anaerobic microorganisms, was the same as the system's hydraulic retention time (HRT). Therefore, the ratio of SRT/HRT was one. The SRT could not be controlled separately from the HRT. Long and adequate SRT meant long HRT which result in large reactor volumes which can cause high capital costs. The anaerobic microorganisms required to ensure improved performance of the digester are not retained for periods longer than the HRT. This results in low treatment efficiencies.

In order to improve upon the performance of the conventional anaerobic digester and overcome some of its disadvantages, it was essential to retain the biomass required for the anaerobic digestion process much longer than the system's HRT. Separation and recycling of effluent suspended back to the anaerobic digester was subsequently incorporated. By recycling the separated biomass, the average retention time of anaerobic microorganisms in the anaerobic digester was increased beyond the system HRT. This modification to the conventional anaerobic digester was termed the anaerobic contact process (Speece (1983); Fig. 2.17(c)) and it enabled the biomass content to be controlled independently of the influent flow rate. That is, the modification enabled the system SRT to be controlled separately from the HRT.

The anaerobic contact process is applicable to a wide range of wastewater concentrations (Table 2.8). The treatment efficiency of an anaerobic contact process is usually much greater than that of a conventional digester. Total COD reductions of 90-95% are possible for highly biodegradable wastewaters with COD concentrations of 2,000 to 10,000 mg/l, and typical organic loading rates in the anaerobic contact systems are between 0.5 and 10 kg COD/m^3.d with HRTs of 0.5 to 5 days (Hall, (1992); Leslie Grady Jr. et al. (1999)). Table 2.8 summaries the advantages and disadvantages of the anaerobic contact process.

2.8 The Anaerobic Filter

For efficient anaerobic treatment of wastewaters, the concentration of biomass and SRT must be high. As part of developmental efforts to achieve these, Coulter et al. (1957) and Young and McCarty (1969) used an upflow packed column. This upflow

Table 2.8 Advantages and disadvantages of Anaerobic Contact Process

ADVANTAGES

- Suitable for wastes with high concentration of soluble organic
- Process can provide uniform substrate, temperature, and pH conditions throughout the reactor
- Reactor can be easily sampled for process monitoring
- Good internal mixing can minimise dead volume accumulation and flow channelling
- Relatively high quality effluent achievable
- Completely mixed reactor volumes can be reduced considerably in comparison to conventional anaerobic digestion
- Aerobic post-treatment sludge can be wasted to the anaerobic reactor for stabilisation.

DISADVANTAGES

- Biomass settleability is critical for successful operation
- Most suitable for wastes with low to intermediate levels of suspended solids
- Pre-treatment of biosludge may be necessary (i.e. temperature shock, vacuum degassing, etc.) to produce a settleable floc
- Relatively short anaerobic HRT results in reduced equalisation capacity for shock inputs

Source: Hall (1992)

packed column was termed the anaerobic filter (Figure 2.17b). It is also referred to as the fixed bed reactor because of the fixed media. The packing material was inert and provided contact surface for the growth of microorganisms. The physical attachment of the micro-organisms to the medium surface prevents biomass washout. In addition, the packing material reduced the Reynolds number of the influent to ensure a low turbulence and efficient sedimentation, and thus allowed the retention of unattached biomass as clumps of cells in the packing interstices. The unattached biomass in the parking interstices account for a large proportion of the retained biomass in the reactor (Speece, 1983; Young and Dahab, 1983; Wilkie et al. 1984). These processes lead to high values of SRT and biomass concentrations per unit reactor volume, and therefore an improved treatment activity.

In the early designs of the upflow anaerobic filter the media was rock-packed. These had low voidage and accumulation of unattached biomass easily plugged the bed. To accommodate the accumulation of unattached biomass without plugging the bed, the filter media in the early designs have largely been replaced by systems that incorporate synthetic packing materials. The advantage of the synthetic packing lies in their large open structures and high void volumes.

As shown in Figure 2.17b, in the anaerobic filter the influent wastewater and re-circulated effluent are distributed across the reactor cross-section and flow upward through the media. Treatment occurs as a result of the unattached and attached biomass retained by the media. Effluent exits at the top of the media section and is collected for discharge. Some effluent is re-circulated to maintain a reasonably uniform hydraulic loading on the filter in spite of varying influent flow rates.

The anaerobic filter process is mainly used for industrial wastewater treatment. Volumetric organic loading rates often range between 5 and 20 kg COD/m^3.d (van Haandel and Lettinga, (1994); Leslie Grady Jr. et al. (1999)). Leslie Grady Jr. et al. (1999) report that hydraulic retention times between 0.5 and 4 days are typical. Table 2.9 summarises the advantages and disadvantages of fixed bed anaerobic processes.

Table 2.9 Advantages and Disadvantages of Fixed Bed Anaerobic Processes

ADVANTAGES

- High biomass concentrations and long SRTs achievable
- Smaller reactor volumes due to high organic loading rates
- Relatively stable operation under variable feed conditions or toxic shocks
- Suitable for wastes with low suspended solids concentrations
- No mechanical mixing required
- Biogas evolution and effluent recycle insure relatively uniform temperature, pH, and substrate concentrations in reactor
- Land area required is relatively small

DISADVANTAGES

- Suspended solids accumulation may negatively impact reactor hydraulics and internal mass transfer characteristics
- Not suitable for high suspended solids wastewaters
- Provision may be required for periodic biomass removal
- Limited access to reactor interior for monitoring and inspection of biomass accumulation
- Relatively short reactor HRT results in reduced equalisation capacity for shock inputs
- Costs of packing material and support systems are high

Source: Hall (1992)

2.9 The Anaerobic Fluidised and Expanded Bed Systems

The anaerobic fluidised bed system (Figure 2.17a) introduced by Jeris (1982), attempts to improve reactor mass transfer characteristics by utilisation of small media particles with very high surface-to-volume ratios (Hall, 1992; Leslie Grady Jr. et al. (1999)). The anaerobic fluidised bed incorporates an upflow reactor partly filled with sand. The upflow velocity is sufficient to fluidise the sand to fill about 75% of the reactor. A very large surface area is provided by the sand, and a uniform biofilm develops on each sand grain. The internal sand grain markedly increases the net density and settling velocity of the attached biofilm and ensures efficient cell retention within the reactor. The large upflow velocities applied to ensure fluidisation promotes turbulence at the biofilm/liquid interface which in turn promotes good mass transfer into and out of the biofilm, and under some conditions exerts sufficient shear to prevent the development of thick biofilms on the media. The high upflow velocities in the fluidised bed system allow the reactor to be designed with relatively large height/diameter ratios and smaller land area requirements.

In latter fluidised bed systems, media of density lower than sand such as anthracite and plastic are used to reduce the required upflow velocity, and consequently the pumping costs. Jewell (1982) developed the expanded bed reactor that uses an upflow velocity less than that required for completed fluidisation of the granular media. Typically, in the expanded bed system, sufficient flow is applied to increase the settled bed volume by 15% to 30% (Jewell, 1982) whereas in the fluidised bed system, higher upflow velocities are utilised to produced 25% to 300% bed expansion (Sutton et al., 1983). In the expanded bed system, the particles are supported partly by the fluid and partly by contact with adjacent particles, and consequently tend to remain in the same relative positions within the bed. In the fluidised bed, the particles are supported wholly by the upward flowing fluid and move freely in the bed.

As mentioned previously, the use of small carrier particles results in high specific surface area and a high active biomass concentration. For expanded bed systems, the specific area of the carrier particles is in the 9,000 to 11,000 m^2/m^3 range, with a void volume of 45 to 55% (Hall, 1992; Iza, 1991). For fluidised bed systems the specific surface area is in the 4,000 to 10,000 m^2/m^3 range, with void volumes of 50 to 90%, depending on the degree of expansion (Leslie Grady Jr. et al. (1999)). The high specific surface areas allow high biomass concentrations to develop, in the order of 15,000 to 35,000 mg/l as VSS (similar to those achieved with the UASB process) (Hall, 1992; Iza, 1991). The high biomass concentrations allow operation at relatively low HRTs and high volumetric organic loading rates while maintaining adequate SRTs for efficient treatment. HRTs in the 0.2 to 2 day range are used, depending on the concentration of the wastewater (Leslie Grady Jr. et al. (1999)). Volumetric organic loading rates of up to 21 kg $COD/m^3.d$ are typical of the anaerobic fluidised and expanded bed reactors (Frankin et al., 1991; Sutton, 1986). Table 2.10 summarises the advantages and disadvantages of fluidised bed and expanded bed anaerobic processes.

Table 2.10 Advantages and Disadvantages of Fluidised and Expanded Bed Anaerobic Processes

ADVANTAGES

- High biomass concentrations and long SRTs achievable
- Excellent mass transfer characteristics
- Compact reactor volumes due to high organic loading rates
- May produce better effluent quality than other anaerobic treatment options
- Relatively stable operation under variable feed conditions or toxic shocks
- Suitable for wastes with low suspended solids concentrations
- No mechanical mixing required
- Biogas evolution and extensive effluent recycle insure relatively uniform temperature, pH, and substrate concentrations in reactor
- Small land area required

DISADVANTAGES

- Lengthy start-up periods may be required
- Power requirements for bed expansion or fluidisation are high
- Control of media and biomass inventories can be difficult
- Accidental washout of media can damage downstream components
- Not suitable for high suspended solids wastewaters
- Relatively short anaerobic reactor HRT results in reduced equalisation capacity for shock inputs
- Mechanical system design is relatively complex
- Cost of carrier medium is high

Source: Hall (1992)

2.10 The Upflow Anaerobic Sludge Blanket Reactor

2.10.1 Introduction

In practice, the overwhelming majority of wastewater treatment plants in developed countries use aerobic metabolism for the removal of organic material. The search for alternate treatment systems due to the high capital, operational and maintenance costs associated with aerobic systems resulted in the development of treatment systems that utilise anaerobic digestion for the removal of organic material from human wastes.

Although anaerobic treatment of wastewater is not a new technology and can be traced from the very beginnings of wastewater treatment, its lack of acceptability as an alternative wastewater treatment system has been due, principally, to the higher removal efficiency of organic matter achieved in aerobic systems (van Haandel and Lettinga, 1994). The lack of fundamental understanding of the anaerobic process was the primary obstacle to its broad implementation. This was rightly pointed out by McCarthy (1964) in his review articles. Correctly operating aerobic systems would remove 90-95 per cent of biodegradable organic matter from raw sewage while early anaerobic systems had removal efficiencies of 30–50 per cent of biodegradable matter. Later anaerobic systems like anaerobic ponds tended to have relatively high removal efficiencies (50-70%) but this was attributed to long retention times of one to five days.

The removal efficiency of the early anaerobic systems depended on the nature of sewage and settling efficiency of the system in use. Removal was based on settling of suspended organic matter. Consequently, there is little, if any contact between the anaerobic micro-organisms in the system and the non-settleable part of the organic matter in the sewage. The result being that the main part of the dissolved or hydrolysed organic matter is not metabolised and leaves the treatment system. This was a fundamental design failure. The importance of a sufficient contact between influent organic matter and the anaerobic bacterial population was not recognised at the time.

Today, the redirection of research efforts towards energy-saving alternatives like anaerobic treatment has resulted in more knowledge and understanding of the anaerobic process. Properly designed modern or high rate anaerobic treatment systems can now attain a high removal efficiency for biodegradable organic matter, even at very short retention times. These high removal efficiencies are being attributed to the presence of a high concentration of active bacterial mass within the anaerobic reactor and the intense interaction between the influent and the active bacterial mass. The upflow anaerobic sludge blanket reactor is one such modern or high rate anaerobic treatment system.

2.10.2 The UASB Reactor Concept and Process Design Considerations

Lettinga and his group at the University of Wageningen in the Netherlands developed the upflow anaerobic sludge blanket (UASB) reactor in the 1970's (Lettinga *et al*, 1980). Their UASB process resembled other USB processes described much earlier in the literature (Stander *et al,* 1967; Cillie *et al*, 1969; and Pretorius, 1971) except that:

a) sludge re-circulation and/or mechanical agitation are kept to a minimum or even completely omitted, and that, in particular,

b) the reactor is equipped in the upper part with a proper system for gas-solids separation.

A schematic diagram of an UASB reactor is shown in Figure 2.18. Lettinga *et al* (1980) state basic ideas underlying the process as:

a) The anaerobic sludge develops and maintains superior settling characteristics if chemical and physical conditions favourable to sludge flocculation and to the maintenance of a well flocculated sludge are provided.

b) A sludge blanket (bed) may be considered as a separate - more or less - fluid phase with its own specific characteristics. A well-established sludge blanket frequently forms a rather stable phase, capable of withstanding relatively high mixing forces. The redispersion of the sludge in the liquid phase therefore may require a significant amount of mixing energy.

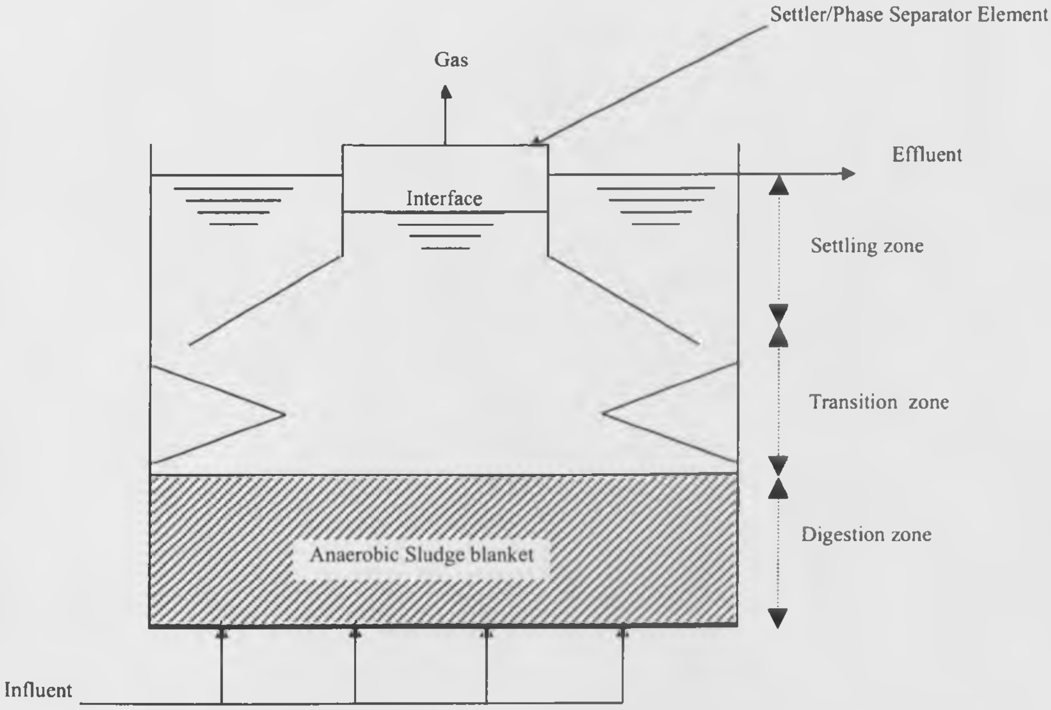


Figure 2-18 Schematic diagram of UASB reactor
Source: Adapted from van Haandel *et al* (1996)

c) The washout of discrete sludge particles (flocs) released from the sludge blanket can be minimised by creating a quiescent zone within the reactor, enabling the sludge particles to flocculate, settle, and/or be entrapped in a secondary sludge blanket (present in the settler compartment).

These ideas are incorporated into the process design of the UASB to meet the basic requirements for a high rate anaerobic wastewater treatment system in the following ways (van Haandel *et al,* 1996):

(a) For the conversion of organic influent material the UASB relies on the formation of well settleable, flocculent or granular type of anaerobic sludge forming a blanket in the bottom section of the reactor, known as the digestion zone. The influent is uniformly distributed over the reactor bottom and follows an upward path to the level of effluent abstraction at the top of the reactor. As the influent passes through the sludge blanket, the organic material is taken up and metabolised by the sludge and to a large extent transformed into biogas.

(b) For the required intense contact between the influent organic material and the bacteria in the sludge, the system relies on agitation caused by the rising biogas bubbles and the kinetic energy of the influent when it enters the reactor. Under most circumstances the natural agitation will be sufficient for good contact between the organic material and the bacteria, so that mechanical mixing is not applied.

(c) A large mass of well-settleable sludge can be retained in the reactor by installing a separator to separate the three phases in the reactor: gas, G (biogas); liquid, L (the effluent); and solid, S (the sludge) in the top part of the reactor

The GLS phase separator is the most characteristic device of the UASB reactor. It divides the reactor into two parts: an upper settling and a lower digestion zone. The phase separator captures the biogas production so that the settling zone is tranquil and sludge particles eventually carried by the liquid flow can settle out and accumulate on the separator elements. Due to the inclined surface of the separator, the settled sludge end up sliding back into the digestion zone of the reactor and once again take part in the degradation of the influent organic material.

2.10.3 Design Criteria

At present, there is no design criteria for UASB treating faecal sludges. The design criteria described in the literature apply to UASB reactors treating sewage and some industrial wastewater. These criteria have been assumed to be applicable to faecal sludges and are used in this research.

2.10.3.1 Loading Rates

The load on a UASB reactor is limited to either the organic or the hydraulic load depending on the nature of the wastewater.

Hydraulic Load

Hydraulic load can be defined as the volume of influent material per unit time. The hydraulic load is thus numerically equal to the average influent flow flowrate. Hydraulic load is used in the design when the wastewater is of a relatively low strength such as domestic sewage. The maximum hydraulic load is limited by the constraint that the upflow velocity in the reactor must not cause excessive sludge wash-out. This upflow velocity usually should not exceed 1 m/h in the UASB reactor and its is calculated as follows:

$$v_i = \frac{Q_i}{A} = \frac{V_r}{A.(HRT)} = \frac{H}{(HRT)} \qquad (2.10)$$

where:

v_i = liquid upward velocity (m/h);
Q_i = average wastewater flow (m^3/h);
A = surface area of the UASB reactor (m^2);
V_r = volume of the reactor (m^3);
(HRT) = hydraulic retention time (h); and
H = height of the UASB reactor (m).

Equation 2.9 further shows the relationship between the upflow velocity, the height of the UASB reactor and the hydraulic retention time. van Haandel and Lettinga (1994) state that from available experimental results, an average retention time of six hours is sufficient in tropical and subtropical regions (T > 18 °C) to achieve a satisfactory treatment efficiency in one compartment UASB reactors.

Organic Load

For concentrated wastewaters, the organic load rather than the hydraulic load becomes the determining factor in the design of the reactor. The organic load (L_o) is defined as the mass of influent organic material per unit time and the specific organic load (l_o) is the mass of influent organic materials per unit time and per unit of reactor volume. The specific organic load is expressed as kilograms COD (applied) per unit reactor volume and per unit time. The specific organic load is calculated as follows:

$$l_o = \frac{L_o}{V_r} = \frac{Q_i S_{ti}}{V_r} = \frac{S_{ti}}{(HRT)} \qquad (2.11)$$

where

l_o = applied specific COD load (kg COD m^{-3} d^{-1})
L_o = organic (COD) load (kg COD d^{-1})
V_r = volume of the reactor (m^3);

Q_i = average wastewater flow (m^3/d);

S_{ti} = influent organic material (COD) concentration (kg/m^3)

(HRT) = hydraulic retention time (d)

van Haandel and Lettinga (1994) state maximum design organic load of organic material may be 20 kg COD/m^3d for wastes containing a high concentration of dissolved organic material of vegetable origin to be digested at or near the optimal temperature for mesophilic digestion.

2.10.3.2 Physical Design considerations

Shape of reactor

UASB reactors are either circular or rectangular in cross section. Circular reactors have the advantage of higher structural stability but are more difficult to construct than a rectangular or square unit. For this reason large UASB reactors are generally constructed in a rectangular or square cross sections and small reactors are generally constructed in cylindrical shape. Furthermore, when more than one reactor unit is constructed, the rectangular shape is advantageous because sidewalls can be shared by different units.

Height (or depth) of reactor

In practice, the choice of the appropriate height (or depth) of the reactor depends on the required performance and economic considerations. A higher depth/volume ratio reduces the required area for the treatment and thereby increases the upflow velocity which results in increased turbulence in the system and hence better contact between biological sludge and incoming wastewater. High upflow velocities can result in excessive sludge washout. The greater the depth of the UASB reactor, the higher the static pressure. High static pressures causes an increase in the solubility of carbon dioxide which may result in a depression of pH. If the pH assumes a lower than optimum value the anaerobic digestion can be jeopardised. Most UASB reactors have a height (or depth) between 4 and 6 m. This range has proved to be the economic optimum and to be adequate from the process point of view (van Haandel and Lettinga, 1994).

2.10.4 Start-up of UASB Reactor Treating Domestic Wastewater

The start-up of an anaerobic treatment process is time consuming and sometimes rather difficult compared to an aerobic treatment process. This is due to the slow growth rate of anaerobic bacteria and adaptation of the bacterial mass to the particular characteristics of the wastewater to be treated. Domestic sewage however differs from other wastewaters of industrial origin in that it already contains the bacterial populations necessary for anaerobic digestion. Thus, a reactor for anaerobic treatment of domestic sewage can be started without the need for inoculation. The bacterial populations for acid and methanogenic digestion develop spontaneously. To cut down on the length of time required for the start-up, the reactor may be seeded.

Reporting on experimental results obtained from a 160 m^3 full-scale UASB reactor treating sewage from the Pedregal township in Campina Grande, Brazil, van Haandel and Lettinga (1994) observed that during the start-up period:

1. **Organic matter removal** measured by BOD and COD removal efficiencies during the initial period of operation were low and even tended to decrease during the first 10 weeks of operation. However after this initial period, the performance of the reactor started to improve and after 20 weeks of operation high and almost constant BOD and COD removal efficiencies were established (Figure 2.19). They attributed the low and decreasing removal efficiency during the initial operation period to the absence of a sufficient quantity of proper bacterial sludge to carry out the anaerobic digestion of the organic material. The bacterial populations develop gradually and sequentially in the reactor, a logical consequence of the fact that the metabolic products of one step form the substrate for the next step. This explains the declining efficiency of organic material removal in the initial period. When the acidogenic, acetogenic and methanogenic bacterial populations are too small, the solubilised products cannot be sufficiently converted and tend to appear in the effluent.

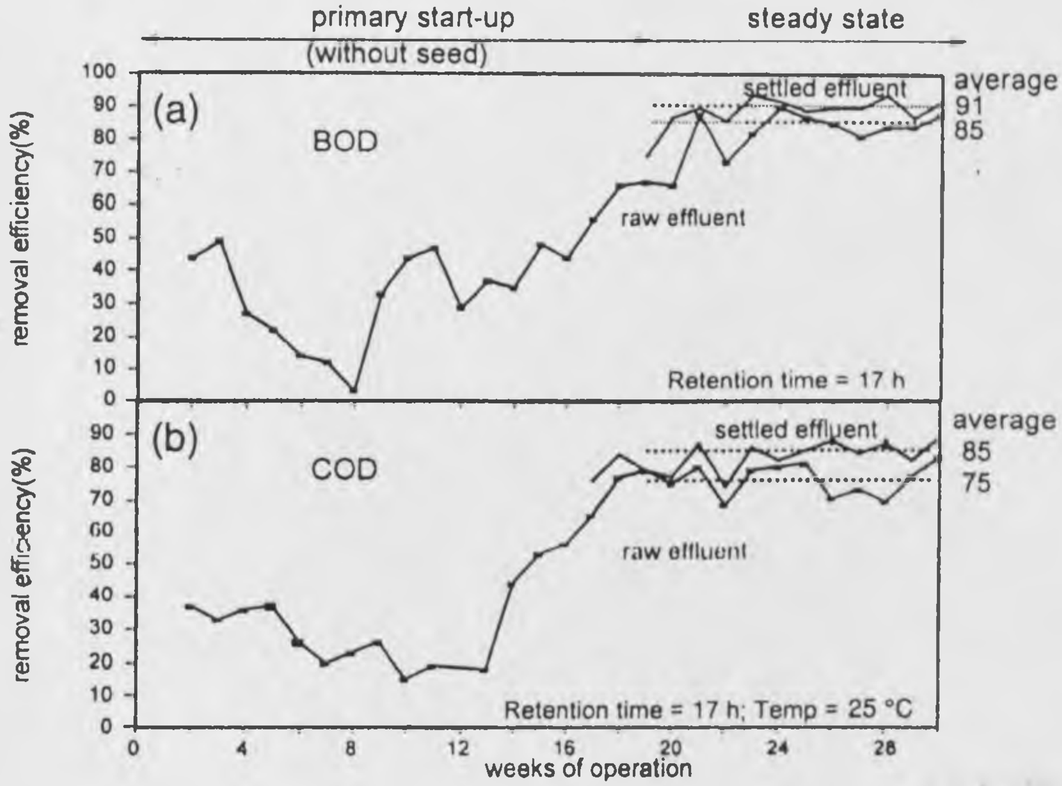


Figure 2-19 BOD and COD removal efficiencies in the Pedegral UASB reactor during the first 30 weeks of operation

Source: van Haandel and Lettinga (1994)

2. **Total suspended solids removal efficiency** increased with time during the initial period of operation (Figure 2.20 (b)). This, as explained by van Haandel and Lettinga (1994), is due to the fact that the amount of sludge present in the reactor was small during the initial operation period, and therefore the likelihood of entrapping suspended solids was reduced. However, as the process proceeded, the amount of sludge mass grew and with it the improvement in the entrapment of the non-settleable suspended solids.

3. **pH value and stability:** The volatile fatty acid (VFA) concentrations in both the influent and effluent were small. During the initial period of operation when the amount of sludge in the reactor was small, the VFA concentration tended to increase, indicating that acid fermentation proceeded at a higher rate than methanogenesis. However, after an adequate amount of sludge accumulated in the reactor, the effluent VFA concentration became smaller than the influent concentration. Furthermore, the effluent pH showed little variation over the

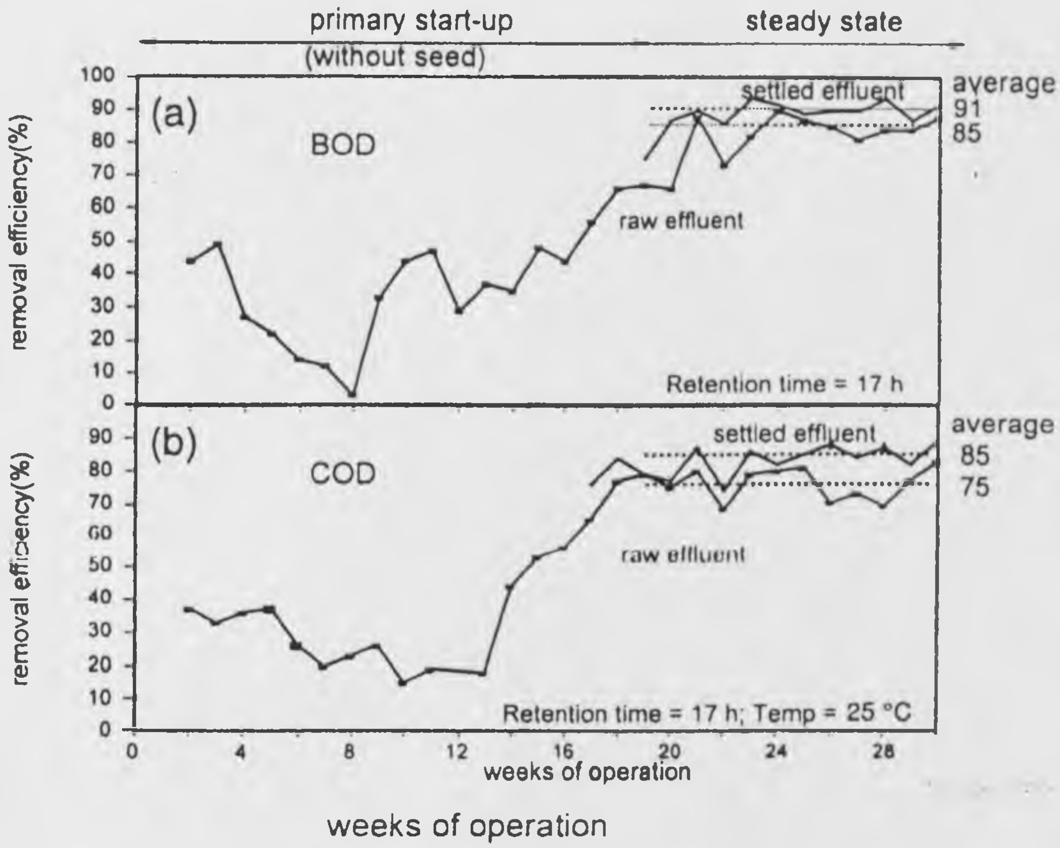


Figure 2-20 **Settleable solids and TSS removal efficiencies in the Pedegral UASB reactor during the first 30 weeks of operation**

Source: van Haandel and Lettinga (1994)

period of operation. There was a slight tendency for the pH to decrease during the first weeks of operation, when acid fermentation prevailed over methanogenesis. From their results, at no stage during the start-up was there any risk of souring the reactor and they conclude that, in general, souring of contents of a UASB reactor is not a problem in anaerobic sewage treatment and that there is no need for chemical pH adjustment.

4. **Nutrient removal:** the results of their determinations of nitrogen and phosphorus concentrations showed that the process resulted in an increase of the nutrient concentrations. The increase was attributed to the mineralisation of organic compounds containing organic nitrogen and phosphorus.

2.10.5 Steady-State Behaviour in the UASB Reactor

During the start-up period of the operation of the UASB reactor, the correct bacterial populations capable of converting organic material into methane develop and grow. Sludge mass begins to accumulate in the reactor and the extent of accumulation of this sludge is limited by the physical size of the UASB reactor. Some time after the beginning of the operation, the UASB reactor becomes filled up with sludge. When this condition is attained, there are two basic ways of dealing with the sludge production of the system (van Haandel and Lettinga, 1994). The first is to discharge the sludge periodically so that the concentration of settleable solids in the effluent remain as low as possible. The second option is to operate the reactor at maximum sludge hold-up, consequently accepting the wash-out of excess sludge. In this latter mode, the concentrations of settleable solids will be relatively high. In practice, the first option is adopted if the UASB reactor is the only biological treatment unit. This ensures that the effluent COD and TSS concentrations as low as possible. The second option is adopted in practice when some kind of post treatment, such as a waste stabilisation pond, is employed.

With respect to sludge build-up, the start-up period can be considered as being complete and steady state established when the sludge mass present in the reactor remains constant, both qualitatively and quantitatively and the effluent quality remains constant at the design load. After the establishment of a steady state, the total daily flux of settleable solids in the effluent is equal to the daily sludge production rate.

With regards to the organic material (COD) present in the waste, a steady state is established when organic matter does not accumulate in the treatment system. When this is attained, the daily mass of influent COD is equal to the sum of the daily mass of COD leaving the system as methane in the excess sludge produced, in the effluent, and the daily amount of COD oxidised.

2.10.6 Mass Balance Equation

Within the UASB reactor (control volume) the mass balance for any given constituent takes the from:

(Net rate of accumulation in the control volume)

= {(rate of flow into the control volume)
- (rate of flow out of the control volume)
+ (net rate of generation in the control volume)}

or simply

Accumulation = input – output + generation.

Each term in the mass balance equation has the units of mass/time. The generation term represents the sum of all reactions in which the constituent of interest participates. If the generation term is positive, the constituent is being produced in the control volume; if it is negative, the constituent is being destroyed.

The organic material (COD) present in the influent , after having being exposed to anaerobic digestion in the UASB reactor will have one of the following forms: (1)sludge COD; (2) methane COD; (3) mineralised COD and (4) remaining COD in the effluent. At a steady-state when organic matter does not accumulate in the treatment system, the daily mass of influent COD is equal to the sum of: (i) the daily mass of COD leaving the system as methane, (ii) in the excess sludge produced, (iii) in the effluent, and (iv) the daily amount of COD oxidised.

$$MS_{ti} = MS_{te} + MS_{tx} + MS_{tm} + MS_{to} \tag{2.12}$$

where

MS_{ti} = daily mass of influent COD
MS_{te} = daily mass of effluent COD
MS_{tx} = daily mass of COD in the discharged sludge
MS_{tm} = daily mass of COD in produced methane
MS_{to} = daily mass of oxidised COD

The COD recovery factor, B_o, is given by

$$B_o = \frac{MS_{te} + MS_{tx} + MS_{tm} + MS_{to}}{MS_{ti}} \tag{2.13}$$

Theoretically, the value of $B_o = 1.00$, but due to errors in the determination of various terms of equation 2.12 and to the fact that the treatment systems usually are not operated under rigorously steady state conditions, the experimental value of Bo deviates from its theoretical value. The magnitude of the deviation being an indication of the accuracy of the experimental procedures.

2.10.7 Biogas Production

The anaerobic digestion of organic material produces biogas, which consists mainly of methane and carbon dioxide. The biogas may also contain small amounts of nitrogen, hydrogen, hydrogen sulphide, and water vapour. The amount of biogas released can vary over a wide range and depends on the concentration of biodegradable organic material and the biological activity in the digester. Due to the high proportion of organic materials in sewage, the methane content in biogas from sewage digesters is always high. It is reported in the literature (van Haandel and Lettinga (1994); Speece (1996)) that, typically, biogas from sewage digesters contains about 65-80 percent methane by volume, and the remainder is made up of a mixture of carbon dioxide, nitrogen, water vapour and a small fraction of hydrogen sulphide. Due to the high proportion of biodegradable organic material in nightsoil and toilet sludge, the methane content in the biogas resulting from the anaerobic digestion would be expected to be high.

A considerable portion of the biogas produced remains dissolved in the liquid phase (particularly the case for carbon dioxide) and leaves the system in the effluent. In addition, methane may also be lost due to desorption at the liquid surface, the loss depending on the size of the surface and whether or not the reactor is covered. Owing to gas losses the mass of collected methane is usually much smaller than the amount produced, which can be calculated from stochiometry. van Haandel and Lettinga (1994) reports that in practice, the losses may be between 20 and 50 percent of the produced biogas. In cases where the methane produced could have some commercial value it is essential to design a good gas collection system to minimised the losses.

Theoretically, the volume of methane produced can be calculated knowing the COD equivalence of methane. The COD equivalent of methane is as follows:

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \qquad (2.14)$$

Knowing the atomic weights of H (1g/mol), C (12g/mol) and O (16g/mol), the above equation shows that for each mole of methane (16 g) consumed, two moles of oxygen equivalent (64 g) are destroyed. That is

$$COD_{CH_4} = \frac{64}{16}\,\text{g COD/g CH}_4 = 4\,\text{g COD/g CH}_4 \tag{2.15}$$

Thus the digestion of 1 kg COD results in a mass of ¼ kg = 250 g of CH_4. 1 mole (16 g) of CH_4 has a volume of (22.4*T*)/273 litres at atmospheric pressure (where T = temperature in K). Hence the volume of methane gas per 1 kg COD is calculated as:

$$= \left(\frac{250}{16}\right) * \left(\frac{22.4T}{273}\right) = 1.28T\ \text{l CH}_4\,/\,\text{kg COD} \tag{2.16}$$

If the partial pressure of methane is p_m and the fraction of collected methane is f_m, the volume of biogas produced, v_b, is given by

$$v_b = \left(f_m * \frac{1.28T}{p_m}\right)\text{l CH}_4/\ \text{kg COD digested} \tag{2.17}$$

The following example, after van Haandel and Lettinga (1994), illustrates how these equations are applied. Assuming a daily per capita contribution of 62.5 g COD and 75 l water from domestic sewage, a digestion efficiency of 80 per cent, a methane recovery fraction of 2/3, and a methane partial pressure of 0.75 atm at 27 °C, the expected methane production from the UASB digester is calculated as follows:

Temperature in K, $T = 273 + 27 = 300$

Methane recovery fraction, $f_m = \frac{2}{3}$

Methane partial pressure, p_m (at 300 K) = 0.75

Biogas yield per kg COD, $v_b = \left(\frac{2}{3} * \frac{1.28 * 300}{0.75}\right)$=341 l/kg COD

Amount of COD digested per litre of sewage = $0.8 * \frac{62.5}{75} = 0.667$ g/l = 667 mg/l

Solubility of methane at atmospheric pressure = 20 mg/l

Solubility of methane in digester = 0.75 * 20 = 15 mg/l

From equation 2.12, COD in liquid phase due to methane = 4 * 15 mg/l

Maximum amount of COD associated with desorbed methane = 667 – 60 = 607 mg/l

Per capita COD mass corresponding to desorbed methane = $75\frac{l}{d}*0.607\frac{g}{l} = 45.5$ g/d

Biogas yield per capita = 341 l/kg COD * 0.0455 kg COD/d = 15.4 l/cap d.

Biogas yield per unit volume of influent = $\frac{15.4l}{75l} = 0.21$ l biogas/l sewage.

The biogas produced can be used as fuel or flared off. Use of the gas as a source of fuel is only feasible when the biogas production is high. Biogas production from the anaerobic digestion of nightsoil and toilet sludge is expected to be high due to the high biodegradable organic content in the waste.

2.10.8 Applicability and present use of the UASB process

Although originally the UASB reactor was developed for treating medium strength (5,000 to 10,000 mg COD/L) types of industrial wastewaters with a low suspended solids concentration, the system has been shown to be quite applicable for other wastes. van Haandel *et al.* (1996) suggest that the influent characteristics that may limit the applicability of the UASB reactor for wastewater treatment are low temperature, a high concentration of suspended solids and presence of toxic compounds. They state that in practice an influent suspended solid concentration beyond 4,000 to 6,000 mg/L becomes non-applicable.

Table 2.11 shows the different wastes that have been successfully treated in full-scale UASB reactors or their variants while Table 2.12 shows in particular the application of UASB reactors for sewage treatment. The period column in Table 2.12 refers to the period for which either the experiment was conducted or the USAB plant was monitored. It is not very clear from the reference if the start-up period was included in the experimental or monitoring period. Tables 2.11 and 2.12 show that the UASB has not been used specifically in the treatment of faecal sludges.

Judging from the characteristics of faecal sludges, it should be possible to use the UASB process to treat faecal sludges after an initial physical pre-treatment to remove grit, large inorganic and non-digestible material from the faecal sludge followed by a significant amount of dilution.

Table 2.11 **Types of wastewaters treated in UASB reactors**

Wastewater	No.	Volume (m^3)
Alcohol	20	52 000
Bakers' yeast	5	9 900
Bakery	2	347
Brewery	30	6 600
Candy	2	350
Canneries	3	800
Chemical	2	600
Chocolate	1	285
Citric Acid	2	6 700
Coffee	2	1 300
Dairy and cheese	6	2 300
Distillery	8	24 000
Domestic Sewage	10	10 000
Fermentation	1	750
Fructose production	1	240
Fruit juice	3	4 600
Landfill leachate	6	2 500
Paper	28	67 200
Pharmaceutical	2	600
Potato processing	27	25 600
Rubber production	1	650
Slaughterhouse	3	950
Sludge liquor	1	1 000
Soft drinks	4	1 380
Starch (barley, corn, wheat)	16	33 500
Sugar processing	19	21 100
Vegetable and fruit	3	2 800
Yeast	4	8 550
Total	205	339 610

Source: van Haandel *et al.* (1996).

Table 2.12 Application of upflow anaerobic reactors to sewage treatment

Place	Vol. (m^3)	Temp. (°C)	Influent concentrations (mg/L)			Inoculum
			COD	BOD (COD_{sol})	TSS	
South Africa	0.008	20	500	(148)	NP	Activated sludge
Netherlands	0.030	21	520-590	(73-75)	NP	Digested sewage sludge
Netherlands	0.120	12-18	420-920	(55-95)	NP	Digested sewage sludge
Netherlands	0.120	18-20	248-581	(163-376)	NP	Granular sludge
Netherlands	0.120	7-18	100-900	53-474	10-700*	Granular sludge
Netherlands	6	10-18	100-900	53-474	10-700*	Granular sludge
Netherlands	20	11-19	100-900	53-474	10-700*	Granular sludge
			150-5500	43-157	50-400*	
Colombia	64	25	267	95	NP	Digested cow manure
Netherlands	0.120	12-20	190-1180	(80-300)	NP	Granular sludge
Netherlands	0.116	12-20	150-600	(70-250)	NP	Granular sludge
Mexico	0.110	12-18	465	NP	154	Adapted aerobic sludge
Brazil	0.120	19-28	627	357	376	None
Italy	336	7-27	205-326	55-153	100-250	None
India	1200	20-30	563	214	418	None
Netherlands	120	>13	391	(291)	-	Granular sludge
Netherlands	205	16-19	391	(291)	-	Self cultivated on sand
Colombia	35	NP	NP	NP	NP	NP
Netherlands	1.2	13.8	976	454	641*	Digested sewage sludge
Netherlands	1.2	12.9	821	467	468*	
Netherlands	1.2	11.7	1716	640	1201*	Granular sludge
Indonesia	0.86	NP	NP	NP	NP	NP
Indonesia	0.86	NP	NP	NP	NP	NP
Thailand	0.030	30	450-750	NP	NP	Different sludges
Brazil	120	18-28	188-459	104-255	67-236	Granular sludge
Colombia	3360	24	380	160	240	none
Brazil	67.5	16-23[a]	402	515	379	Digested sludge
Netherlands	0.200	15.8	650	346	217	Digested sludge
Netherlands	0.120	15.8	397	254	33	Granular sludge
Puerto Rico	0.059	≅20	782	352	393	Digested sludge
India	12000	18-32	1183	484	1000	NP
India	6000	18-32	404	205	362	NP
Brazil	477	NP	600	NP	303	Non adapted sludge

NP: not provided; $_{sol}$: soluble; [a]: air temperature; *: expressed as COD

Source: Seghezzo *et al.*, (1998)

Table 2.12 Applications of upflow anaerobic reactors for sewage treatment (continued)

Place	HRT. (h)	Removal efficiencies in the reactor (%)			Start-up (months)	Period (months)
		COD	BOD (COD_{sol})	TSS		
South Africa	24	90	(49)	60-65	1	1
Netherlands	9	57-79	(50-60)	30-70	NP	1
Netherlands	32-40	48-70	(30-45)	90	NP	3
Netherlands	12	72	(62)	NP	NP	17
Netherlands	4-14	45-72	(38-59)	50-89	NP	12
Netherlands	9-16	46-60	(42-48)	55-75	NP	12
Netherlands	6.2-18	31-49	(23-46)	NP	NP	12
Colombia	6-8	75-82	75-93	70-80	6	9
Netherlands	7-8	30-75	(20-60)	NP	NP	NP
Netherlands	2-3	NP	(20-60)	NP	NP	NP
Mexico	12-18	65	NP	73	NP	>12
Brazil	4	74	78	72	4	9
Italy	12-42	31-56	40-70$^+$	55-80$^+$	NP	12
India	6	74	75	75	2.5	12
Netherlands	2-7	16-34	(20-51)	None	NP	35
Netherlands	1.5-5.8	≅ 30	(≅ 40)	None	NP	33
Colombia	5-19	66-72	79-80	69-70	NP	48
Netherlands	44.3	33	50	47.0*	NP	28
Netherlands	57.2	3.8	14.5	5.8*	NP	24
Netherlands	202.5	60	50	77.1*	NP	13
Indonesia	360	90-93	92-95	93-97	NP	60
Indonesia	34	67-77	Up to 82	74-81	NP	60
Thailand	3-12	90	NP	NP	>2	4
Brazil	5-15	60	70	70	>2	24
Colombia	5.0	45-60	64-78	≅ 60	>6	>36
Brazil	7.0	74	80	87	NP	14
Netherlands	3.0	37-38	26.6	83	None	5
Netherlands	2.0	27-48	(32-58)	NP	None	3
Puerto Rico	6-24	57.8	NP	76.9	≅ 4	16
India	8	51-63	53-69	46-64	5	13
India	8	62-72	65-71	70-78	5	11
Brazil	13	68	NP	76	2	>7

NP: not provided; $_{sol}$: soluble; ^{a}: air temperature; *: expressed as COD; $^+$: obtained at temperatures of 15-20°C, HRT of 12 h and V_{up} of 0.58 m/h

Chapter Three

3 Materials and Methods

3.1 Introduction

The chapter is focused on the materials and methods used for the experimental work carried out for this research project. The experimental work for the research project was carried out in three phases as follows:

Phase1: Characterisation of faecal sludges in Kumasi

Phase 2: Anaerobic digestion of primary sludge arising from domestic sewage using a laboratory-scale UASB in Leeds, UK.

Phase 3: Anaerobic digestion of faecal sludges in Kumasi, Ghana.

Description of the materials and methods used for each of the phases follows.

3.2 Phase 1: Characterisation of Faecal Sludge in Kumasi, Ghana

The first stage of the research work involved the characterisation of faecal sludges in Kumasi, Ghana. This was carried out between mid-April to mid-July 1998. The characterisation involved laboratory analyses of samples of faecal sludges taken from truck tankers discharging the waste in a manner described in Chapter One.

3.2.1 Raw Material

The raw material for this first phase of the work was faecal sludge. The faecal sludge came mainly from public toilets - bucket latrines, non-flush aqua privies, Kumasi ventilated improved pits latrines (KVIPs) – and septic tanks serving both household and communal water closets. In most cases, the tanker truckloads were mixtures of faecal sludge from various sources as tanker trucks tried to maximise their income from desludging activities. This being the reality of the situation, samples taken were all classified as faecal sludge and hence no distinction was made as to whether it was purely nightsoil, toilet sludge or septage.

3.2.2 Sampling

When sampling, the techniques used must ensure that representative samples are obtained because the data from the analysis of the samples will serve as a basis for designing the pilot treatment system. Given the very nature of faecal sludge and mode in which it is collected, transported and discharged, it was crucial to obtain a good representative sample when collecting the faecal sludge from the discharging tanker truck. This was done by taking three 5-l grab samples and then mixing them before taking samples for the laboratory analysis. The three 5-l samples were taken at the start of the discharge, midway through the discharge and towards the end of the discharge from the tanker trucks.

3.2.3 Laboratory Analysis

The samples were then taken to the Environmental Quality Engineering (EQE) laboratory located within the Department of Civil Engineering at the Kwame Nkrumah University of Science and Technology (KNUST), Kumasi, for analysis. The samples were analysed for the following physical and chemical characteristics:

1. Biochemical oxygen demand (BOD)
2. Chemical oxygen demand (COD)
3. Total solids (TS)
4. Total volatile solids (TVS)
5. Ammonia nitrogen (NH_3-N)
6. pH

3.2.3.1 Biochemical Oxygen Demand (BOD)

Biochemical oxygen demand (BOD) is usually defined as the amount of oxygen required by bacteria while stabilising decomposable organic matter under aerobic conditions. The BOD test is essentially a bioassay procedure involving the measurement of oxygen consumed by living organisms (mainly bacteria) while utilising the organic matter present in waste, under conditions as similar as possible to those that occur in nature.

The BOD was determined using the 5-day BOD test described in method 5210 B in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". The method consists of filling with sample, to overflowing, an airtight bottle of 300 ml capacity and incubating it at 20°C for 5 days. Dissolved oxygen (DO) is measured initially and after incubation, and the BOD is computed from the difference between the initial and final dissolved oxygen. The DO was measured using the Wrinkler or iodometric method and its modifications as described in method 4500 O B in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". Because the sample contains a large proportion of micro-organisms, seeding was not necessary. When dilution water is not seeded, the BOD_5 is calculated as follows:

$$\mathrm{BOD}_5, \mathrm{mg/l} = \frac{D_1 - D_2}{P}$$

where:

D_1 = DO of diluted sample immediately after preparation, mg/l,
D_2 = DO of diluted sample after 5 d incubation at 20°C, mg/l
P = decimal volumetric fraction of sample used.

3.2.3.2 Chemical Oxygen Demand (COD)

The chemical oxygen demand (COD) is used as a measure of the oxygen equivalent of the organic matter content of a sample that is susceptible to oxidation by a strong chemical oxidant in an acidic medium. During the determination of COD, organic matter is converted to carbon dioxide and water regardless of the biological assimilability of the substances. This represents a major limitation of the COD test in that one cannot differentiate between biologically oxidizable and biologically inert organic matter. However for samples from a specific source, COD can be related empirically to BOD.

The COD was determined using the dichromate open reflux method as described in method 5220 B in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". As stated in the "Standard Methods for the Examination of

Water and Wastewater" the dichromate reflux method is preferred over procedures using other oxidants because of superior oxidising ability, applicability to a wide variety of samples, and ease of manipulation.

The sample is refluxed in a strongly acid solution with a known excess of potassium dichromate ($K_2Cr_2O_7$). The reaction involved may be represented in a general way by the following chemical equation:

$$C_nH_aO_b + cCr_2O_7^{2-} + 8cH^+ \xrightarrow{\text{heat + catalyst}} nCO_2 + \frac{a + 8c}{2} H_2O + 2cCr^{3+}$$

$$\text{where } c = \frac{2}{3}n + \frac{a}{6} - \frac{b}{3}$$

After digestion, the remaining unreduced $K_2Cr_2O_7$ is titrated with ferrous ammonium sulphate (FAS) {$Fe(NH_4)_2(SO_4)_2$} to determine the amount of $K_2Cr_2O_7$ consumed. The reaction between FAS and dichromate may be represented as follows:

$$6Fe^{2+} + Cr_2O_7^{2-} + 14H^+ \rightarrow 6Fe^{3+} + 2Cr^{3+} + 7H_2O$$

The oxidizable organic matter is calculated in terms of oxygen equivalent as follows:

$$\text{COD as mg } O_2/l = \frac{(A - B) \text{ x } M \text{ x } 8000}{\text{ml sample}}$$

where:

A = ml FAS used for blank,
B = ml FAS used for sample, and
M = molarity of FAS.

3.2.3.3 Total Solids (TS)

"Total solids" is the term applied to the material residue left in the vessel after evaporation of a sample and its subsequent drying in an oven at 103 to 105 °C. The total solids content of the faecal sludge was determined using the total solids dried at

103 – 105 °C method described in method 2540 B in the 19th edition of "Standard Methods for the Examination of Water and Wastewater".

A measured well-mixed sample is evaporated in a weighed dish and dried to constant weight in an oven at 103 – 105 °C. The increase in weight over that of the empty dish represents the total solids. The total solids is calculated as follows:

$$\text{TS mg/l} = \frac{(\text{A - B}) \times 1000}{\text{sample volume, ml}}$$

where:

A = weight of dried residue + dish, mg, and
B = weight of dish, mg.

3.2.3.4 Total Volatile Solids (TVS)

The weight loss when the total solids, as determined in the previous section, is ignited at 550 ± 50 °C is called "total volatile solids". When the total solids are ignited, the organic fraction oxidises and is driven off as gas at the ignition temperature, and the inorganic fraction remains behind as ash. Volatile solids analysis is applied most commonly to wastewater sludges to measure their biological stability.

The volatile content of the total solids is determined by igniting at 550 °C as described in method 2540 E in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". The total volatile solids is calculated as follows:

$$\text{TVS mg/l} = \frac{(\text{A - B}) \times 1000}{\text{sample volume, ml}}$$

where:

A = weight of residue + dish before ignition, mg, and
B = weight of residue + dish after ignition, mg.

3.2.3.5 Ammonia Nitrogen (NH_3-N)

The forms of nitrogen of greatest interest with respect to water and wastewater are nitrate, nitrite, ammonia and organic nitrogen. In the determination of these various forms of nitrogen, it is customary to report all results in terms of nitrogen so that values may be interpreted from one form to another without the use of a factor. The term ammonia nitrogen thus refers to all nitrogen that exists in aqueous solution as either the ammonium ion or ammonia, depending on the pH of the solution (as described in section 2.5.4.3), in accordance with the following equilibrium reaction:

$$NH_3 + H_2O \leftrightarrow NH_4^+ + OH^-$$

Ammonia nitrogen for the faecal sludge is determined by the methods described in procedures 4500-NH_3 A, 4500-NH_3 B, and 4500-NH_3 C in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". The ammonia nitrogen concentration in the sludge was calculated as follows:

$$\text{mg } NH_3\text{ - N/kg} = \frac{(A - B) \text{ x } 280}{\text{g dry wt sample}}$$

where:

A = volume of H_2SO_4 titrated for sample, ml, and
B = volume of H_2SO_4 titrated for blank, ml.

3.2.3.6 pH

pH is a term used universally to express the intensity of the acid or alkaline condition of a solution. It is a way of expressing the hydrogen-ion concentration in the solution. In wastewater treatment employing biological processes, pH must be controlled within a range favourable to the particular microorganisms involved. The pH of the faecal sludges were measured using a pH meter.

3.3 Phase 2: Anaerobic Digestion of Primary Sludge Using a Laboratory-Scale UASB Reactor in Leeds

The second stage of the research work involved the anaerobic digestion of primary sludge using the UASB reactor. The essence of this stage was to get hands-on experience using the UASB reactor before trying it on faecal sludges. This phase was carried out between mid-May and early September 1999 in Leeds.

3.3.1 Raw Material

The raw material used in this phase was primary sludge from the Owlwood Sewage Treatment Works treating only domestic sewage. This material was chosen to avoid any interference from industrial waste. The sewage works is located near Kippax in West Yorkshire.

The primary sludge was taken from concrete storage tanks (Plate 11) and was normally up to a week old. This was chosen as it was considered to be the material closest to faecal sludge. The primary sludge was then diluted with final effluent from the Knostrop Sewage Treatment Works in Leeds. This was done to enable the pump to pump the waste into the reactor without experiencing any blockages. The initial dilution was 1 in 20 and this was gradually increased to 1 in 10. The dilution ratio was chosen after conducting a quick analysis of the primary sludge to determine its characteristics.

3.3.2 Characterisation of Primary Sludge

The primary sludge used as the raw material was first characterised. It was essential to know the nature of the waste to be treated and also to assist in the design of the laboratory-scale UASB needed for the experiments. The physical and chemical parameters determined were pH, chemical oxygen demand (COD), total solids (TS), total volatile solids (TVS) and ammonia nitrogen (NH_3-N). The analysis for the listed parameters was carried out at the wastewater treatment laboratory of the School of Civil Engineering in the University. The determination of TS and TVS was same described section 3.2.3.3 and 3.2.3.4 respectively. A summary of the characteristics of the primary sludge from the Owlwood Sewage works is provided in Chapter 4.

Plate 11 **Taking sludge samples at Owl Wood Sewage Treatment Works**

3.3.2.1 Chemical Oxygen Demand (COD)

The COD was determined using the dichromate closed reflux method (Plate 12) as described in method 5220 C in the 19th edition of "Standard methods for the Examination of Water and Wastewater". The COD was calculated as in section 3.2.3.2

3.3.2.2 Ammonia Nitrogen (NH_3-N)

The ammonia nitrogen in the primary sludge was determined using the ammonium ion selective electrode (ISE) by Phillip Harris Scientific (1996), a method similar to the ammonia-selective electrode method described in method 4500-NH_3 D in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". In this procedure, 1M lithium acetate solution prepared by dissolving 51 g of lithium acetate dihydrate ($CH_3CO_2Li.2H_2O$) in distilled water and making it up to 500 ml, was used as an Ionic Strength Adjustment Buffer (ISAB).

Plate 12 **COD determination – digestion tubes in a heating block**

A series of standard solutions of ammonium chloride covering concentrations 100, 10 and 1 mg NH_3-N/l were prepared in Nessler tubes by making decimal dilutions of stock 1000 mg NH_3-N/l standard ammonium chloride solution with distilled water. 100 ml of the sample to be measured was placed in an additional Nessler tube. Using a pipette, 2 ml of the ISAB was added to the sample and 1 mg/L standard, and 1.8 ml to the 10 mg/l and 100 mg/l standards and stirred thoroughly. The tubes were then placed in a temperature bath (Plates 13 and 14) to ensure the standards and the sample are at the same temperature and that the temperature remained constant for the testing. The ISE and a thermometer were then immersed into the 1 mg/l standard to record the mV and temperature reading. The procedure was repeated for each standard in increasing concentration, rinsing the electrodes between measurements. A calibration curve of mV versus ammonia nitrogen concentration (mg/l) on linear/log graph paper was plotted. The ISE and thermometer were immersed in the sample and the mV recorded. Using the calibration curve, the ammonia nitrogen concentration was read.

Plate 13 Experimental set-up for the determination of ammonia nitrogen

Plate 14 Standards and sample immersed in water bath

3.3.2.3 pH

A pH meter (Plate 15) was used to measure the pH of the primary sludge.

Plate 15 **pH meter**

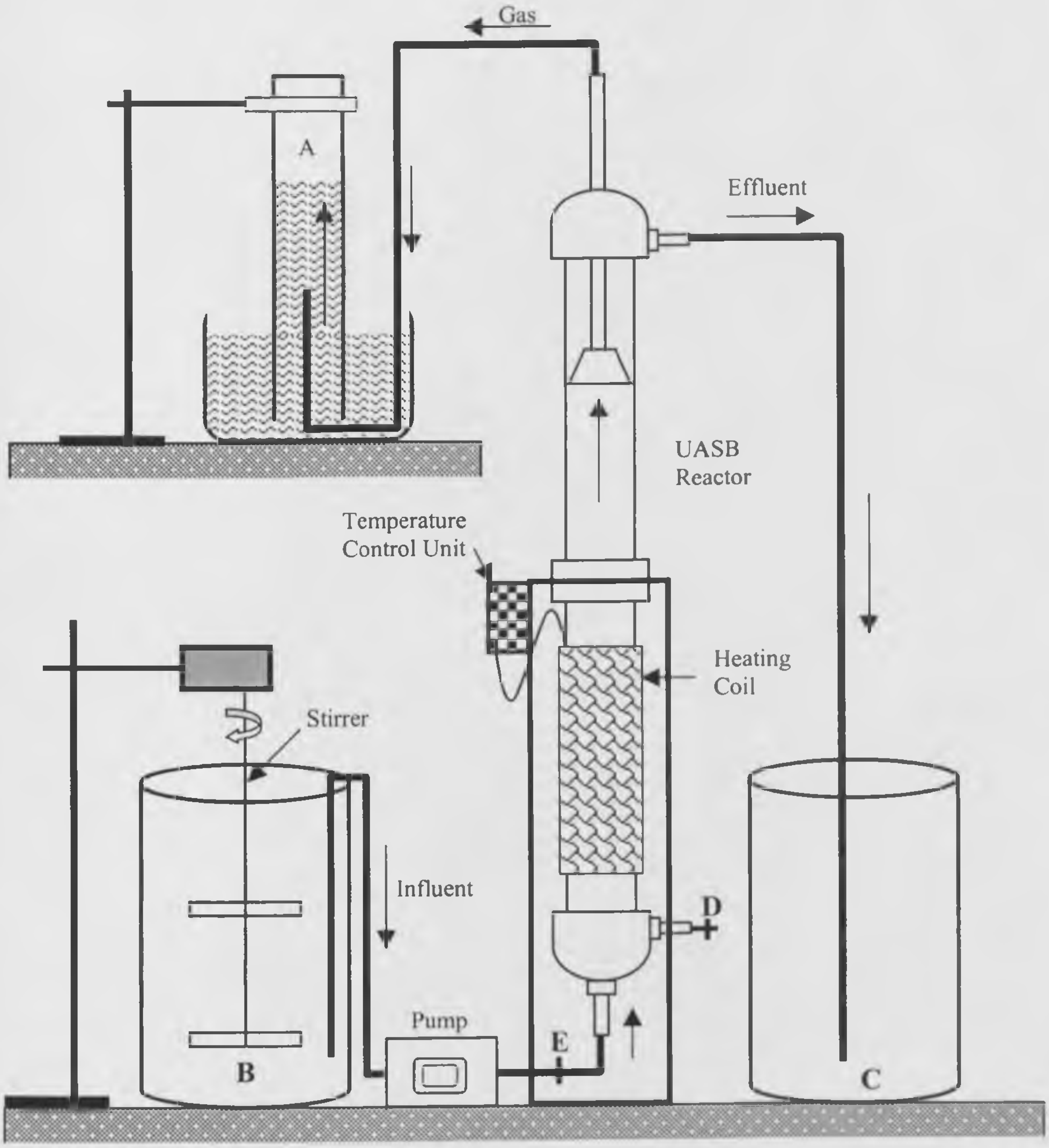
3.3.3 Experimental Set-up

The set-up used for the experiment is shown in Figure 3.1 and Plate 16. The set-up was sited in the wastewater laboratory at the Knostrop Sewage Treatment Works. The set-up consisted of a 15-litre UASB reactor, with a heating coil wrapped around the lower half of the reactor (Plate 17). The heating coil with the temperature control unit was used to maintain a temperature of 37 °C to ensure optimal temperature conditions for the anaerobic digestion taking place in the reactor. A 2-litre measuring cylinder filled with water and inverted in a bucket containing water was used to collect the gas produced as a result of the digestion process (Plate 18).

As shown in Figure 3.1, the primary sludge was placed in a 125-litre barrel and diluted with final effluent from the Knostrop Sewage Treatment Works. The influent was gently stirred to prevent the settling of sludge particles and also keep them well mixed (Plate 19). The stirring was gently done to prevent any turbulence that would have introduced air into the influent waste. The influent was then pumped continuously through the UASB reactor using a peristaltic pump (Plate 20) and the effluent collected in another barrel. The test was run at a pre-determined flowrate after several trials.

Samples for analysis were taken just after the pump for the influent (Plate 21) and at the end of the effluent tubing (Plate 22). The flowrate were also determined at the time of taking the samples by recording the times required to collected a know volume of the sample. The reactor was seeded with sludge from an UASB reactor treating sugar wastes at the British Sugar factory in York. The seeding was done to ensure a quick start-up of the reactor.

Plate 16 **Experimental Set-up for Phase 2**



A - Gas collected from the UASB reactor by downward displacement of water

B – vessel containing influent diluted sludge

C – vessel to collect effluent from reactor

D – Valve for wasting excess sludge

E – Valve for taking influent sample

Figure 3-1 **Schematic diagram of the experimental set-up for the anaerobic digestion of primary sludge**

Plate 17 Heating coil wrapped around lower portion of reactor

Plate 18 Gas collection unit for Phase 2 experiment in Leeds

Plate 19 Stirring the influent to keep sludge particles in suspension and ensure a well mixed influent.

Plate 20 A WATSON MARLOW 313S Peristaltic Pump

Plate 21 Collecting influent sample

Plate 22 Collecting effluent sample

3.3.4 Laboratory Analysis

The influent and effluent samples collected were analysed for the following physical and chemical parameters:

1. Chemical oxygen demand (COD) – total and filtered
2. Total solids (TS)
3. Total volatile solids (TVS)
4. Total Suspended Solids (TSS)
5. Volatile Suspended Solids (VSS)
6. Ammonia nitrogen (NH_3-N)
7. pH

The COD, TS, TVS, TSS and VSS analysis were carried out at the wastewater laboratory at Knostrop Sewage Treatment Works while the ammonia nitrogen and pH determination were carried out at the wastewater treatment laboratory in the School of Civil Engineering at the University. The determination of COD, TS, TVS, ammonia nitrogen and pH are same as described in previous sections. A summary of the results is presented in Chapter 4.

3.3.4.1 Total Suspended Solids (TSS)

The total suspended solids content of the influent and effluent samples for the UASB reactor were determined using the total suspended solids dried at 103 – 105 °C method described in method 2540 D in the 19th edition of "Standard Methods for the Examination of Water and Wastewater".

A well-mixed sample was filtered through a weighed Whatman GF/C filter paper and the residue retained on the filter dried to a constant weight at 103 to 105 °C. The increase in weight of the filter paper represents the total suspended solids. The total suspended solids is calculated as follows:

$$\text{TSS mg/l} = \frac{(\text{A} - \text{B}) \times 1000}{\text{sample volume, ml}}$$

where:

A = weight of filter + dried residue, mg, and

B = weight of filter, mg.

3.3.4.2 Volatile Suspended Solids (VSS)

The weight loss when the total suspended solids is ignited at 550 ± 50 °C is called "volatile suspended solids". The volatile content of the total suspended solids was determined using the same method described for the determination of total volatile solids in section 3.2.3.4. The volatile suspended solids is calculated as follows:

$$\text{VSS mg/l} = \frac{(\text{A} - \text{B}) \times 1000}{\text{sample volume, ml}}$$

where:

A = weight of filter + dried residue before ignition, mg, and

B = weight of filter + residue after ignition, mg.

3.4 Anaerobic digestion of faecal sludge using a laboratory-scale UASB in Kumasi, Ghana

The third stage of the research work involved the anaerobic digestion of faecal sludges using the UASB reactor. Following the hands on experience acquired under Phase 2, a larger reactor of capacity 50 litres was designed and used for the third phase. The third phase was undertaken in Kumasi, Ghana between early February and the end of May 2000.

3.4.1 Raw Material

The raw material used in this third phase of the research work was faecal sludges. The faecal sludge was collected directly from tanker trucks discharging their waste (Plate 23) into 10 litre containers (Plate 24) and transported to the experimental site. The faecal sludge in this state contains a lot of grit, large particles, plastics, pieces of wood, carrier bags etc (Plate 25) and is also concentrated. It is thus diluted using tap water (Plates 26) and then filtered using a sieve with holes dimension of about 3mm x 3mm to get rid of all the large solid particles that will cause a blockage of the pump (Plate 27). The dilution ratio used ranged from 1 in 6 to 1in 8 depending on the

Plate 23 Collecting faecal sludge sample from discharging tanker truck

Plate 24 Filling 10-litre container with faecal sludge

Plate 25 Faecal sludge discharged site showing kinds of solid materials that may be present in the sludge

Plate 26 Diluting faecal sludge with tap water

Plate 27 **Sieving diluted faecal sludge**

concentration of the sample collected from the tanker truck and also the organic loading that was to be applied. The diluted and filtered faecal sludge then served as the influent into the UASB reactor.

3.4.2 Sampling

As mentioned in section 3.2.2 it is crucial to obtain as closely as possible a representative sample when collecting the faecal sludge from the discharging tanker truck given the very nature of faecal sludge and mode in which it is collected. Using four 10-litre plastic bottles, grab samples were taken at the start of the discharge, midway through the discharge and towards the end of the discharge from the tanker trucks. The collected samples were then prepared as described in the previous section to be used as the influent for the UASB reactor.

3.4.3 Experimental Set-up

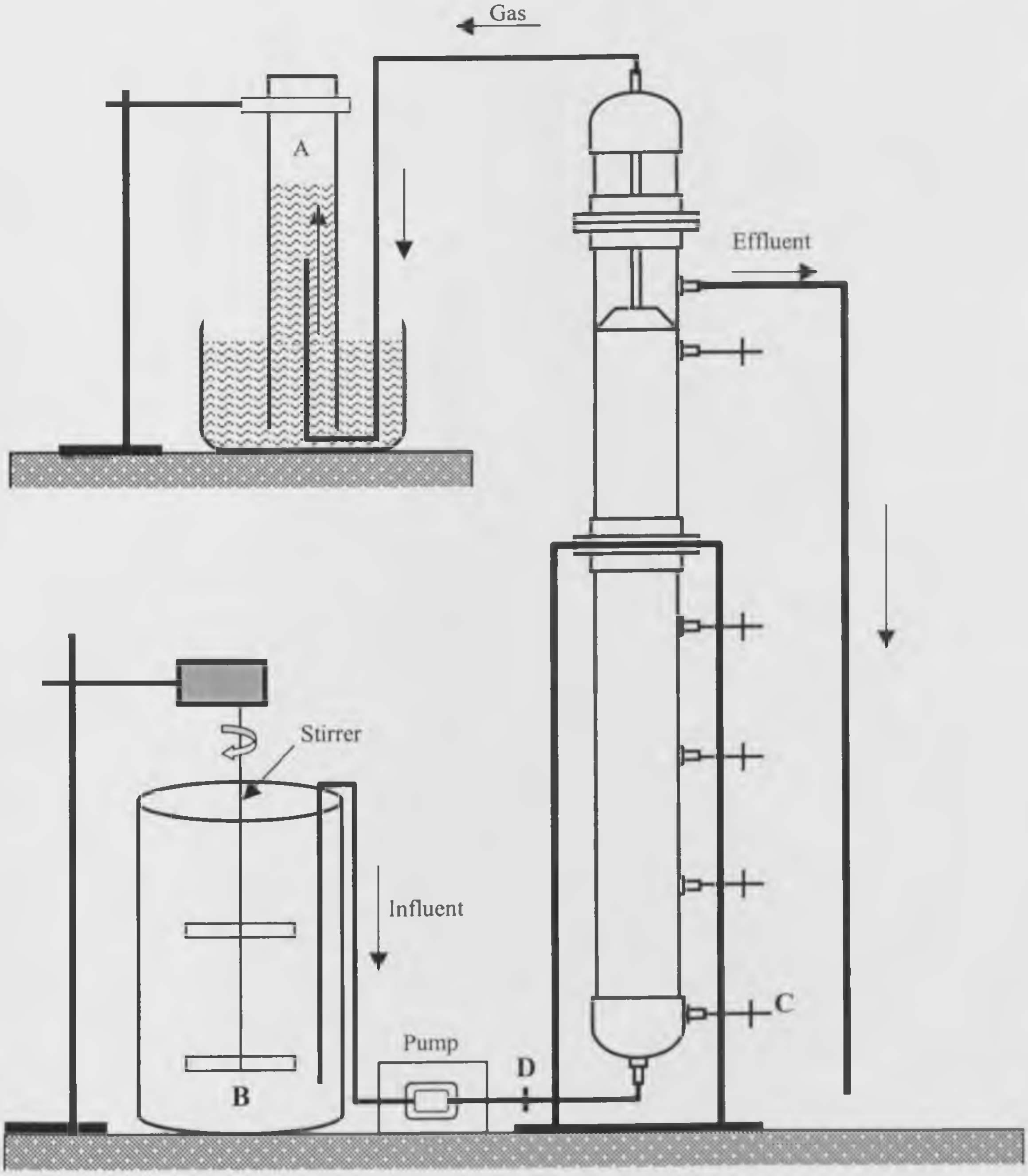
The set-up used for the experiment is shown in Figure 3.2 and Plate 28. For security reasons, reliable electricity supply and ease of accessibility, the experimental set-up was sited in an open space behind the offices but within the compound of the Training Research and Networking for Development (TREND) Group, in Kumasi.

The set-up consisted of a 50-litre UASB reactor (without a heating coil). A 2-litre measuring cylinder filled with water and inverted in a bucket containing water was used to collect the gas produced as a result of the digestion process (Plate 29).

The prepared influent faecal sludge was poured into a 250-litre barrel (Plate 30) and gently stirred to prevent the settling of sludge particles and also keep them well mixed. The stirring was gently done to prevent any turbulence that would have introduced air into the influent waste. The influent was then pumped continuously through the UASB reactor using a peristaltic pump (Plate 31) and the effluent allowed to flow into a pit filled with stones (Plate 32).

Based on the results obtained in the Leeds tests, a flowrate that allowed 12 hours retention time was maintained throughout the experiment. The reactor was not seeded at start-up. Samples for analysis were taken just after the pump for the influent and at the end of the effluent tubing.

Plate 28 **Experimental Set-up for the anaerobic digestion of faecal sludge using the UASB**



A - Gas collected from the UASB reactor by downward displacement of water

B – vessel containing influent sample of sieved and diluted faecal sludge

C – Valve for wasting excess sludge

D – Valve for sampling influent

Figure 3-2 **Schematic diagram of experimental set-up for the anaerobic digestion of faecal sludge**

Plate 29 Gas collection unit for Phase 3 experiment in Kumasi

Plate 30 Filling the 250-litre barrel with the influent

Plate 31 A WATSON MARLOW 505S Peristaltic Pump used in pumping the influent

Plate 32 Pit filled with stones – receives effluent from reactor

3.4.4 Laboratory Analysis

The influent and effluent samples collected were analysed for the following physical and chemical parameters:

1. Biochemical oxygen demand (BOD)
2. Chemical oxygen demand (COD)
3. Total solids (TS)
4. Total volatile solids (TVS)
5. Total Suspended Solids (TSS)
6. Total Kjeldahl nitrogen (TKN)

All the analysis were carried out at the Environmental Quality Engineering laboratory located within the Department of Civil Engineering at the Kwame Nkrumah University of Science and Technology, Kumasi. The determination of BOD, COD, TS, TVS, and TSS are same as described in previous sections. With regards to BOD, not all samples were examined because the laboratory was closed over the weekends and on public holidays. A summary of the results is presented in Chapter 4.

3.4.4.1 Total Kjeldahl Nitrogen (TKN)

Total nitrogen is comprised of organic nitrogen, ammonia, nitrite and nitrate. Total Kjeldahl nitrogen refers to the total of organic and ammonia nitrogen. The TKN was determined by the semi-micro-kjeldahl method as described in method 4500-N_{org} C in the 19th edition of "Standard Methods for the Examination of Water and Wastewater". The samples were first boiled to drive off the ammonia present in solution and then digested. Thus values obtained represent the organic nitrogen concentrations in the samples.

The principle of the method is the conversion of organic nitrogen into ammonium in the presence of concentrated sulphuric acid (H_2SO_4), potassium sulphate (K_2SO_4), and mercuric oxide as catalyst. Free ammonia, if not removed prior to the digestion, is also converted to ammonium. After the digestion, a base is added to release ammonia from the ammonium complex formed during the digestion. The ammonia is then distilled from the alkaline medium and absorbed in boric acid. Using the titrimetric method described in method 4500-NH_3 C in the 19th edition of "Standard Methods for the Examination of Water and Wastewater", the ammonia absorbed by the boric acid is determined.

The ammonia nitrogen concentration in the analysed samples were calculated as follows:

$$\text{mg NH}_3\text{ - N/kg} = \frac{(A\ -\ B)\ x\ 280}{\text{g dry wt sample}}$$

where:

A = volume of H_2SO_4 titrated for sample, ml, and

B = volume of H_2SO_4 titrated for blank, ml.

Chapter Four

4 Results

4.1 Introduction

This chapter presents the results of the various tests carried out in the previous chapter. For ease of reading and referencing, the presentation of the results follows the same order as described in Chapter 3.

4.2 Results for tests carried out under Phase 1

Table 4-1 Characteristics of nightsoil/toilet sludge in Kumasi, Ghana

Parameter	Value	
	Average	Range
pH	-	8.1 - 8.5
BOD (mg/l)	23,300	14,200 - 52,000
COD (mg/l)	86,700	36,600 - 175,000
COD/BOD	4:1	2.4:1 - 7.2:1
TS (mg/l)	55,700	31,300 - 87,000
TVS (mg/l)	39,700	15,000 - 65,400
TVS/TS	71%	48% - 76%
TKN (mg/l)	2,400	700 - 4050

(Analysis of 8 samples taken between 14/5/98 - 26/6/98)

The values for BOD, COD, TS, TVS and ammonia nitrogen that can be expected for nightsoil and toilet sludge were very high (Table 4-1). Apart from being high, the values were also quite variable with a large range between the minimum and maximum. The standard deviations for BOD, COD, TS, TVS and ammonia nitrogen were respectively 12,900 mg/l, 44,100 mg/l, 15,700 mg/l, 14,200 mg/l and 1,090 mg/l. This stems from the fact the faecal sludges collected by the vacuum tankers were of variable age and as a result had undergone varying degrees of biodegradation. Faecal sludge may be relative fresh. In the case of bucket latrines, that is up to a few days old. Others could be a few months to a couple of years old as was the case for toilet sludge from "dry" aqua privies and KVIPs, and septage from septic tanks.

The BOD of the faecal sludge ranged from 14,200 to 52,000 mg/l with a mean value of 23,300 mg/l. The higher BOD values were for the fresher faecal sludges such as nightsoil, which contained a higher biodegradable organic content than the toilet sludge and septage. The COD ranged from 36,600 to 175,000 mg/l with a mean value of 86,700 mg/l. The ratio of COD/BOD ranged from 2.4:1 to 7.2:1 an average value of 4:1. The lower the value, the higher the biodegradable organic content in the faecal sludge.

Total solids and total volatile solids concentrations were also very high. The total solids concentration varied between 31,300 and 87,000 mg/l with a mean value of 55,700 mg/l while the total volatile solids ranged from 15,000 to 65,400 mg/l with an average value of 39,700 mg/l. The ratio of TVS/TS had a minimum value of 48% and a maximum value of 76%, i.e. the volatile organic content varied from nearly half to just over three-quarters of the total solids. This high volatile fraction makes it quite attractive for anaerobic treatment.

The ammonia-nitrogen content ranged from 700 to 4050 mg/l, with a mean value of 2,400 mg/l. Higher values were recorded for nightsoil and toilet sludge while values for septage were lower. This is because septage was diluted with flush water in addition to a higher biodegradation compared with toilet sludge and nightsoil. The high values could be critical if faecal sludge is to be treated anaerobically. It will be essential to reduce the concentration, possibly by dilution before treatment.

4.3 Results for tests carried out under Phase 2

As described in section 3.3.3, a 15-litre UASB reactor was used for anaerobic digestion of the primary sludge. Throughout the experiment, the heating coil wrapped around the lower portion of the reactor was constantly on to maintain a temperature of 37 °C, which is considered optimal for mesophilic anaerobic digestion. The retention time in the reactor ranged from 7.8 to 12.4 hours with a mean value of 9.8 hours (mean upflow velocity 0.18 m/h). The variation was due primarily to the manual setting of the pump to achieve the design HRT of 10 h and also flushing the tubing to prevent blockage. It was noticed during the experiment that solids build up in the tube reducing the effective diameter of the tubing, which resulted in decreased flow, and thus increasing the retention time.

4.3.1 Characteristics of primary sludge

Table 4-2 Characteristics of Primary Sludge from Owlwood Sewage Works

Parameter	Value	
	Average	Range
PH	-	5.28 - 5.36
COD (mg/l)	41,940	40,840 - 43,230
TS (mg/l)	36,500	35,800 - 37,130
TVS (mg/l)	29,000	28,100 - 29,680
TVS/TS	79.6%	78.4% - 80.2%
NH_3-N (mg/l)	337	324 - 350

(Analysis of 4 samples taken between 23/11/98 - 27/11/98)

The COD, TS, TVS and ammonia values obtained in the Owlwood primary sludge were high (Table 4-2), but lower than the values obtained for faecal sludge and the range of the values were not as varied as that for the faecal sludge. The low variability of the parameters compared with those of faecal sludge was because the primary sludge came from the same source while the faecal sludge was from different sources.

The ratio of TVS/TS show very high values, ranging from 78.4% to 80.2% with an average value of 79.6%. This shows a high percentage of organic content in the total solids which is good if the primary sludge is to be treated using an anaerobic treatment process. The pH values were rather low (acidic conditions), possibly indicating that the first stage of anaerobic digestion which involves the hydrolysis of proteins into amino acids and lipids into fatty acids might have taken place. At the first stage of anaerobic digestion, ammonia is also produced during the hydrolysis of proteins and hence could be responsible for the high concentration of ammonia in the primary sludge.

4.3.2 Chemical Oxygen Demand (COD)

COD was measured on samples from the influent into the reactor; the effluent from the reactor; and the filtered effluent from the reactor (filtrate from the determination of effluent suspended solids) (Figure 4-1 and Table 4-3). Also shown on the graph are the percentage removal efficiencies for the total COD and filtered COD.

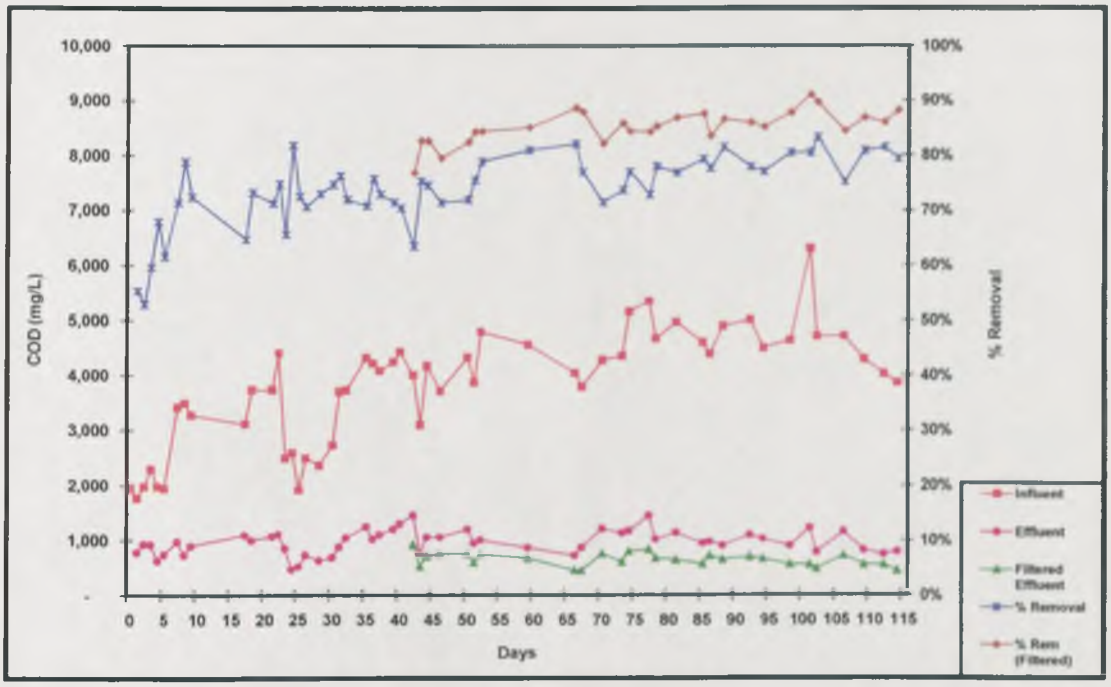


Figure 4-1 COD values for the influent/effluent and removal efficiency with time for primary sludge

Table 4-3 Summary of COD values for influent/effluent samples from the UASB reactor using primary sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	6310	1460	84
Minimum	1770	470	53
Mean	3800	960	74
Standard Deviation	1020	210	7

The influent COD concentration ranged from 1770 to 6310 mg/l with a mean value of 3800 mg/l and a standard deviation of 1020 mg/l, while the effluent concentration (total) ranged from 470 to 1460 mg/l with an average value of 960 mg/l and a standard deviation of 210 mg/l. With regards to the filtered effluent, the COD concentration ranged from 450 to 920 mg/l with an average value of 650 mg/l and a standard deviation of 120 mg/l. COD removal efficiencies were very good. The total COD removal was never less than 53%; was as high as 84%; and it averaged 74% with a standard deviation of 7%. The removal efficiencies for filtered effluent was even better, ranging between 77% and 91% and a mean value of 85% with a standard deviation of 3%. Removal of suspended solids thus improved the COD removal efficiency by 7% to 24%. A considerable portion of the total COD was therefore due to the soluble component.

It is apparent from Figure 4-1 that the influent COD increased especially in the first half of the experimental period. This apparent upward trend in the influent COD concentration was the result of gradually decreasing the dilution ratio used from 1:20 to 1:10. The effluent COD values, although varied, the degree of variation was much less (effluent COD standard deviation was 210 mg/l) than that of the influent especially (influent COD standard deviation was 1020 mg/l) in the last third of the graph. The variation was even less with the filtered effluent (filtered effluent COD standard deviation was 120 mg/l). This could be an indication of steady-state conditions developing. The removal efficiencies for both total and filtered COD showed an upward trend as the experiment progressed, with the filtered value always being higher than the total value.

4.3.3 Total Solids (TS)

Total solids (TS) was measured for the influent into the reactor and the effluent from the reactor. The values obtained were used to calculate the percent removal efficiency (Figure 4-2 and Table 4-4).

The total solids concentration in influent ranged from 2270 to 6520 mg/l with a mean value of 3690 mg/l whilst that of the effluent ranged from 980 to 1970 mg/l with an average value of 1350 mg/l. There was good solids removal with the removal

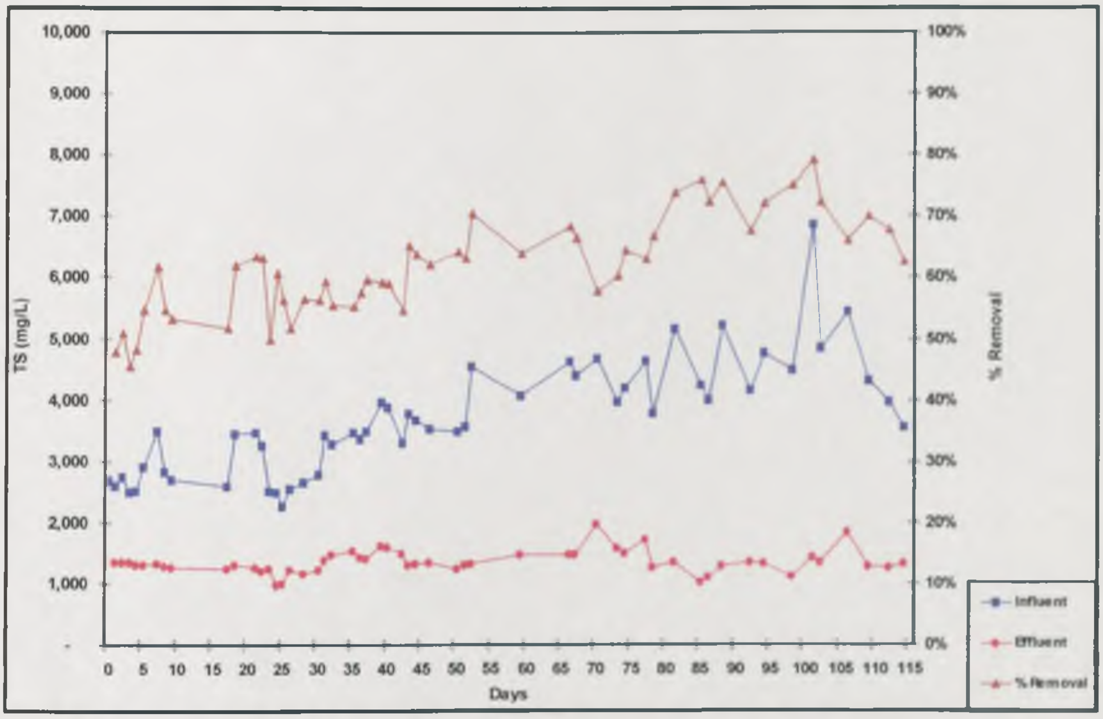


Figure 4-2 TS values for influent/effluent and removal efficiency with time for primary sludge

Table 4-4 Summary of TS values for influent/effluent samples from the UASB reactor using primary sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	6520	1970	78
Minimum	2270	980	46
Mean	3690	1350	62
Standard Deviation	900	180	8

efficiency never below 46%. It was as high as 78% in an instance and it averaged 62% with a standard deviation of 8%. The graph shows that there was a general upward trend in the total solids removal as the experiment progressed. This could be due to the formation of the sludge blanket which tend trap more solids from the influent as it increases in density.

It is apparent from the graph (Figure 4.2) that the influent total solids concentration had an upward trend, especially in the first portion of the experimental period. Again this was the result of gradual reduction of the dilution ratio from 1:20 to 1:10 as the experiment progressed. The graph also shows that the effluent concentration was quite constant although the influent varied a lot (the standard deviation for the effluent TS was 180 mg/l compared to 900 mg/l for the influent TS). The erratic nature of the influent concentration may be due to the fact that grab samples of the primary sludge were used.

4.3.4 Total Volatile Solids (TVS)

The total volatile solids content was measured for the influent into the reactor and effluent from the reactor to determine the organic fraction of total solids (Figure 4-3 and Table 4-5).

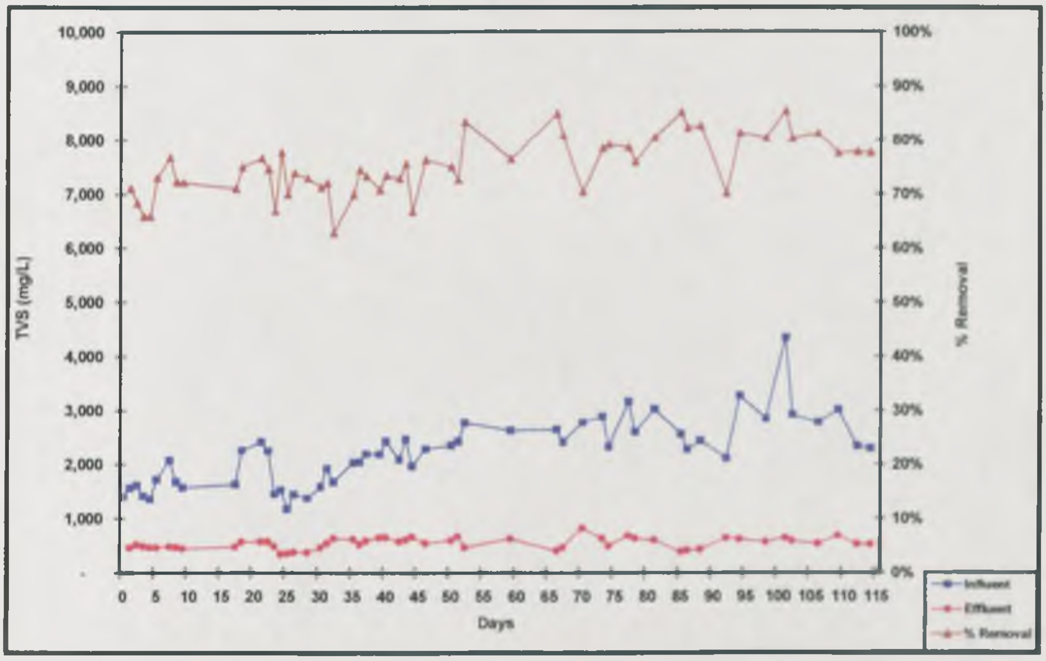


Figure 4-3 **TVS values for influent/effluent and % removal efficiency with time for primary sludge.**

The influent concentration of total volatile solids ranged between 1170 and 4350 mg/l. It averaged 2220 mg/l with a standard deviation of 600 mg/l. With regards to the effluent, the total volatile solids ranged from 340 to 820 mg/l and had a mean value of 530 mg/l with a standard deviation 100 mg/l. As evident from the graph, the effluent concentration varied very little compared with the influent, which also show an upward trend from the start of the experiment. As explained in the earlier sections, this upward trend in the influent concentration was the result of decreasing the dilution ratio used for the raw primary sludge.

Table 4-5 Summary of TVS values for influent/effluent samples from the UASB reactor using primary sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	4350	820	85
Minimum	1170	340	63
Mean	2220	530	75
Standard Deviation	600	100	5

Removal efficiency for total volatile solids was very good, ranging between 63% and 85%. The removal efficiency had a mean of 75% and a standard deviation of 5%. As the experiment progressed, the removal efficiency also went up. Most of the above average removal efficiency was recorded in the latter half of the experiment. The high removal rate indicating the high conversion of organic matter during the anaerobic digestion in the reactor.

4.3.5 Ratio of total volatile solids (TVS) to total solids (TS)

The ratio of TVS to TS was determined using the values obtained from the determination of TVS and TS (Figure 4.4).

The ratio of TVS/TS for the influent ranged from 47 to 72% with a mean of 60% and a standard deviation of 6% while that for the effluent ranged from 27 to 52% with an average of 39% and a standard deviation of 6%. As shown on the graph, there is no apparent trend except that at any instance, the ratio for the effluent is less than that for the influent. The decrease in the ratio could be seen as the extent of utilisation of organic fraction in the total solids. The decrease in ratio was never less than 4%. At times it was as high as 33% and it average 21%.

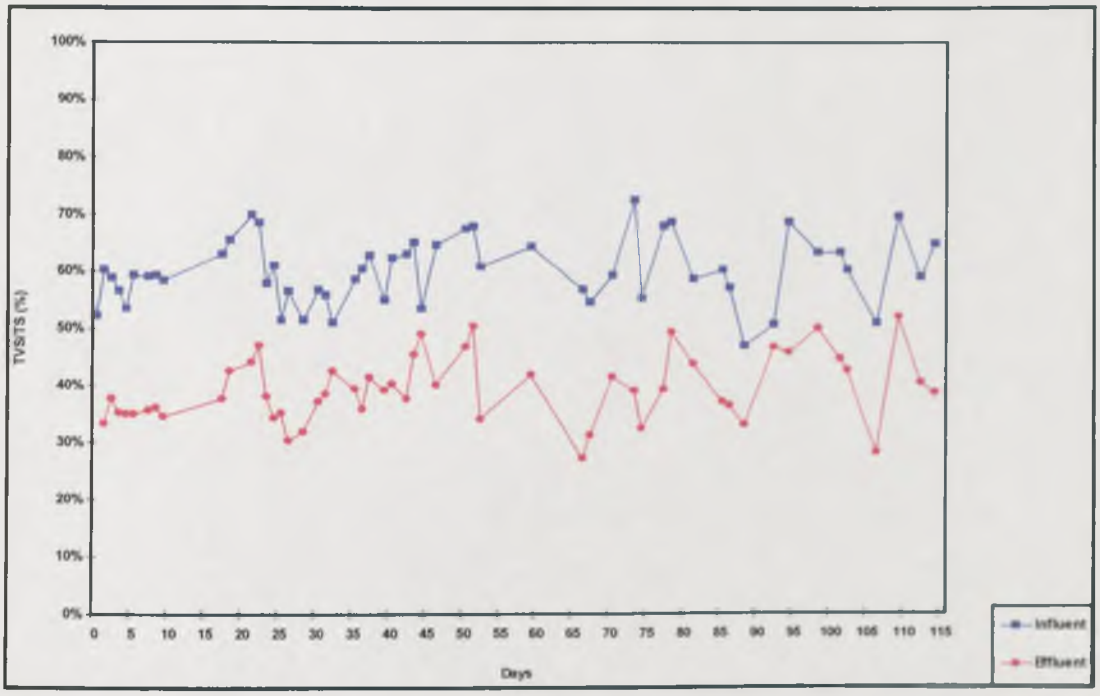


Figure 4-4 Ratio of TVS/TS for influent/effluent with time for primary sludge.

4.3.6 Total Suspended Solids (TSS)

Figure 4-5 and Table 4-6 show the results obtained in the determination of the total suspended solids content in the influent into the reactor and the effluent from the reactor. Also shown on the graph and in the table are the percent removal efficiency.

The total suspended solids ranged from 1500 to 5600 mg/l for the influent and 130 to 580 mg/l for the effluent. The respective mean concentrations and standard deviations for the influent and effluent are 2900 mg/l and 820 mg/l, and 250 mg/l and 100 mg/l. Suspended solids removal was never less than 80%. It was as high as 96% and had an average value of 91% with a standard deviation of 4%.

It is apparent from the graph that there was an upward trend in the influent concentration. This was primarily due to the decrease in the dilution ratio from 1:20 to 1:10 as the experiment progressed. The gradual decrease was to enable the seed bacteria to adapt to the increasing load. Furthermore the graph shows that the degree of variation in the effluent was less than that for the influent. As mentioned earlier,

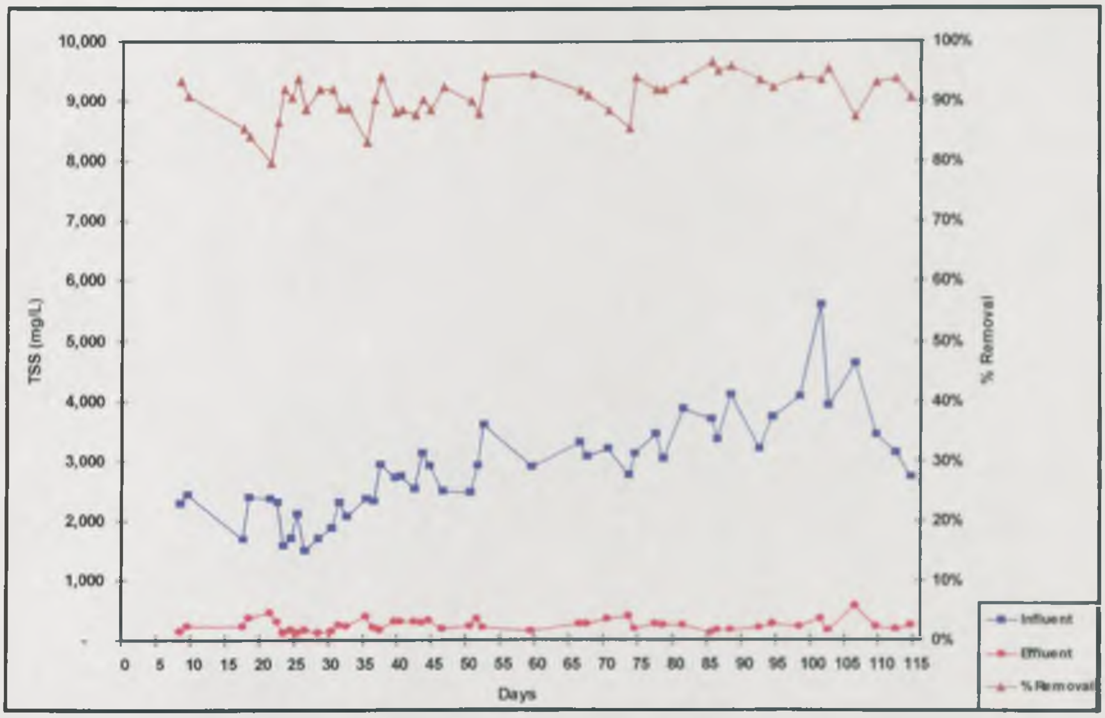


Figure 4-5 TSS values for influent/effluent and % removal efficiency with time for primary sludge.

Table 4-6 Summary of TSS values for influent/effluent samples from the UASB reactor using primary sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	5600	580	96
Minimum	1500	130	80
Mean	2900	250	91
Standard Deviation	820	100	4

the higher degree of variation in the influent concentration could be attributed to the mode of collecting primary sludge samples. The gap at the start of the graph was due to the fact that the determination of total suspended solids started later and not at the very beginning of the study.

4.3.7 Ratio of total suspended solids (TSS) to total solids (TS)

The ratio of TSS to TS was determined using the values obtained from the determination of TSS and TS (Figure 4.6). The ratio of TSS/TS for the influent ranged from 59% to 93% with a mean of 75% while that for the effluent ranged from 10% to 38%% with an average of 19%. There is no apparent trend except that at any instance, the ratio for the effluent is less than that for the influent. The difference between the two ratios was never less than 31% while at times it was as high as 80% and it averaged 21%. The low TSS/TS ratio for the effluent indicates a higher proportion of total suspended solids being removed out of the total solids. This could be attributed to good settling and digestion of the organic fraction of the TSS within the UASB reactor.

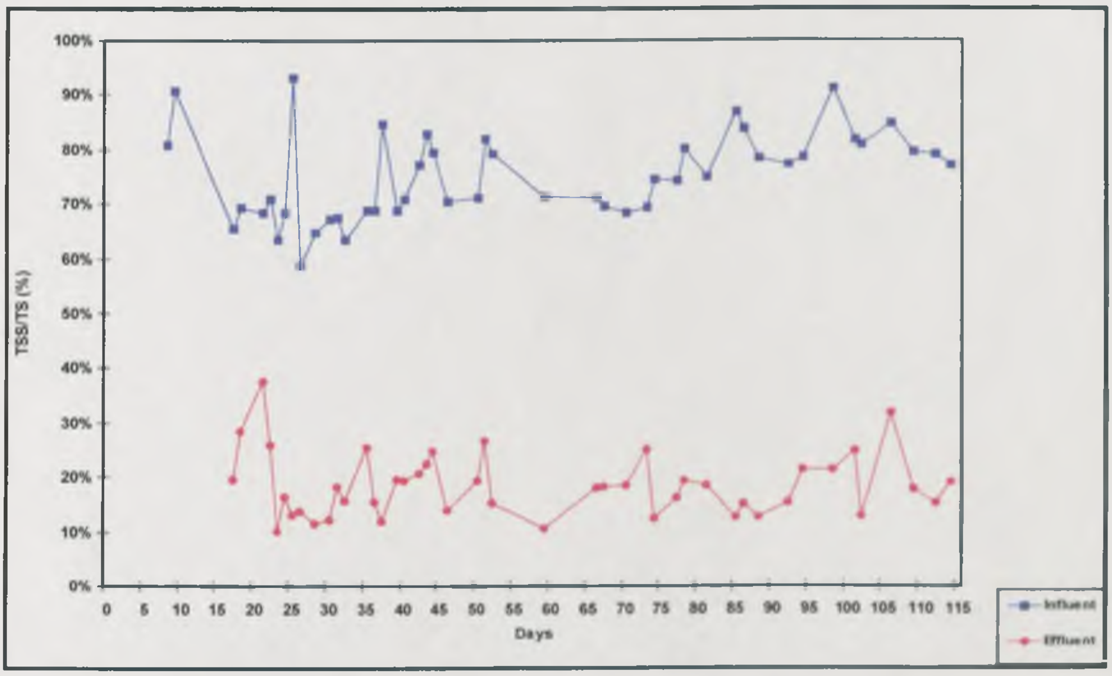


Figure 4-6 Ratio of TSS/TS for influent/effluent with time for primary sludge.

4.3.8 Removal efficiencies for TS and TSS

The graph below (Figure 4.7) shows the removal efficiencies for total solids and total suspended solids over the experimental period. Total solids removal ranged from 46% to 78% with a mean of 62% while removal of total suspended solids ranged from 80% to 96% and had an average of 91%.. As seen from the graph, the removal efficiency for total suspended solids was higher than that of the total solids at any time. The removal of the total solids showed an upward trend with the progress of the experiment. Furthermore, the total suspended solids removal in the last third look fairly stable without much variations.

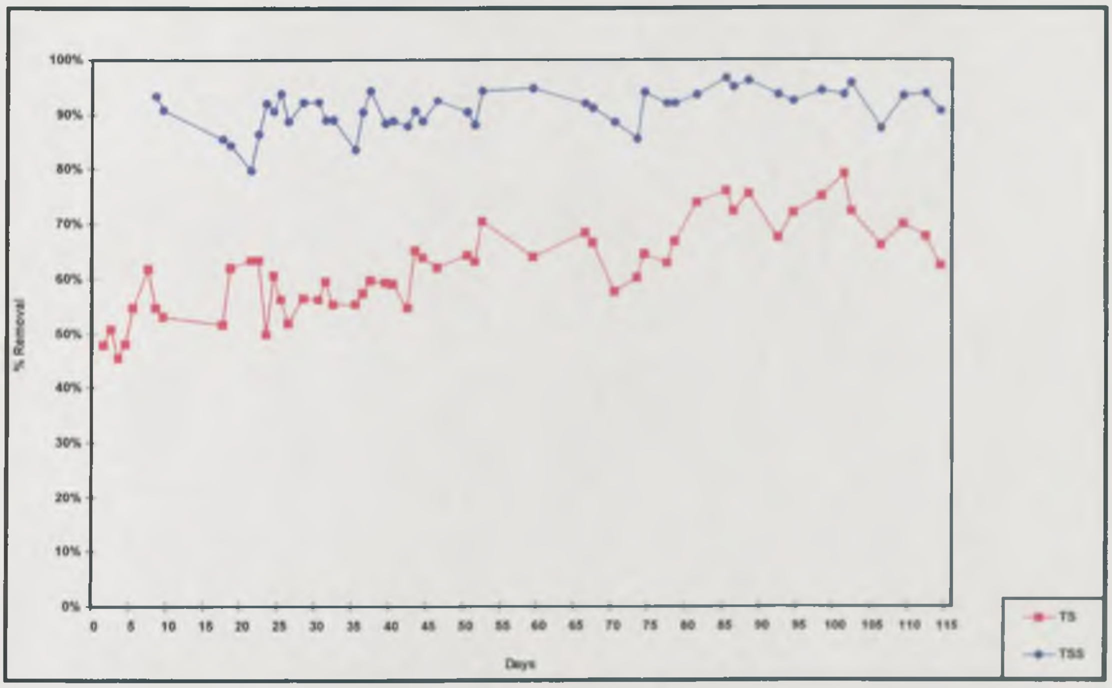


Figure 4-7 Percent removal efficiencies for TS and TSS with time for primary sludge.

4.3.9 Volatile suspended solids

The volatile solids content, a measure of the organic fraction of the total suspended solids, was measured for the influent into the reactor and effluent from the reactor (Figure 4.8 and Table 4-7).

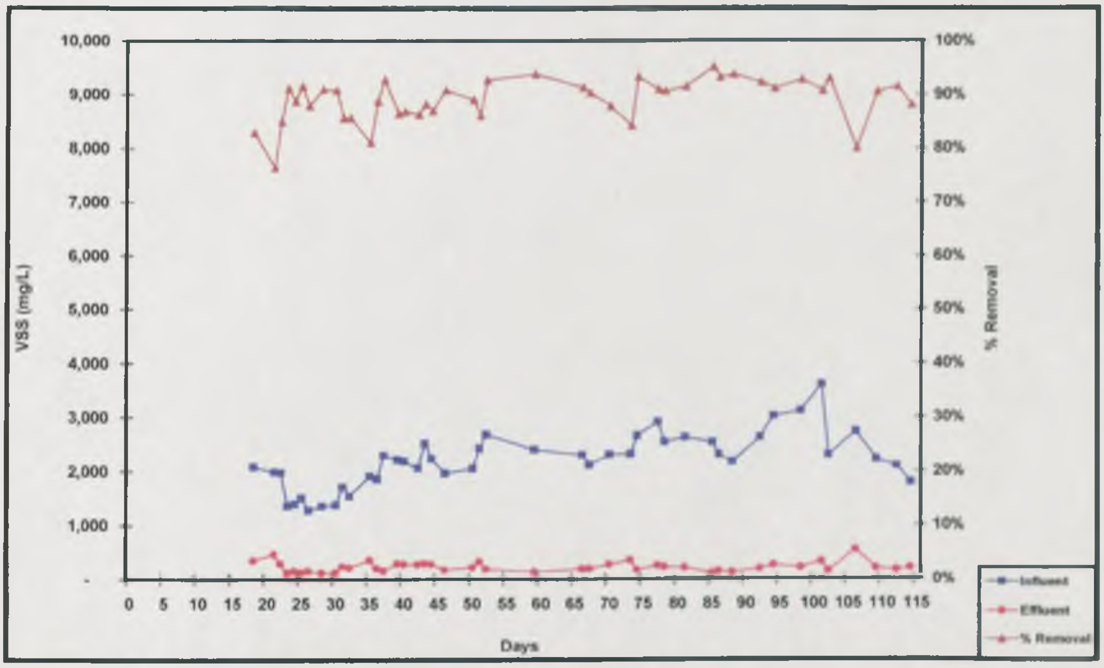


Figure 4-8 VSS values in influent/effluent and % removal with time for primary sludge.

Table 4-7 Summary of VSS values for influent/effluent samples from the UASB reactor using primary sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	3600	540	95
Minimum	1290	120	89
Mean	2200	230	77
Standard Deviation	500	90	4

The gap at the beginning of graph is due to the fact that the measurement of volatile suspended solids did not start at the outset of the study. Influent volatile suspended solids concentration varied from 1290 to 3600 mg/l with a mean value of 2200 mg/l and a standard deviation of 500 mg/l. As evident from Figure 4.8, the influent concentration showed an upward trend especially in the first third of the graph. This was due to the gradual decrease in the dilution ratio from 1:20 to 1:10.

Effluent volatile suspended solids concentration ranged between 120 to 540 mg/l and had an average of 230 mg/l with a standard deviation of 90 mg/l. The graph shows that for most of the time the effluent concentration varied very little although the same cannot be said about the influent concentration. Volatile suspended solids removal was very good and was never below 77%. It was at times as high as 95% and it averaged 89% with a standard deviation of 4%. This high removal rate further underlines the efficiency of the UASB reactor in the treatment of primary sludge. With the exception of a sudden dip in the removal efficiency towards the end of the study, the removal efficiency in the latter half of the study was fairly constant at about 90%.

4.3.10 Ratio of volatile suspended solids (VSS) to total suspended solids (TSS)

Using the results obtained from the determination of VSS and TSS, the ratio of VSS to TSS was calculated (Figure 4.9). The ratio shows the percentage of total suspended solids that may be organic in nature. The ratio of VSS/TSS for the influent ranged from 53% to 87% and averaged 76% while that of the effluent varied between 74% to 99% with a mean of 92%. As shown on the graph, at any instant, the effluent ratio was higher than the influent ratio. This could be due to the higher removal rate of total suspended solids compared to the removal rate of volatile content. The graph shows an apparent decreasing trend in the influent VSS/TSS ratio. This may be due to an increase in the non-volatile content of the total suspended solids, which could also account for the higher removal rate associated with the total suspended solids. The effluent ratio showed very little variations compared to the influent ratio, except the sudden decrease midway through the experiment.

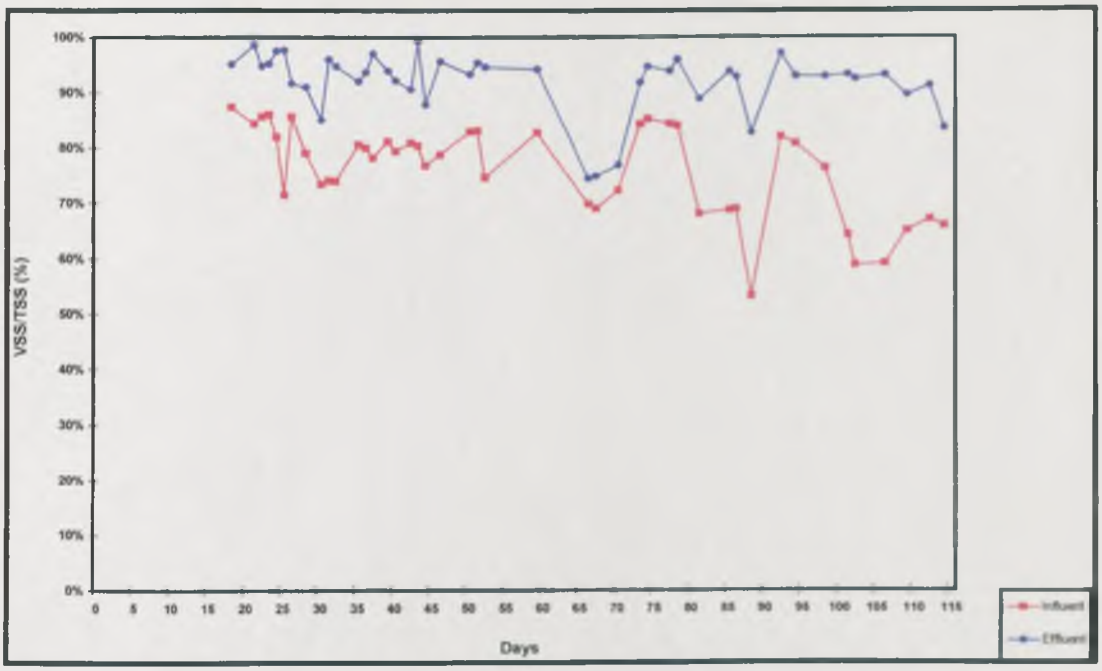


Figure 4-9 Ratio of VSS/TSS in influent/effluent with time for primary sludge.

4.3.11 Ammonia

The concentrations of free ammonia measured in the influent and effluent samples are presented below (Figure 4-10 and Table 4-8). The break in the graph is due to the fact that the determination could not be carried out during the month of June due to unavailability of the equipment used for the measurement.

Table 4-8 Summary of Ammonia values for influent/effluent samples from the UASB reactor using primary sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Rise
Maximum	54	69	49
Minimum	30	37	22
Mean	41	53	30
Standard Deviation	6	8	6

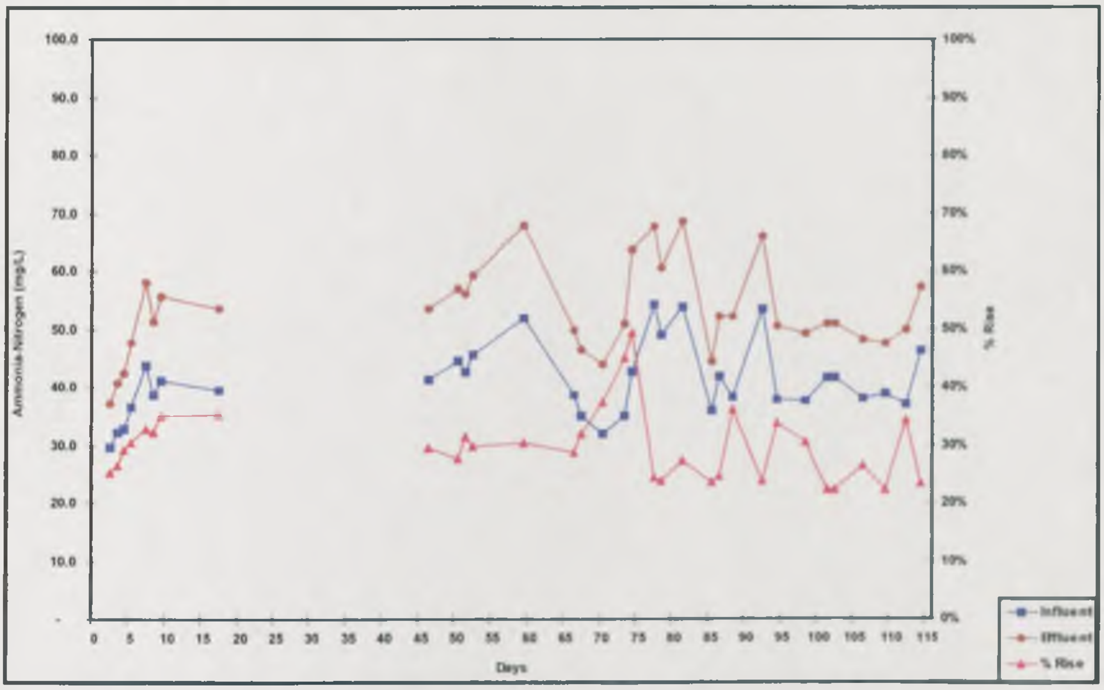


Figure 4-10 Graph showing NH_3-N values in influent/effluent with time for primary sludge.

The influent ammonia concentration ranged from 30 to 54 mg/l with a mean value of 41 mg/l and a standard deviation of 6 mg/l. With regards the effluent, the minimum concentration was 37 mg/l while the maximum was 69 mg/l. The effluent had an average concentration of 53 mg/l and a standard deviation of 8 mg/l. There was an increase in the concentration of ammonia, which is expected in the anaerobic digestion process (Figure 2.11). The percentage rise in the concentration ranged from 21% to 49% and averaged 30%. The initial upward trend before the gap was the result of decreasing the dilution ratio. The concentrations fluctuated except towards the end of the study when it tended to be fairly constant.

The rise in the concentration of ammonia is expected because during anaerobic digestion organic nitrogen is converted into ammonia as nitrogen bacteria consume soluble organic matter containing nitrogen. In primary sludge of domestic origin, there is abundance of nutrients and as such the ammonia produced tends to accumulate in the medium.

4.3.12 Hydraulic retention time (HRT) and percent removals for COD, TS, and TSS

The graph below (Figure 4.11) shows the hydraulic retention times (HRT) over the experimental period and removal efficiencies for COD, TS and TSS.

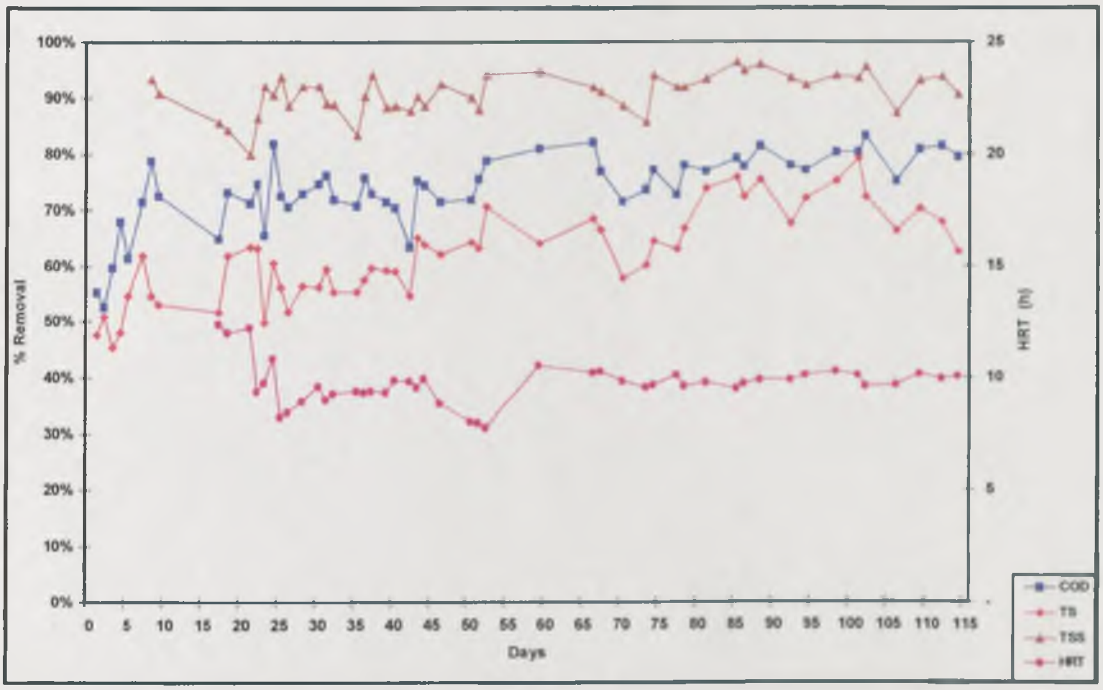


Figure 4-11 Removal efficiencies for COD, TS & TSS and the variation of HRT with time for primary sludge.

The hydraulic retention time (HRT) ranged between 7.8 to 12.4 hours and had an average value of 9.8 hours. The wide variations in some instances as shown in the graph was mainly due to the incorrect setting of the pump after clearing the tubes of grit which will otherwise block the tubes. It was the objective of the study to carry out the study with a retention time of 9 to 12 hours. The graph shows that the retention time in the latter half of the study was fairly constant and close to 10 hours. During this period the COD and total solids removals had a slightly upward trend while the total suspended solids removal was also fairly constant.

4.3.13 Organic loading rate (OLR) and percent removal for COD.

The organic loading rates were calculated using the influent COD values and the determined hydraulic retention times. Figure 4.12 shows the applied organic loading rate and the percent removal efficiency for COD.

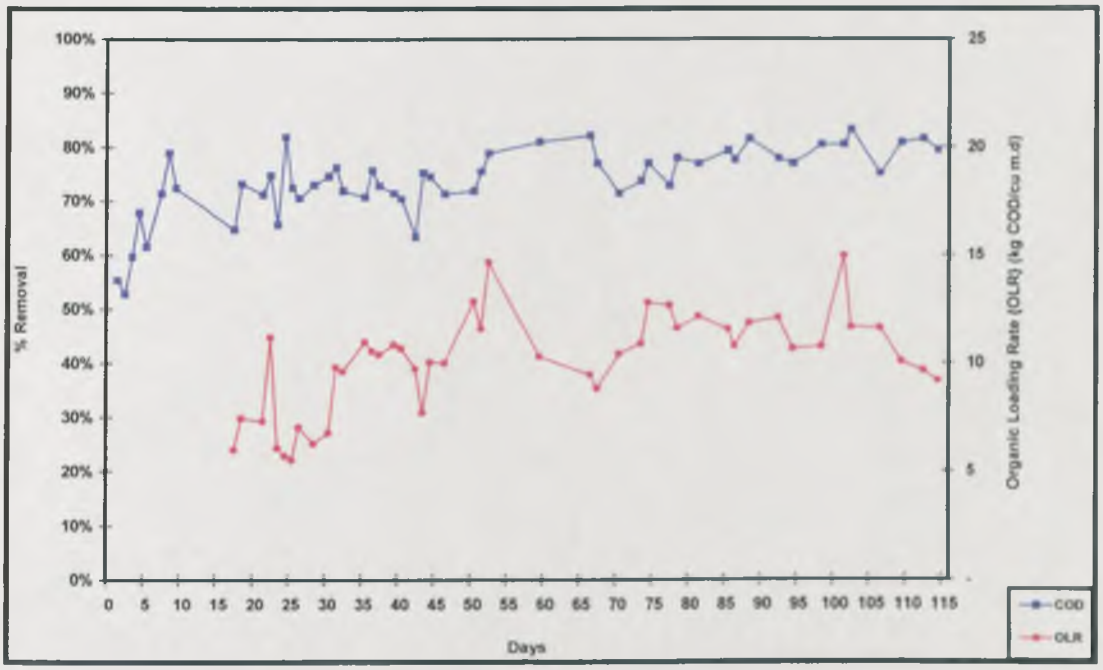


Figure 4-12 Removal efficiency for COD and OLR with time for primary sludge.

The applied organic loading ranged from 5.6 to 15.0 kg COD/m^3.d and it averaged 10.0 kg COD/m^3.d. The organic loading rate was influenced both by the influent COD concentration and the hydraulic retention time especially in the first half of the study. However, in the second half, the influence could be attributed to the variations in the influent COD concentrations because the hydraulic retention times in the second half was fairly constant. From the graph it is difficult to tell whether the organic loading rate had any significant effect on the COD removal when it was well within the recommended loading rates for UASBs.

4.3.14 pH

The pH values were measured for the influent into the reactor and the effluent from the reactor (Figure 4.13).

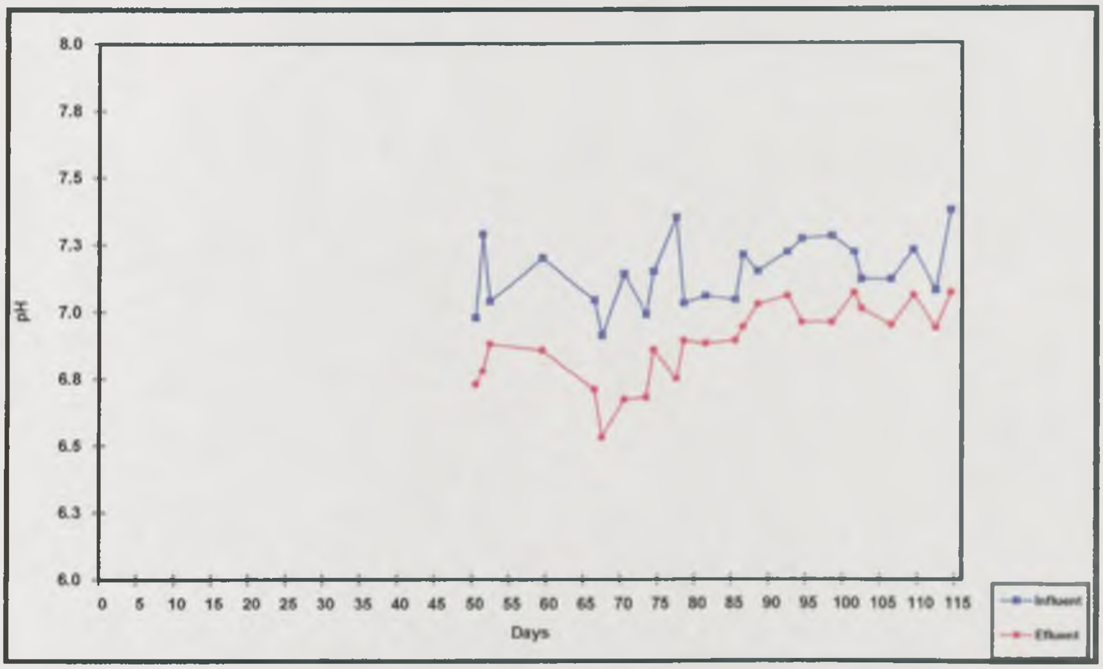


Figure 4-13 pH values in influent/effluent for primary sludge.

The pH values ranged from 6.7 to 7.4 for the influent and from 6.4 to 7.1 for the effluent. The mean pH values for the influent and effluent were respectively 7.1 and 6.9. Except for a couple of instances, the pH was well within 7.0 ± 0.5 units. In general there was a fall in the value of the pH as the sample passed through the reactor. This fall was not unexpected as the whole anaerobic digestion process produces intermediate acidic products. The gap in the first portion of the graph was due to the fact that the pH measurements started later on in the study.

4.3.15 Methane Production for Primary Sludge

Biogas production was observed and the volume of the biogas collected were recorded throughout the duration of the experiment. The production of biogas was used mainly as an indication of the progress of the digestion process. The controlled room temperature of the of the experimental site was 20°CThe volume of methane in the biogas was then calculated using equations 2.14, 2.15, 2.16 and 2.17 (section 2.10.7) and assuming that the volume of methane in the collected gas was 70% (Figures 4-14, 4-15, 4-16 and Table 4-9).

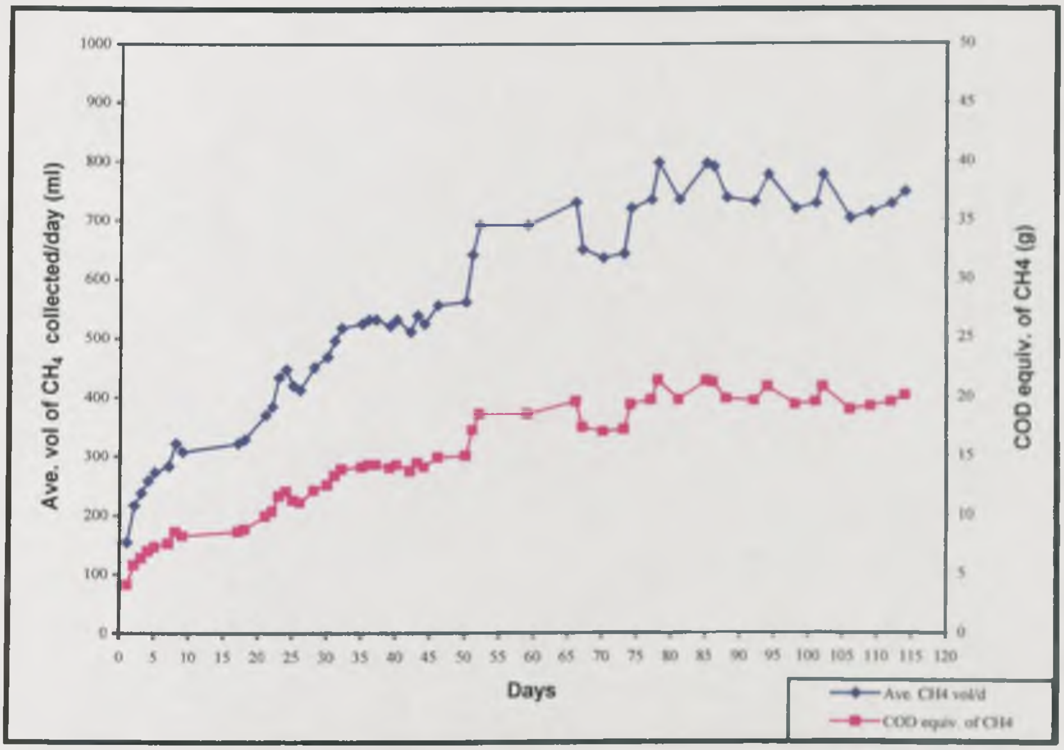


Figure 4-14 Calculated volume of CH_4 and COD equivalent of CH_4 in the collected biogas for primary sludge

Table 4-9 Summary of CH_4 production values for primary sludge

Parameter	Cal. vol. of CH_4 produced daily (ml)	COD equiv. of CH_4 (g COD)	COD_{CH4}/ $COD_{tot. rem.}$ %	Vol of CH_4/ $COD_{tot\ rem}$ l/kg	Vol of CH_4/ TVS_{rem} l/kg
Maximum	800	21	19	7.0	13.6
Minimum	150	4	10	3.9	6.0
Mean	550	15	14	5.4	9.3
Standard Deviation	180	5	2	0.8	1.8

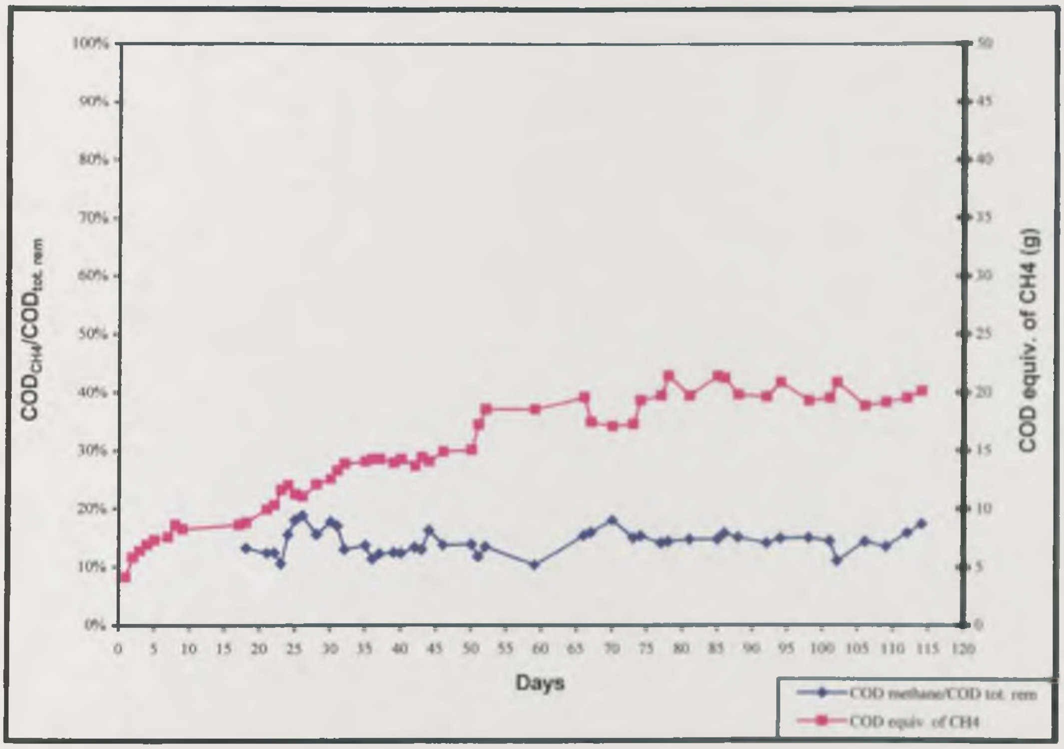


Figure 4-15 COD equivalent of CH_4 and the ratio of $COD_{CH4}/COD_{total\ removed}$ for primary sludge with time

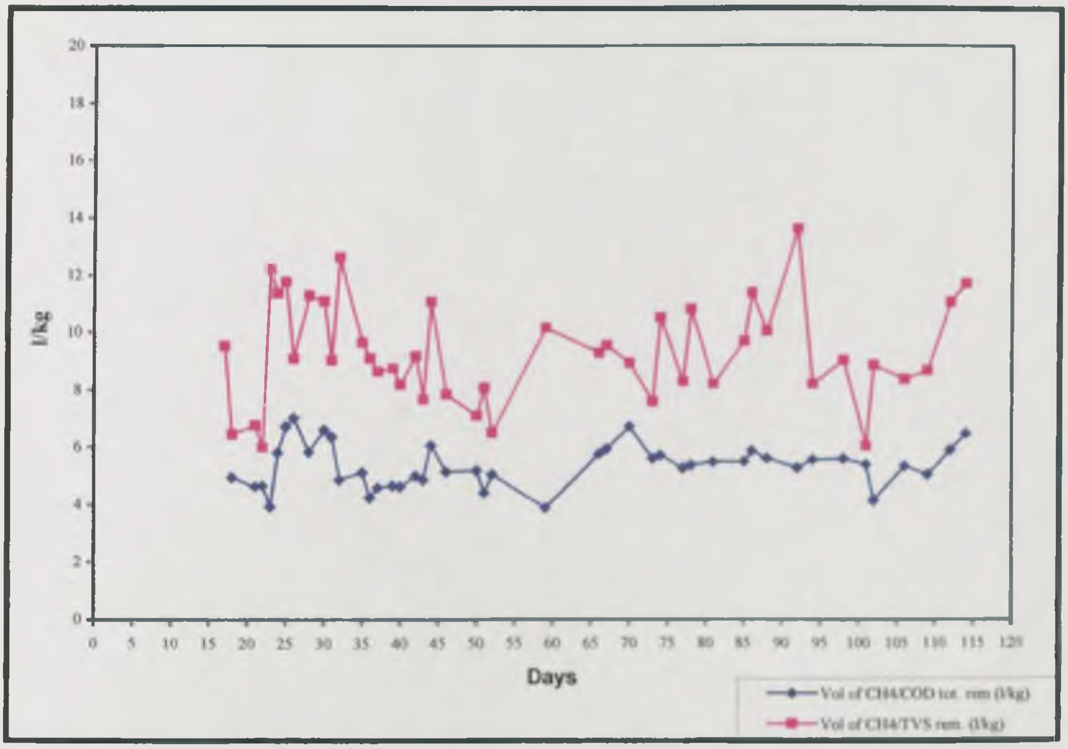


Figure 4-16 Volume of CH_4/COD $_{total\ removed}$ and volume of CH_4/TVS $_{removed}$ for primary sludge with time.

The calculated volume of methane in the collected biogas produced daily ranged from 150 ml/d at the start to as high as 800 ml/d. The mean volume of methane over the experimental duration was 550 ml/d with a standard deviation of 150 ml/d. The calculated volume of methane in the biogas produced daily rose steadily from the start of the experiment as shown in Figure 4-14. However after the 7th week, the volume appears to have stabilised. From day 52 till the end of the experiment, the calculated volume of methane in the biogas produced ranged from 640 ml/d to 800 ml/d, and had a mean value of 730 ml/d with a standard deviation of 45 ml/d. The variation in the calculated volume of methane in the biogas produced after the 7th week was reduced substantially comparing the standard deviation for this period (45 ml/d) to the standard deviation for the whole experimental duration (150 ml/d). That is, for the whole test period the degree of variation about the mean was 27 % while for the latter half this reduced substantially to 6%. Figure 4-14 also shows the COD equivalent of the calculated volume of methane in the biogas produced. The graph pattern is similar to that of the calculated volumes because the COD equivalence methane was calculated using the calculated volumes of methane in the biogas produced. The minimum COD equivalence of methane was 4 g COD and the maximum was 21 g COD during the experimental period. The mean for the test period was 15 g COD with a standard deviation of 5 g COD. As expected and shown in the graph, after the 7th week, the COD equivalent of the calculated volume of methane appears to have stabilised. During this latter period the COD equivalent of methane ranged from 17 to 21 g COD with a mean of 20 g COD and a standard deviation of 1 g COD.

Figure 4-15 shows the ratio of COD of methane to the total COD removed in the UASB rector during the test. There is no apparent trend in the graph as it is almost horizontal with some variations. The ratio ranged from 10% to 19% with a mean of 14% and a standard deviation of 2%. That is, on the average, the calculated amount of methane in the biogas collected accounts for 14% of the total COD removed in the UASB reactor, not accounting for biogas losses, which could be between 20 to 50% (van Haandel and Lettinga, 1994) of the collected biogas volume.

The ratio of the volume of methane produced per g of COD and TVS removed is shown in Figure 4-16. There is no apparent trend in the graphs. The calculated volume of methane per g of TVS removed was always higher than that for the COD. The calculated volume of methane in the biogas produced per kg of COD removed ranged between 3.9 to 7.0 l/kg COD. It averaged 5.4 l/kg COD and had a standard deviation of 0.8 l/kg COD. With regards to TVS, the calculated volume of methane in the biogas produced per kg of TVS removed ranged between 6.0 to 13.6 l/kg TVS. It averaged 9.3 l/kg TVS and had a standard deviation of 1.8 l/kg TVS. Although the values for TVS were higher than that for COD, the graph shows that the values for TVS were erratic and varied considerably (had a standard 1.8 compared to 0.8 for COD values). These values compared with the expected values and discussed in section 5.5.5 of the thesis.

4.4 Results for tests carried out under Phase 3

4.4.1 Ambient temperature measurements

As described in section 3.4.3, a 50-litre UASB reactor was used for the anaerobic digestion of the faecal sludge. Unlike the set-up used in Phase 2, no conventional heating was applied to the reactor in this case. The UASB reactor was mounted in the open and hence the main source of heating was solar radiation. The air temperatures obtained from the meteorological office in Kumasi, Ghana during the experimental period are shown in Figure 4.17. No temperature measurements of the reactor contents were taken.

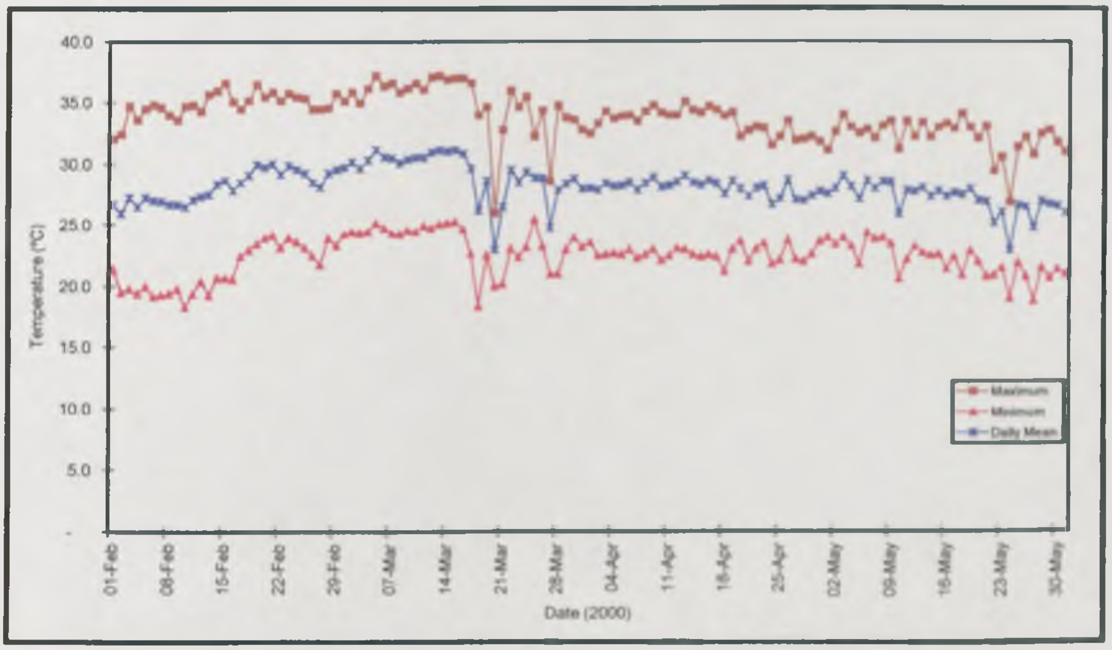


Figure 4-17 Temperature values over the experimental period

The maximum daily temperatures ranged from 26.0 to 37.2 °C with a mean maximum of 33.8 °C. The minimum daily temperature was never lower than 18.3 °C and was as high as 25.5°C. The average value for the minimum daily temperature was 22.4 °C. The mean daily temperature, calculated from the daily maximum and minimum temperatures ranged from 23.0 to 31.2 °C and it averaged 28.1 °C. The period from February to May does not include the hottest months of the year in Ghana, which is normally from November to January. With the mean daily temperature averaging 28.1 °C, the UASB reactor should not require conventional heating for its operation.

4.4.2 Chemical Oxygen Demand

The COD of the faecal sludge was measured for the influent into and the effluent from the UASB reactor and the results obtained are shown Figure 4-18 and Table 4-10.

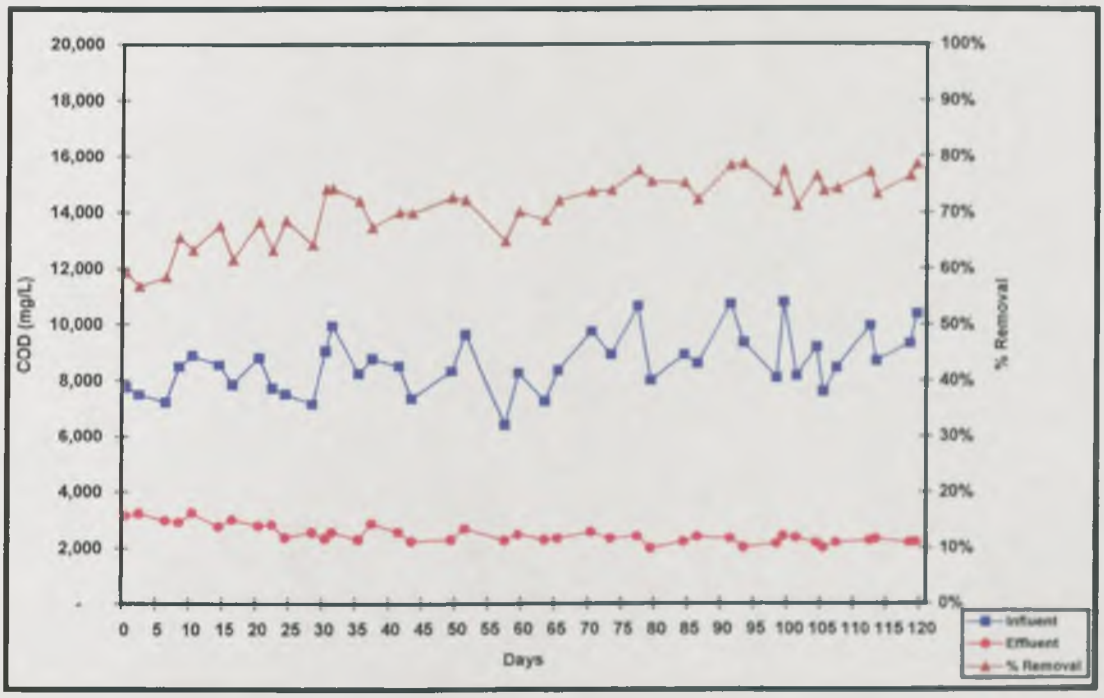


Figure 4-18 COD values and % removal efficiency with time for faecal sludge

Table 4-10 Summary of COD values for influent/effluent samples from the UASB reactor using faecal sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	10780	3270	79
Minimum	6400	1970	57
Mean	8610	2480	71
Standard Deviation	1040	340	6

The influent COD concentration ranged from 6400 to 10780 mg/l and it averaged 8610 mg/l with a standard deviation of 1040 mg/l. The effluent COD was never less than 1970 mg/l and was as high as 3270 mg/l. The effluent averaged 2480 mg/l and had a standard deviation of 340 mg/l. The graph shows a slightly downward trend in the effluent concentrations, an indication of improving effluent quality as the study progressed although the influent concentration was quite varied. The variation in the influent concentration was due to the fact that the origin of faecal sludge varied much. The downward trend in the effluent concentration was reflected in the COD removal which showed an upward trend from start to end. The removal rate varied from 57% to 79% and it averaged 71% and a standard deviation of 6%.

4.4.3 Total Solids

The total solids of the faecal sludge was measured for the influent into and the effluent from the reactor and the values obtained from the determination are shown in Figure 4-19 and Table 4-11.

Table 4-11 Summary of TS values for influent/effluent samples from the UASB reactor using faecal sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	10970	3800	70
Minimum	5560	1990	53
Mean	7560	2900	61
Standard Deviation	1390	500	6

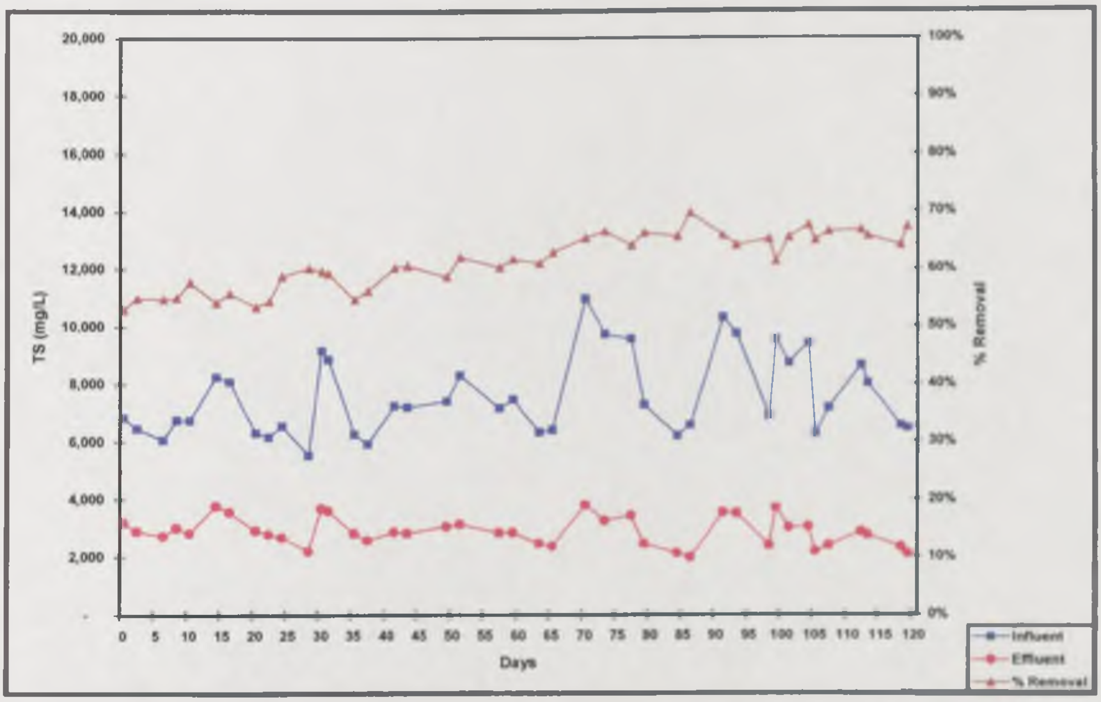


Figure 4-19 TS values and % removal with time for faecal sludge.

The influent total solids concentrations were quite variable, ranging between 5560 and 10,970 mg/l. It averaged 7560 mg/l with a standard deviation of 1390 mg/l. As mentioned in earlier paragraphs, the variable nature of the influent is due to the varied sources of the faecal sludge. The effluent concentration was fairly variable compared to the influent. It was not less than 1990 mg/l and in some instances as high as 3800 mg/l. The mean effluent concentration was 2900 mg/l and had standard deviation of 500 mg/l. Total solids removal rate was quite good and comparable to that for the COD. The minimum removal efficiency was 53% while the maximum was 70%. The mean removal efficiency was 61% with a standard deviation of 6%. The graph shows an upward trend for the removal efficiency, a very good indication that as the study progressed, the removal efficiency was rising. The increasing trend from start to end is similar to that for the COD removal.

4.4.4 Total Volatile Solids

The total volatile solids content was measured for the influent into the reactor and effluent from the reactor to determine the organic fraction of total solids (Figure 4-20 and Table 4-12).

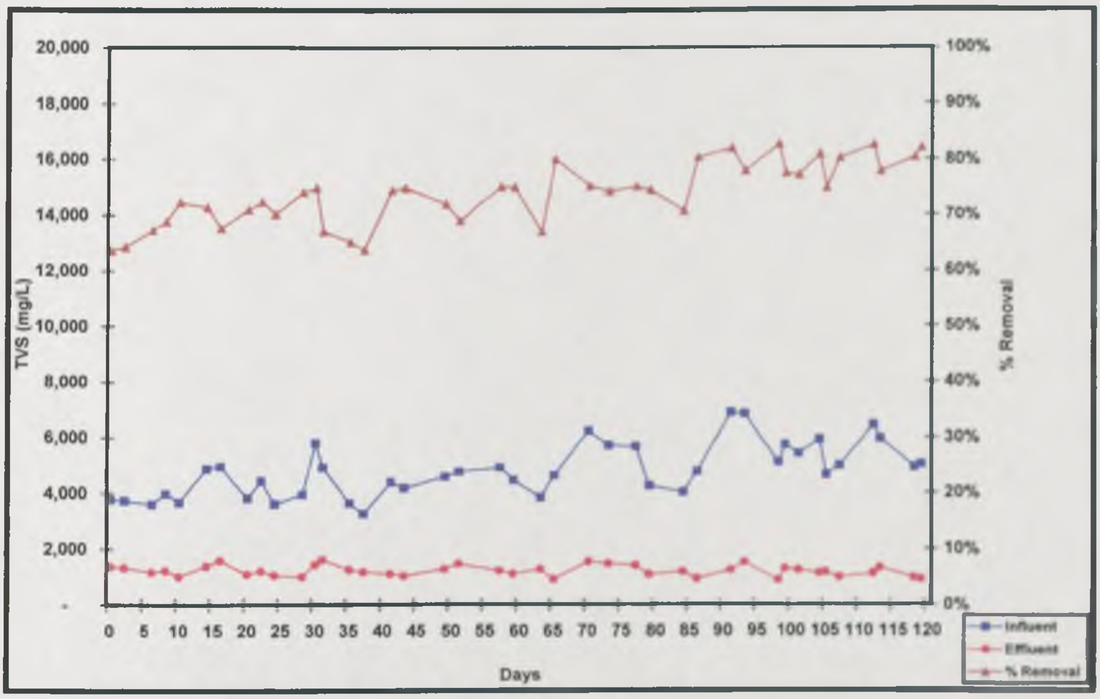


Figure 4-20 TVS values and % removal with time for faecal sludge.

Table 4-12 Summary of TVS values for influent/effluent samples from the UASB reactor using faecal sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	6890	1630	83
Minimum	3290	880	64
Mean	4800	1220	74
Standard Deviation	930	200	6

The influent TVS ranged from 3290 to 6890 mg/l with an average of 4800 mg/l and a standard deviation of 930 mg/l. The influent was quite variable as indicated by the standard deviation. An unexplained upward trend is noticeable in the graph, which maybe due to better quality faecal sludge samples. The effluent quality was fairly stable (standard deviation was 200 mg/l). It varied between 880 and 1630 mg/l and had an average value of 1220 mg/l. The removal efficient was very good and it showed an upward trend from start to finish. The minimum removal efficiency was 64% and the maximum 83%. The mean removal efficiency was 74% with a standard deviation of 6%. The mean removal efficiency is a very good indication of high uptake rate for the organic fraction of the total solids.

4.4.5 Ratio of total volatile solids (TVS) to total solids (TS) for faecal sludge

The ratio of TVS to TS for both influent and effluent was determined using the values obtained from the determination of TVS and TS (Figure 4.21). Comparison of the influent and effluent values gives an indication of the utilisation of the organic content in the reactor.

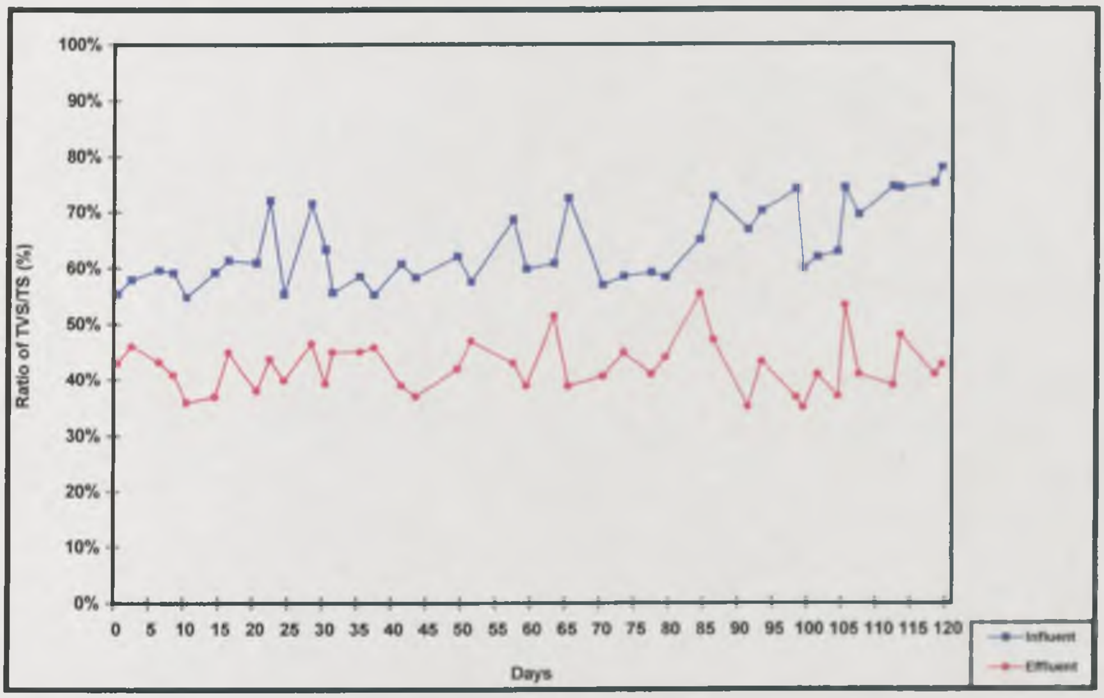


Figure 4-21 Ratio of TVS/TS with time for faecal sludge.

The influent TVS/TS ratio ranged between 55 to 78% with an average value of 64% while the effluent had a minimum value of 35% and a maximum value of 55%. The mean effluent ratio was 42%, a mean drop of 22% from the influent TVS/TS ratio. As indicated on the graph, the influent ratio was always higher than the effluent ratio, an indication of the utilisation of the organic fraction during the anaerobic digestion.

4.4.6 Total Suspended Solids (TSS)

Figure 4.22 and Table4-13 show the results obtained in the determination of the total suspended solids (TSS) content in the influent into the reactor and the effluent from the reactor. Also shown on the graph is the percent removal efficiency.

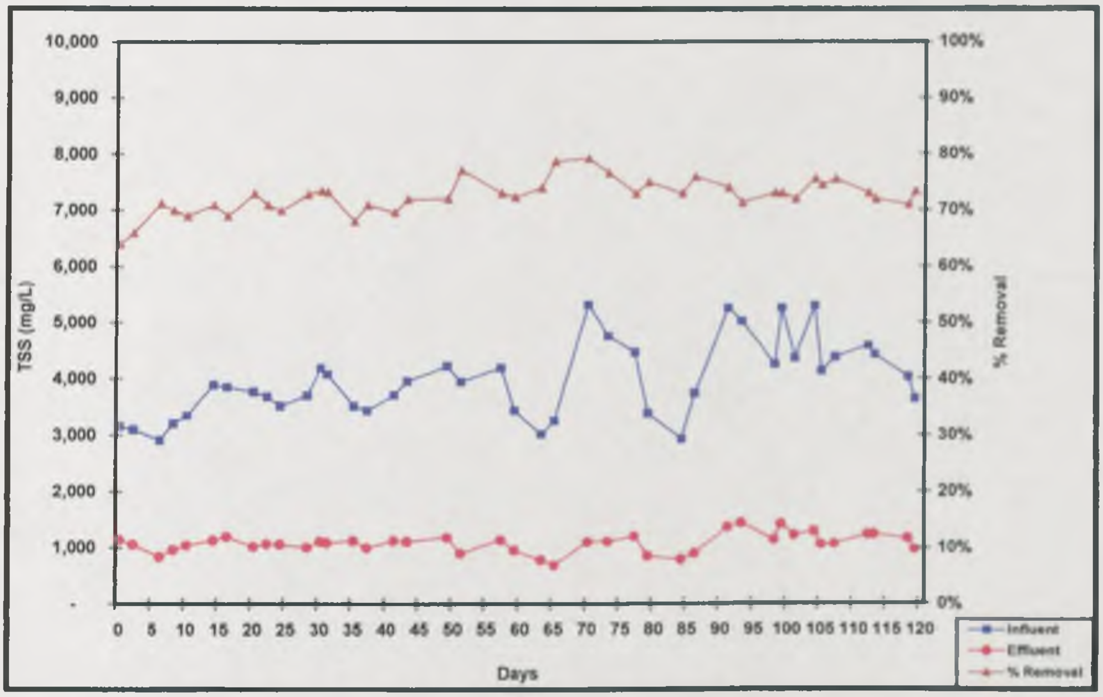


Figure 4-22 TSS values and % removal efficiency with time for faecal sludge.

Table 4-13 Summary of TSS values for influent/effluent samples from the UASB reactor using faecal sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	5320	1440	79
Minimum	2920	690	64
Mean	3960	1080	73
Standard Deviation	670	160	3

The influent concentration of the total suspended solids ranged from 2920 to 5320 mg/l and had a mean of 3960 mg/l with a standard deviation of 670 mg/l. The graph shows quite a variation in the influent TSS values and this is indicated by the large value of the standard deviation. This variation in the influent TSS is due to the different sources of faecal sludge samples. With regards to the effluent, the total suspended solids concentration varied between 690 and 1440 mg/l and averaged

1080 mg/l with a standard deviation of 160 mg/l. As shown on the graph and indicated by the small value of the standard deviation, the degree of variation of the effluent TSS was very little compared to the influent TSS concentrations. The removal efficiency was very good, ranging between 64% and79% with an average of 73% and a standard deviation of 3%. The graph of removal efficiency showed a slight upward trend from start to finish. The fairly little variation in the effluent TSS may be an indication of the stability of the system.

4.4.7 Ratio of total suspended solids (TSS) to total solids (TS)

The ratio of TSS to TS was determined using the values obtained from the determination of TSS and TS and the values are shown in Figure 4-23.

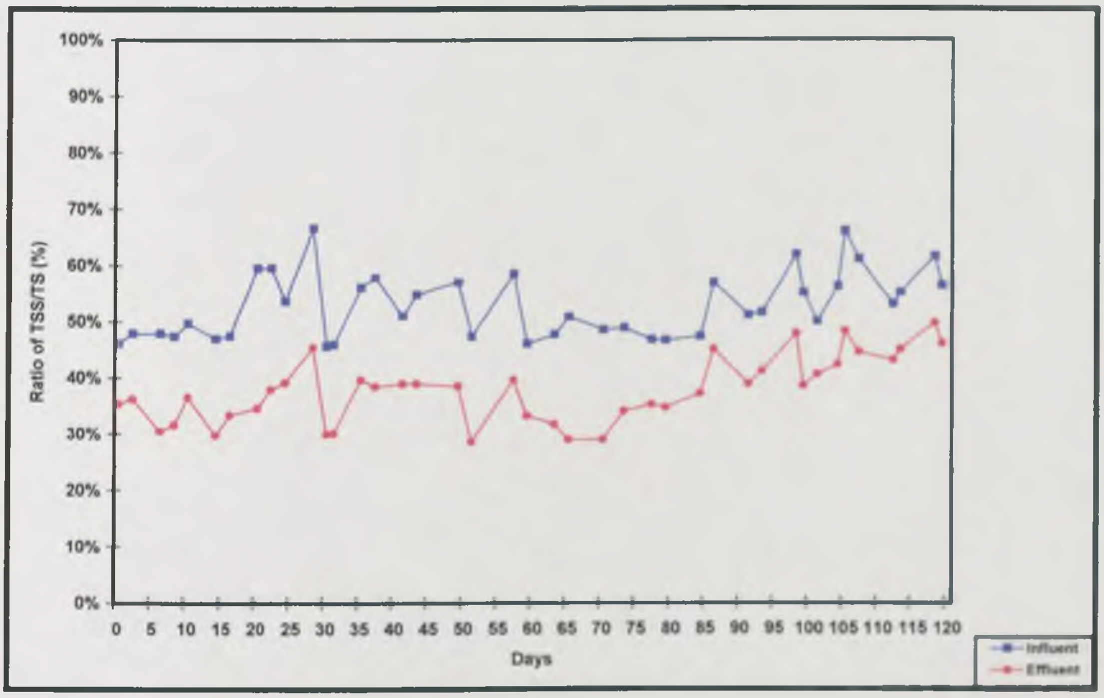


Figure 4-23 Ratio of TSS/TS with time for faecal sludge

The ratio of TSS/TS for the influent ranged from 46 to 67% and averaged 53% with a standard deviation of 6%. With regards to the effluent, the TSS/TS ratio was never less than 29% and was as high as 50%. The mean TSS/TS ratio for the effluent was 38% and had a standard deviation of 6%. The graph shows that at any instance, the influent ratio was higher than the effluent ratio, the mean drop in TSS/TS ratio being

15%. Thus a lesser proportion of the total solids in the effluent were suspended compared to the influent. Both influent and effluent ratios were quite variable and do not show any trend in the graphs.

4.4.8 Removal efficiencies of TS and TSS for faecal sludge

Figure 4-24 shows the removal efficiencies for total solids and total suspended solids over the experimental period. As shown in the figure, both graphs show an upward trend in the removal efficiencies for total solids and total suspended solids. It is also evident from the graph that the removal efficiency for total suspended solids was higher than that of the total solids, the difference ranging from 6% to 20%. The differences were higher at the start and tended to narrow as the study progressed, with the removal efficiency of TS showing a higher rise than that of TSS. That is the increase in the removal efficiency of total solids as the study progressed was better than that for total suspended solids.

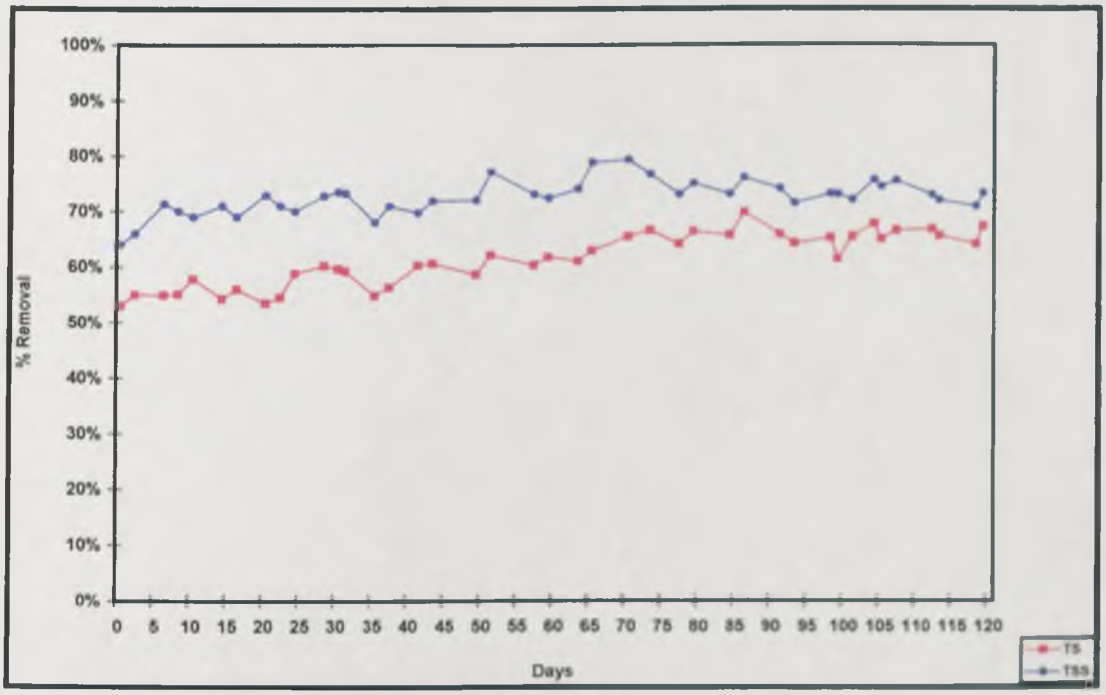


Figure 4-24 **Percent removal efficiencies of TS & TSS with time for faecal sludge.**

4.4.9 Organic Nitrogen

As mentioned in section 3.4.4.1 in the TKN analysis, the sample was first boiled to drive off ammonia and then digested. The measured values therefore represent the organic nitrogen in the samples analysed. Organic nitrogen was measured for the influent into and effluent from the UASB reactor and the values obtained are shown in Figure 4.25 and Table 4-14.

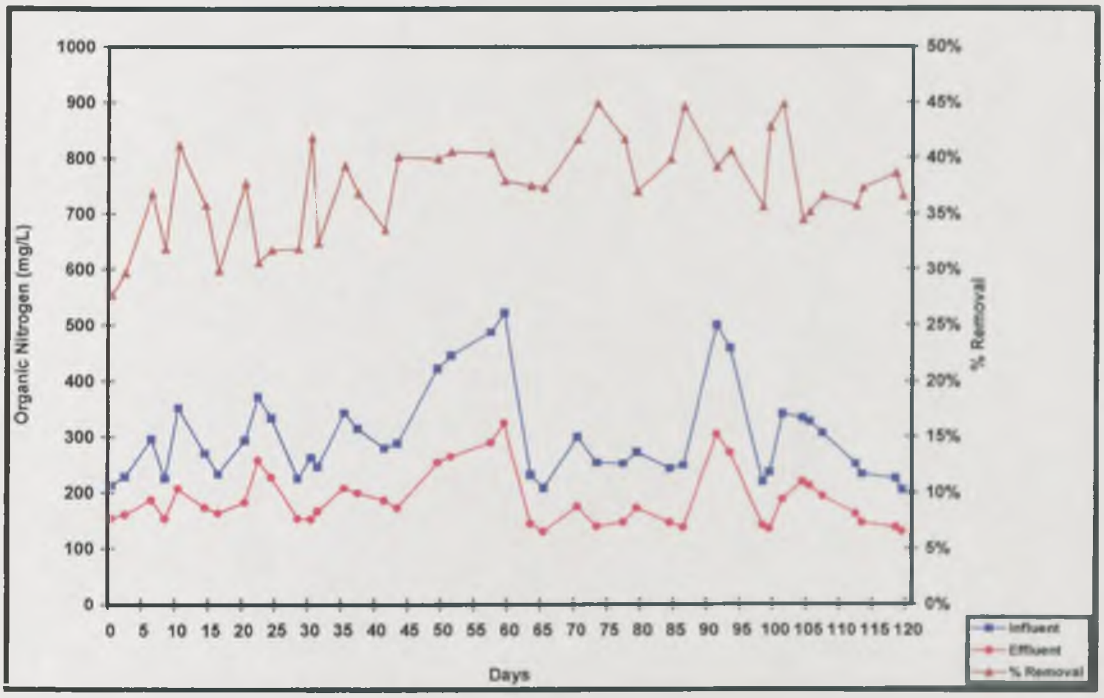


Figure 4-25 Values of organic nitrogen and % removal efficiency with time for faecal sludge.

Table 4-14 Summary of organic nitrogen values for influent/effluent samples from the UASB reactor using faecal sludge

Parameter	Influent (mg/l)	Effluent (mg/l)	% Removal
Maximum	520	325	45
Minimum	205	130	28
Mean	300	185	37
Standard Deviation	85	50	4

The influent organic nitrogen ranged from 205 to 520 mg/l and had a mean of 300 mg/l with a standard deviation of 85 mg/l. With regards to the effluent, the organic nitrogen varied between 130 and 325 mg/l. The mean effluent organic nitrogen was 185 mg/l and the standard deviation was 50 mg/l. The influent values are quite variable and the effluent values do not appear to follow any trend. From the graph it could be seen that the effluent variation was patterned after the influent concentrations, an indication of the influence of the influent concentration.

The anaerobic digestion process led to a decline in the organic nitrogen content. During anaerobic digestion soluble organic nitrogen is converted into ammonia nitrogen as bacteria consume organic matter containing nitrogen. The consumption of soluble organic matter containing nitrogen results in a decline in the organic nitrogen content. The organic nitrogen removal efficiency ranged from 28% to 45% with an average of 37% and a standard deviation of 4%.

4.4.10 Organic loading rate (OLR) and percent removal efficiencies of COD, TS and TVS for faecal sludge.

Figure 4-26 shows the applied organic loading rate and the percent removal efficiencies for COD, TS and TVS. The applied organic loading rate was calculated using the influent COD concentration and the measured hydraulic retention time. The applied organic loading rate varied between 12.5 and 21.5 kg COD/m^3.d and it had a mean value of 17.1 kg COD/m^3.d. The graph indicates that the OLR was quite variable. This was due mostly to the variable nature of the influent COD since the hydraulic retention remained fairly constant throughout the study period. The removal rate for COD, TS, and TVS all showed an upward trend from start to end of the study.

The retention time in the reactor ranged from 11.7 to 12.4 hours with a mean value of 12.1 hours (mean upflow velocity of 0.14 m/h). A dilution ratio in the range of 1:6 – 1:8 was applied to the faecal sludge throughout the experimental study.

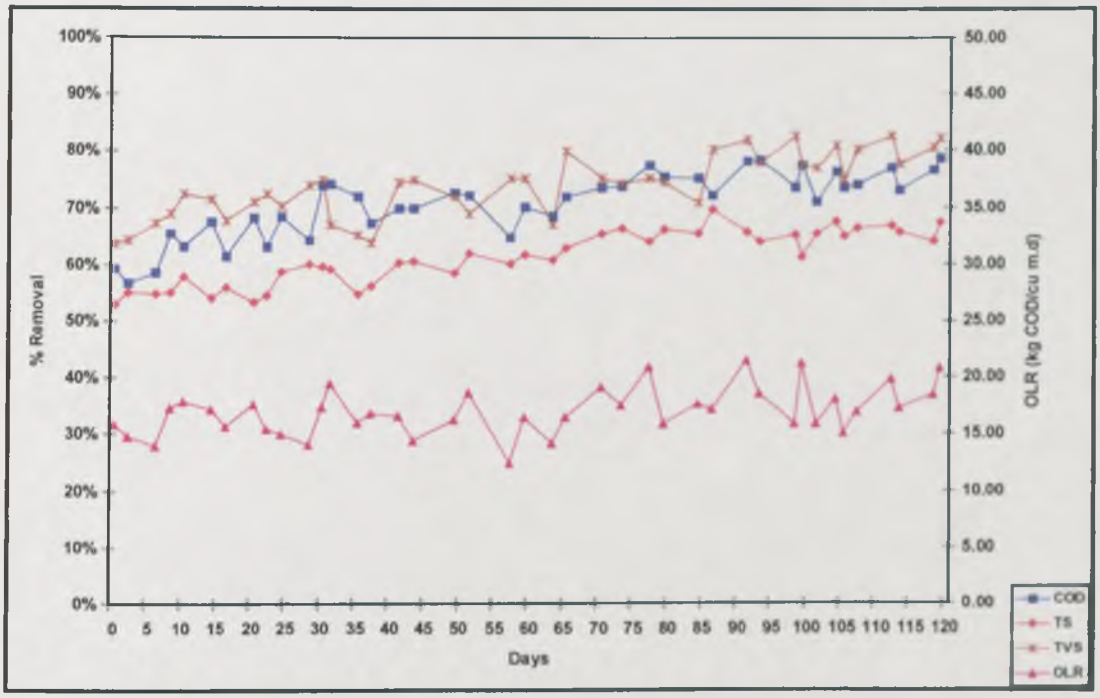


Figure 4-26 Percent removal efficiencies for COD, TS, TVS and the variation of OLR with time for faecal sludge

4.4.11 Methane Production for Faecal Sludge

The production of biogas was observed and volumes of the biogas collected were recorded throughout the duration of the experiment. Again the production of biogas was used mainly as an indication of the progress of the digestion process. The average air temperature over the experimental period was 28°C. The volume of methane in the biogas collected was then calculated using equations 2.14, 2.15, 2.16 and 2.17 (section 2.10.7) and assuming that the volume of methane in the collected gas was 70% (Figures 4-27, 4-28, 4-29 and Table 4-15).

Table 4-15 Summary of CH_4 production values for faecal sludge

Parameter	Cal. vol. of CH_4 produced daily (ml)	COD equiv. of CH_4 (g COD)	COD_{CH4}/ $COD_{tot\ rem}$ %	Vol of CH_4/ $COD_{tot\ rem}$ l/kg	Vol of CH_4/ TVS_{rem} l/kg
Maximum	4000	105	20	7.6	16.0
Minimum	1900	50	11	4.1	6.8
Mean	3280	85	14	5.5	9.6
Standard Deviation	480	12	2	0.6	2.0

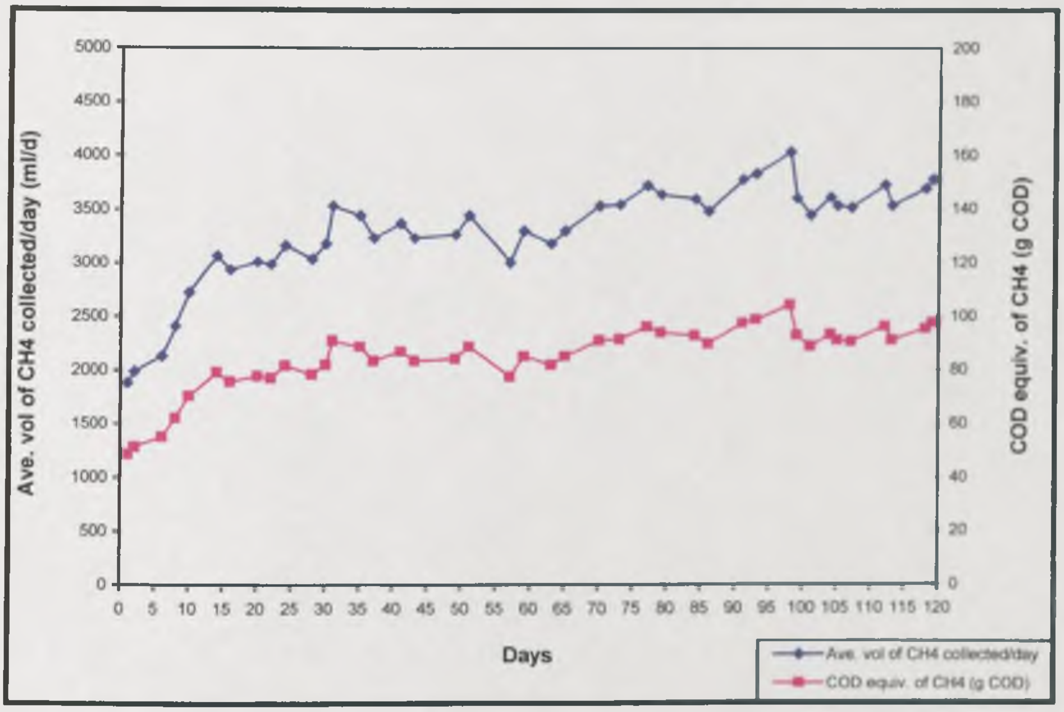


Figure 4-27 Calculated volume of CH_4 and COD equivalent of CH_4 in the collected biogas for faecal sludge

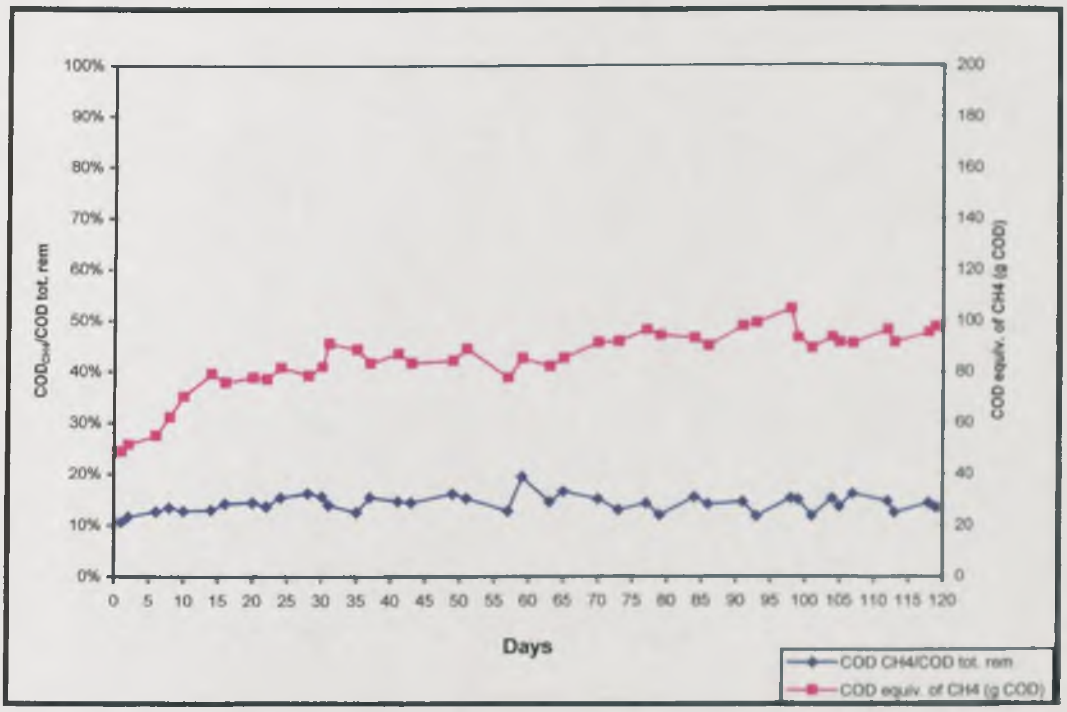


Figure 4-28 COD equivalent of CH_4 and the ratio of $COD_{CH4}/COD_{total\ removed}$ for faecal sludge

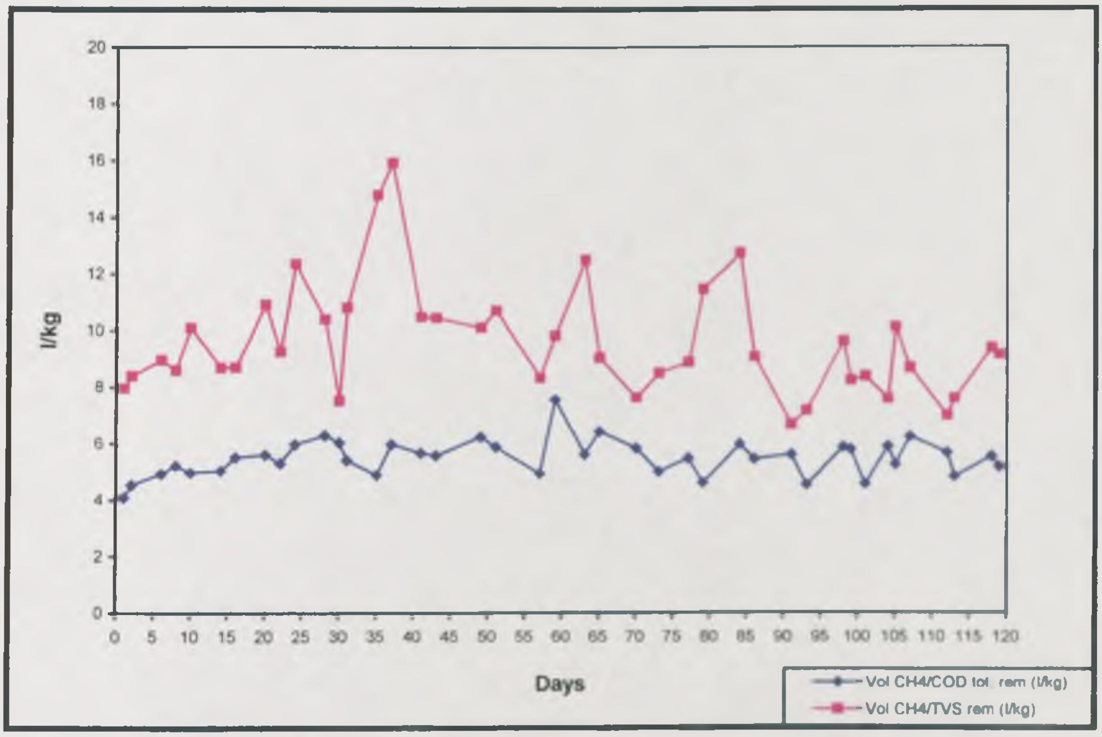


Figure 4-29 **Volume of CH_4/COD $_{\text{total removed}}$ and volume of CH_4/TVS $_{\text{removed}}$ for faecal sludge**

Volume of methane in the collected biogas produced daily was calculated and ranged between 1900 ml/d and 4000 ml/d. The mean volume of methane over the experimental duration was 3280 ml/d with a standard deviation of 480 ml/d. The calculated volume of methane in the biogas generally increased as the experiment proceeded (Figure 4-27). For the latter half of the experimental period (after the 7th week) the minimum calculated volume of methane was 3000 ml/d. The mean calculated volume of methane in the collected biogas for this latter half was 3560 ml/d with a standard deviation of 240 ml/d. Thus for the whole test period the degree of variation about the mean was 14%, while for the latter half this reduced to 6%. That is, the variation in the volume of biogas collected about the mean volume tended to decrease as the experiment progressed. Figure 4-27 also shows the COD equivalent of the calculated volume of methane in the biogas produced. The pattern of the graph is similar to that of calculated volume because the COD equivalence methane were calculated using the calculated volumes of methane in the biogas produced. The minimum COD equivalence of methane was 50 g COD and the maximum was 105 g COD during the experimental period. The mean COD equivalence for the calculated volume of methane in the biogas collected was 85 g COD with a standard deviation of 12 g COD. During the latter same latter half of the period considered, the COD equivalent of the calculated volume of methane in the biogas collected ranged between 80 and 105 g COD with a mean of 90 g COD and a standard deviation of 6 g COD.

Figure 4-28 shows the ratio of COD of methane to the total COD removed in the UASB rector during the experiment. There is no apparent trend in the graph as it is almost horizontal with some variations. This result is similar to the result obtained for primary sludge. The ratio ranged from 11% to 20% with a mean of 14% and a standard deviation of 2%. That is on the average, the calculated amount of methane in the biogas collected accounts for 14% of the total COD removed in the UASB reactor. This result is the same as that obtained for primary sludge. Again this value does not account for biogas losses, which could be between 20 to 50% of the collected biogas volume (van Haandel and Lettinga, 1994).

The ratio of the volume of methane produced per g of COD and TVS removed is shown in Figure 4-29. As in the case of primary sludge, the calculated volume of methane per g of TVS removed was always higher than that for the COD. The calculated volume of methane in the biogas produced per kg of COD removed ranged between 4.1 to 7.6 l/kg COD. It averaged 5.5 l/kg COD and had a standard deviation of 0.6 l/kg COD. With regards to TVS, the calculated volume of methane in the biogas produced per kg of TVS removed ranged between 6.8 to 16.0 l/kg TVS. It averaged 9.6 l/kg TVS and had a standard deviation of 2 l/kg TVS. These values compared with the expected values and discussed in section 5.5.5 of the thesis.

There is no apparent trend in the graphs except to mention that the variations in the volumes associated with TVS removed were substantial compared to variations in volumes associated with total COD removed (the standard deviation about the mean for TVS values was 21% and that for the COD values was 11%).

Chapter Five

5 Discussion

5.1 Introduction

This chapter discusses the results of the experimental work presented in the previous chapter. As revealed by the literature review the UASB reactor has not been used specifically for the treatment of faecal sludges (as defined in this thesis). Its extensive usage in recent times has been for the treatment of domestic sewage and industrial wastewater. In the discussion, the results obtained have been compared to results obtained from USAB reactors treating domestic sewage. The potential of using the USAB for the treatment of faecal sludges, based on the experimental results, is presented.

5.2 Physical and Chemical Characteristics of Faecal Sludges

The physical and chemical characteristics of faecal sludges exhibit extreme variations (Table 4.1). These variations have also been observed in a number of earlier studies (Pescod, 1971; Pradt 1971; Um and Kim, 1986; Guo *et al.*, 1991; WRRI/SANDEC, 1994; and Strauss *et al.*, 1997) and have been attributed to several of factors which included:

i. origin of the sludge,
ii. type of on on-site sanitation system,
iii. amount of ageing that has taken place,
iv. time of the year,
v. extent of stormwater and groundwater infiltration, and
vi. user habits.

The variation is highlighted in Table 5.1 where characteristics of toilet sludge from Accra, Ghana are compared to those obtained in Kumasi, Ghana during this research. In Accra, the minimum BOD was 3,800 mg/l while the minimum for Kumasi was 14,250 mg/l; the minimum value for Kumasi was close to the maximum for Accra, which was 15,000 mg/l. The mean BOD for Kumasi (23,300 mg/l) was nearly three times the mean BOD reported for Accra (8,800 mg/l). With regards to COD, the minimum for Kumasi (36,600 mg/l) was over three times the minimum for Accra

Table 5.1 Comparison of the characteristic of faecal sludges from Kumasi to that reported in the literature and for Accra.

Parameter	Origin		
	Kumasi, Ghana[1]	Accra, Ghana[2]	Literature
pH	8.1 – 8.5	-	7.2 – 8.8
BOD (mg/l)	14,250 – 52,000 (23,300)[3]	3,800 – 15,000 (8,800)	8,000 – 23,000
COD (mg/l)	36,600 – 175,000 (86,700)	10,400 – 97,000 (47,600)	10,000 – 97,000
COD:BOD	2.4:1 – 7.2:1 (3.7:1)	-	1.5:1 – 5:1
TS (mg/l)	31,300 – 87,000 (55,700)	-	12,000 – 45,100
TVS (% of TS)	48 – 76 (71)	62	≥ 50
SS (mg/l)	-	2,000 – 19,000 (6,400)	7,000 – 20,000
VSS (mg/l)	-	58 (% SS)	5,800 – 13,000
NH_4-N (mg/l)	700 – 4050 (2400)	-	2,100 – 6,000
Helminth eggs (no./l)	-	29,000 (3,600 - 62,000)	-

[1] Source: Present study; [2] Source: WRRI/SANDEC, (1994); [3] Values in brackets are means.

(10,400 mg/l) while the maximum value (175,000 mg/l) was almost twice that for Accra (97,000 mg/l). One major factor that could account for this extreme variation of the values in this study to that reported for Accra is the amount of water used at the on-site sanitation facility. Water may be used to flush the toilets in the case of WCs or used for cleaning the toilets in the case of KVIPs and aqua privies. Water used dilutes the faecal sludge and ultimately affects its concentration. Since the mid 1980's the local government authorities in the major cities in Ghana have banned the use of bucket latrines and privatised the management of public toilets. Schemes were put in place to assist homeowners to convert existing bucket latrines to other hygienically more acceptable sanitation systems. Public bucket latrines are also being converted to other systems such as WCs and aqua privies. The city of Accra, which is the capital, is ahead of the other cities with these improvements and now has more water dependent on-site sanitation facilities than the other cities including Kumasi. It is therefore not surprising that the faecal sludges from public on-site sanitation facilities in Kumasi are more concentrated than that in Accra.

The characteristics of the faecal sludge from this study compare quite satisfactorily with the summary characteristics compiled from the literature (Table 5.1). In general, the values for Kumasi were higher than those reported in the literature. The physico-chemical characteristics indicate that the faecal sludge contains high concentrations of organic matter. Total volatile solids, expressed as a percentage of total solids, range from 48 to 76% and had a mean of 71%. The high percentage of organic material in the faecal sludge, coupled with the wide fluctuations of the physico-chemical characteristics make the anaerobic digestion process a preferred option for faecal sludge treatment.

5.3 Physical and Chemical Characteristics of Untreated Primary Sludges

The physico-chemical characteristics of the untreated primary sludge showed little variation compared with those for the faecal sludges. The values compared well with those in the literature (Table 5.2)

Table 5.2 Comparison of untreated primary sludge characteristics

Parameter	Owlwood STW[1]		Metcalf & Eddy, Inc (1991)	
	Range	Average	Range	Average
pH	5.28 - 5.36	-	5.0 – 8.0	6.0
COD (mg/l)	40,840 - 43,230	41,940	-	-
TS (mg/l)	35,800 - 37,130	36,500	20,000 – 80,000	50,000
TVS (mg/l)	28,100 - 29,680	29,000	-	-
TVS/TS	78.4% - 80.2%	79.6%	60% - 80%	65%
NH_3 (mg/l)	324 – 350	337	-	-

[1] Analysis of 4 grab samples

The total solids concentration from the Owlwood sewage treatment works (STW) fell well within the reported values except that the average value 36,500 mg/l was much lower than that cited in the literature of 50,000 mg/l. Again the ratio of TVS/TS for the untreated primary sludge from Owlwood was within the reported range, but the mean value of 79.6% was much higher than the reported typical average of 65% and closer to the upper limit of 80%. From the TVS/TS ratio, it can be said that the untreated primary sludge from the Owlwood STW, which treats only domestic sewage, contains a high proportion of organic matter. This is to be expected for untreated primary sewage sludge of domestic origin. With regards to the pH the untreated primary sludge was acidic in the range 5.28 – 5.36, the mean of which would be less than the reported average value (6.0) in the literature.

5.4 Operating conditions and parameters

5.4.1 Start-up of the UASB reactors

It is widely reported that the start-up of an anaerobic treatment plant for most wastewaters is time consuming and rather a difficult process, due to the fact that a large bacterial mass, adapted to the particular characteristics of wastewater has to develop. This is particularly true for industrial wastewaters. Faecal sludge, however, differs from most industrial wastewaters in that it already contains the bacterial populations necessary for anaerobic digestion. Thus, a reactor for anaerobic treatment of wastes of faecal origin can be started without the need for inoculation.

A cautious approach was adopted for the start-up of the Leeds reactor by inoculating it with seed from the UASB reactor treating sugar wastes at the British Sugar factory located in York, North Yorkshire, England. The concentration of the waste was gradually increased over the first few weeks by decreasing the dilution ratio from 1:20 to 1:10. This was done to allow the seed to adapt to the waste. The start-up was less problematic than expected and treatment proceeded smoothly from the beginning. Treatment efficiencies for COD, TS and TVS gradually increased from start to finish of the experiment as shown in Figures 4.1, 4.2 and 4.3 respectively.

From the experience obtained with the Leeds reactor, the reactor used for the Kumasi experiment was started by simply feeding it with diluted faecal sludge without any seeding and gradual loading. No difficulties were experienced with the start-up, and the start-up proceeded relatively rapidly for an anaerobic process. Figure 4.18, 4.19 and 4.20 show that the treatment efficiencies for COD, TS and TVS gradually increased from start to end as the Phase 3 experiment progressed. From experiences obtained so far from UASB reactors treating domestic sewage, van Haandel and Lettinga (1994) concluded that UASB systems for sewage treatment could be started up without any serious problems using the empty reactor at the design flow.

With regards to the start-up of UASB reactor treating faecal wastes, the critical factor could be the length of the start-up period and how quickly the desired effluent quality can be achieved. Both experimental results clearly indicated that with time, the effluent quality improves. For both experiments, COD removal was consistently over 60% after the first six weeks. For the Leeds experiment with untreated primary sludge, the mean COD removal efficiency from the 10th week to the end of the experiment was 78% with a standard deviation of 3%. With regards to the Kumasi experiment using faecal sludge, the mean COD removal efficiency from the 10th week to the end of the experiment was 75% with a standard deviation of 2%. These removal efficiencies were quite high and with minimal variation from the means as depicted by the values of the standard deviations. To obtain such high removal efficiencies from the 10th week was quite remarkable. van Haandel and Lettinga (1994) state start-up periods of 12 to 20 weeks for UASB reactors treating domestic sewage. COD removal efficiencies for UASB reactors treating domestic sewage during the start-up period, as reported by van Haandel and Lettinga (1994), were rather lower than that obtained in this research. In fact in their circumstances, they report of decreasing removal efficiency during the initial operation and attributed it to the absence of a sufficient quantity of proper bacterial sludge to carry out the anaerobic digestion of organic material. With UASB reactor treating sludge there is a sufficient quantity of sludge mass from the beginning. As a result, the removal efficiencies follow upward trends from the beginning as the current experiment demonstrates. In the case reported by van Haandel and Lettinga (1994), the performance of the reactor (in terms of COD removal) started to improve, and high and almost constant COD removal efficiencies were established after about 20 weeks of operation.

As mentioned in the previous paragraph, similar removal efficiencies were achieved after about 10 weeks for the laboratory scale UASB reactors treating primary and faecal sludge. This was half the period required for a UASB treating domestic sewage.

5.4.2 Loading Rates

Loading rates applied to UASB reactors are limited by either the hydraulic or organic load depending on the nature of the wastewater.

5.4.2.1 Hydraulic Loading Rate

The hydraulic load, which is numerically equal to the average influent flowrate, is the limiting load on a UASB reactor in the case of dilute wastewater. As shown in equation 2.9 the hydraulic load directly affects both the upflow velocity and retention time for a given reactor configuration. The upflow velocity increases as the hydraulic retention time decreases. High upflow velocity can result in excessive sludge washout. One basic requirement for high rate anaerobic treatment in the UASB reactor is that a large and active sludge mass is retained in the reactor. The maximum hydraulic load that can be applied is therefore limited by the constraint that the upflow velocity in the reactor must not cause excessive sludge washout. van Haandel *et al.* (1996) stated that the upflow velocity must not exceed 1 m/h for sewage treatment in a conventional UASB system.

In the experiments conducted in this research the upflow velocities were much lower than the recommended maximum. For the experiment in Leeds, the average upflow velocity was 0.18 m/h (mean hydraulic retention time 9.8 hours) while for that in Kumasi it was 0.14 m/h (mean hydraulic retention time 12.1 hours). Typical upflow velocities cited in the literature for UASB reactors treating domestic sewage range from 0.24 m/h to 1.43 m/h, with removal efficiencies decreasing as the upflow velocity increases. Retention times also ranged from as low as 1.5 hours to as high as 8.5 days (Seghezzo *et al.*, 1998). In both experiments conducted in this research, hydraulic load was not a limiting factor as the waste treated was concentrated.

5.4.2.2 Organic Loading Rate

For concentrated wastes, the organic load rather than the hydraulic load becomes the limiting factor for the UASB system. The organic loading rate indicates the daily organic load per unit of reactor volume. In practice the organic load is often expressed as kilograms COD (applied) per unit reactor volume per unit time. Applied COD loading rates reported in the literature for UASB systems treating domestic sewage range from as low as 0.19 kg COD/m^3.d to 7.33 kg COD/m^3.d. van Haandel *et al.* (1996) proposed that the maximum design load of organic material may be 20 kg COD/m^3.d for wastes containing a high concentration of dissolved organic material of vegetable origin to be digested at or near the optimal temperature for mesophilic digestion. No reasons were stated for this proposed maximum limit.

In the experiment conducted in Leeds the applied COD loading rate ranged from 5.6 to 15.0 kg COD/m^3.d with an average of 10.1 kg COD/m^3.d. The applied COD loading rate varied from 12.5 to 21.5 kg COD/m^3.d and it averaged 17.1 kg COD/m^3.d for the experiment in Kumasi. These applied loads are much higher than those applied in pilot and full scale UASB reactors treating domestic sewage (Seghezzo *et al.* 1998; van Haandel and Lettinga, 1994). The COD removal efficiencies obtained (as illustrated in Figures 4.1 and 4.15) do not in any way suggest that the systems were overloaded. The experiments thus illustrate that it is possible to increase the organic loading rates to the levels used in the experiments and still have substantial treatment performances.

5.4.3 Temperature

Anaerobic digestion, like other biological processes, is strongly dependant upon temperature. The temperature at which the UASB reactor is operated has an influence on the treatment efficiency. van Haandel and Lettinga (1994), from an extensive studies conducted on full scale UASB plants in Cali, Colombia and São Paulo, Brazil, concluded that "within the temperature range 20-25 °C and applied loading conditions, the temperature does not exert any significant influence on the UASB performance." On the other hand, results obtained by DeMan (1990) and Van der Last (1991) show that at temperatures below 18 °C and notably below 15 °C both the rate and extent of organic matter removal declines considerably. It is clear from the data that "an average retention time of six hours is sufficient in tropical and subtropical regions where the temperature is above 18 °C to achieve a satisfactory treatment efficiency in one compartment UASB reactors" (van Haandel and Lettinga, 1994). van Haandel and Lettinga (1994) further state that "experimental results obtained for moderate climatic conditions indicate that the liquid retention time for conventional designs has to be increased to 12 – 14 hours for a temperature of 10 – 12 °C.

Data for both laboratory and full scale UASB reactors show that the reactor has been used for the anaerobic treatment of sewage in varying temperature conditions ranging from as low as 7 °C to 32 °C (Seghezzo *et al.* 1998). The experiment conducted in Leeds (which has temperate climatic conditions) was carried out under controlled temperature conditions. The bottom half of the UASB reactor was heated (Figure 3.1) and maintained at a temperature of 37 °C, which is considered optimal for mesophilic digestion. This was a precautionary approach to ensure that temperature conditions were optimal for anaerobic digestion. With regards to the experiment carried out in Kumasi (which has a tropical climate) ambient temperature conditions were utilised and no supplementary heating of the reactor was necessary. For the period of the experiment, the mean daily temperature ranged from 23.0 °C to 31.2 °C with an average of 28.1 °C. The ambient temperature range was well within the mesophilic temperature range. No temperature measurements of the UASB reactor contents were taken. The removal efficiencies clearly indicate that the ambient temperatures are adequate for anaerobic digestion in the USAB reactors.

5.5 Performance of the UASB reactors

5.5.1 COD removal efficiencies.

A considerable spread in removal efficiencies of COD in UASB reactors treating domestic sewage is generally observed (Seghezzo *et al.* 1998; van Haandel and Lettinga, 1994). The spread is due primarily to the different types of sewage being treated and the different operational conditions. In general the COD removal efficiencies for a UASB reactor operating in tropical and sub-tropical temperatures (temperatures above 18 °C) are higher than those operating in cold climatic conditions. The COD removal efficiencies are also affected by the hydraulic retention time. For reactors that operate in tropical and sub-tropical temperatures with retention times over 4 hours, the COD removal efficiencies ranged from 45 to 93%. The COD influent concentrations under these conditions also varied from 188 to 1183 mg/l.

Experiments conducted in Leeds using diluted primary sludge were carried out at a nominal temperature of 37 °C and with an average retention time of 9.8 hours. Influent COD concentrations were very high when compared to sewage, ranging from 1770 to 6310 mg/l with a mean value of 3800 mg/l. The COD removal efficiencies were high and well within the values reported for domestic sewage. For the whole duration of the experiment, the COD removal ranged from 53% to 83% with an average of 74%. From the 10th week till the end of the experiment, the COD removal was never less than 71% and was as high as 83%. It averaged 78% during this period. This is very good indication that when steady state conditions have been established under the right operating conditions, the UASB reactor can consistently produce high removal efficiencies with regards to COD removal. The effluent COD concentration from the 10th week till the end of the experiment ranged from 750 to 1450 mg/l with a mean of 1030 mg/l.

The ratio of the filtered COD to total COD in the effluent ranged from 45% to 78% and had a mean of 64%. Thus, on the average, about 36% of the effluent COD can be attributed to suspended solids in the final effluent.

As shown in Figure 4.1, variations in the effluent COD concentrations and corresponding COD removal efficiencies were minimal compared with the influent concentrations in the latter half of the experiment. This could be due to the fact that the reactor was approaching steady state conditions. The effluent COD concentration was too high for direct discharge into surface waters. A post-treatment unit will be required for further treatment to reduce the concentration to an acceptable level.

With regards to the experiments carried out in Kumasi using diluted faecal sludge, ambient temperature conditions were utilised and the hydraulic retention time averaged 12.1 hours. A constant dilution ratio was used from start to end of the experiment and the reactor was not seeded. For the whole duration of the experiment, influent COD concentrations ranged from 6400 to 10,780 mg/l with an average of 8610 mg/l. The COD removal efficiencies over the duration of the experiment were comparable to those reported in the literature, ranging from 57% to 79% with an average of 71%. From the 10th week till the end of the experiments, there was an improvement in the mean COD removal efficiency compared to the removal efficiency for the whole duration of the experiment. From the 10th week till the end of the experiment, the average COD removal efficiency was 75% with a standard deviation of 2%. The effluent COD concentration during this period ranged between 1970 and 2420 mg/l and averaged 2230 mg/l. As shown in Figure 4.18, a general upward trend in the COD removal efficiency with a corresponding improvement in the effluent quality was observed. Although the influent COD concentrations were quite variable as expected, the effluent COD concentrations remained fairly constant.

On the whole, the high removal efficiencies for COD are a good indication of the fact that the UASB, under proper operating conditions, could be used for the pre-treatment of faecal sludges before the conventional faecal sludge treatment plants (FSTPs) as used in Ghana. The pre-treatment would reduce the strength of the waste to acceptable levels that could be handled by the FSTPs without any adverse effects.

5.5.2 Solids removal efficiencies

Removal efficiencies for total solids (TS), total volatile solids (TVS), total suspended solids (TSS) and volatile suspended solids (VSS) were measured for the Leeds experiment and TS, TVS and TSS were measured for the Kumasi experiments.

5.5.2.1 Total Solids (TS)

The removal efficiencies of total solids for both experiments increased gradually over the course of the experiment. In the Leeds experiment, the removal efficiency ranged from 46% to 78% with an average of 62%. Influent concentrations were high, ranging between 2270 mg/l to 6520 mg/l with an average of 3690 mg/l over the whole period of the experiment. The effluent total solids concentration ranged from 980 mg/l to 1970 mg/l with a mean value of 1350 mg/l.

From the 10th week till the end of the experiment, the total solids concentration in the influent for the diluted primary sludge ranged from 3780 to 6520 mg/l with an average of 4570 mg/l. The effluent concentration during this period varied between 1020 and 1970 mg/l, and had a mean of 1380 mg/l. The total solids removal efficiency from the 10th week till the end of the experiment averaged 69%, an increase of 7% over the over the mean for the whole duration of the experiment. This clearly indicates that as the experiment progressed there was an improvement in the performance of the UASB reactor. During this period, the total solids removal efficiency was never less than 58%, an increase of 12% over the minimum TS removal efficiency for the whole experimental duration.

Figure 4.2 shows that the influent total solids concentration were erratic and also showed a general increasing trend over the experimental period while the effluent total solids concentration was fairly constant. Figure 4.2 also shows that towards the end of the experiment, the TS removal efficiency showed a decreasing trend. This was due to decreasing total solids concentration in the influent rather than deteriorating effluent quality.

Removal efficiencies for total solids in the Kumasi experiments were of similar magnitude to those experienced in Leeds, but marginally lower. Furthermore, the total solids concentrations of the faecal sludge were higher than that of the primary sludge. The total solids removal efficiency for faecal sludge was never lower than 53% and was as high as 70%. The average total solids removal efficiency was 61%, which is nearly the same as that for the primary sludge. The total solids concentration in the influent was between 5560 and 10,970 mg/l with a mean of 7560 mg/l. With regards to the effluent, the total solids concentration varied between 1990 and 3800 mg/l and had a mean of 2900 mg/l.

From the 10th week till the end of the experiment, the influent total solids concentration in the diluted faecal sludge ranged from 6190 to 10300 mg/l and it averaged 8060 mg/l. During this period the effluent concentration ranged from 1990 to 3690 mg/l and it averaged 2770 mg/l. These values show that the average influent total solids concentration increased while the average effluent concentration decreased. A good indication of the improvement in the performance of the UASB reactor as the experiment progressed with time. The mean removal efficiency from the 10th week till the end of the experiment was 61%, and increase of 8% over the mean for the whole experimental period.

Figure 4.16 shows that the influent total solids concentration was very erratic and quite variable while the effluent varied very little from one point to the next. This clearly illustrates the ability of the UASB reactor to handle varying influent loads. Although the removal efficiency was very good for both experiments, the effluent concentrations are still high for direct discharge into the environment. It will be essential to have post-treatment of the effluent to improve the effluent quality.

5.5.2.2 Total Suspended Solids (TSS)

Removal efficiencies for total suspended solids (TSS) were higher than that for total solids in both experiments. The total suspended solids removal efficiencies ranged from 80% to 97% with a mean of 91% and 64% to 79% with a mean of 73% for the primary and faecal sludges respectively. In terms of actual concentrations, the influent TSS ranged from 1500 to 5600 mg/l with a mean of 2900 mg/l and from 2920 to 5320 mg/l with an average of 3960 mg/l for primary and faecal sludges respectively. TSS effluent concentration for the Leeds experiment using primary sludge was from 130 to 580 mg/l and it averaged 250 mg/l. Effluent TSS concentrations for the Kumasi experiment using faecal sludge were a higher with a minimum of 690 mg/l, a maximum 1440 mg/l and a mean 1080 mg/l. The wider variation between the influent and effluent TSS for the Leeds experiment was due to the different but decreasing dilution ratios employed from the start of the operation.

During the Leeds experiments, the effluent TSS concentration was fairly constant although the influent TSS concentrations were erratic and increased over the experimental period (Figure 4.5). From the 10th week till the end of the experiment the mean influent TSS concentration, 3610 mg/l, was higher than the mean for the whole experimental period, which was 2900 mg/l. The effluent TSS concentration during this latter period was 130 mg/l, which was the same as that for the whole experimental period, which was 130 mg/l. Once again, this shows an improvement in the performance of the UASB reactor as the experiment progressed with time. The minimum TSS removal efficiency from the 10th week till the end of the experiment increased by 6% over the minimum for the whole experimental period to 86%.

The effluent TSS concentrations in the Kumasi experiment using diluted faecal sludge were fairly constant while the influent TSS concentrations were quite variable (Figure 4.22). From the 10th week till the end of the experiment, the mean influent TSS concentration increased from 3960 mg/l to 4350 mg/l. The mean effluent TSS concentration and the removal efficiency for this period increased marginally over the values for the whole experiment. The minimum TSS removal efficiency from the 10th week to the end of the experiment was never less than 71%, an increase of 7% over the minimum removal efficiency for the whole experiment duration.

Total suspended solids removal efficiencies reported in the literature ranged from as low as 30% to as high as 97% (Seghezzo *et al.*, 1998) for UASB reactors treating domestic sewage with hydraulic retention times of at least 4 hours and at tropical and subtropical temperature conditions. Compared to the removal efficiencies obtained for the experiments conducted in Leeds and Kumasi, the initial TSS removal efficiencies reported for UASB treating domestic sewage was low (van Haandel and Lettinga, 1994). The initial low TSS removal efficiencies reported in the literature for domestic sewage has been attributed to the low concentration of sludge blanket at the initial operation periods, which reduces the likelihood of entrapping non-settleable solids. As the operation proceeds with time, the amount of sludge mass increases, improving the entrapment of non-settable suspended solids and hence increasing the removal efficiency. With regards to primary sludge and faecal sludge, the concentration of sludge mass from the outset was high and hence entrapment of TSS was high from the outset, improving with time. This situation might account for the high removal efficiencies experienced from the start of the experiment in contrast to those reported by van Haandel and Lettinga (1994) due to the low concentration of the sludge blanket. The early development of the sludge blanket is very important for the performance of the UASB reactor and in ensuring a good quality and stable effluent from the anaerobic digester.

van Haandel *et al.* (1996) state that in practice the UASB reactor becomes non-applicable for influent suspended solids concentration beyond 4000 to 6000 mg/l. Experiences from the experiments clearly indicate that this may not be wholly correct and will be dependent on the type of waste and the nature of the solids. For the Leeds experiment the TSS concentrations in the latter half were very close to 4000 mg/l and exceeded it in some instances. With regards to the Kumasi experiment, for a significant period in the last half the TSS concentrations were over 4000 mg/l. In both experiments these high values were very close to and above the limits but did not appear to adversely affect either the removal efficiency or the effluent quality. For primary and faecal sludges originating purely from domestic sources, the solids content after preliminary screenings will be made up of mainly biodigestible organic matter, which can be easily digested anaerobically.

5.5.2.3 Total Volatile Solids (TVS)

The removal efficiencies of total volatile solids (TVS) were of similar order to that of the TSS removal efficiency, but higher than the TS removal efficiency. The TVS removal efficiencies for the diluted primary sludge ranged from 63% to 85% with an average of 75%. In terms of actual concentrations, the influent TVS for the diluted primary sludge ranged from 1170 to 4350 mg/l with a mean of 2220 mg/l. The effluent TVS concentrations varied between 340 and 820 mg/l and it averaged 530 mg/l. From the 10th week till the end of the experiment, the mean influent TVS concentration was 2770 mg/l while the mean effluent concentration was 570 mg/l. Despite this significant rise in the mean TVS influent concentration during the latter half of the experiment, the mean effluent TVS concentration remained almost the same. The mean TVS removal efficiency during this period was 79%.

The TVS removal efficiency for faecal sludge was never less than 64% and was as high as 83%. The mean TVS removal efficiency for faecal sludge was 74%. In terms of actual concentrations, the TVS concentration in the influent faecal sludge varied from 3290 to 6890 mg/l having an average of 4800 mg/l. The effluent TVS concentrations range from 880 to 1630 mg/l for the diluted faecal sludge and had an average of 1220 mg/l. The values clearly show that the TVS concentrations of the faecal sludge were higher than that of the primary sludge. This may be due in part to the different dilution ratios used. From the 10th week till the end of the experiment, the TVS removal efficiency averaged 78%, a 4% rise over the average for the whole duration of the experiment. During the period the mean TVS influent concentration increased significantly to 5440 mg/l while the mean effluent concentration decreased marginally compared to the respective mean values for the whole duration of the experiment.

The high TVS removal efficiencies indicate the effectiveness of the UASB reactor in digesting the diluted primary and faecal sludges under anaerobic digestion when the operating conditions are right. Figures 4.3 and 4.20 show that the effluent concentrations of the total volatile solids for both experiments were fairly constant although the influent concentrations varied considerable. This demonstrates the ability of the UASB reactor to handle quite well varying loads applied to the reactor. The fairly constant effluent quality would be beneficial in the design of any post-treatment facility to improve upon the effluent quality from the reactor.

The ratio of TVS to TS decreased as the waste passed through the UASB reactor. That is the volatile content of the total solids, which gives a measure of the organic component decreased during the anaerobic digestion in the UASB reactor. The mean influent TVS/TS ratios were 60% and 64% respectively for the primary and faecal sludges. In the effluent, TVS/TS ratios were 39% and 42% for the primary and faecal sludges respectively. Thus in either experiment, there was a drop of just over 20% in the TVS/TS ratio as a result of the anaerobic digestion in the UASB reactor.

5.5.3 pH value and Stability

Anaerobic digestion like other biological processes are pH dependent. It is essential to maintain the optimal pH range conducive to the micro-organisms responsible for the anaerobic digestion processes. Maintenance of the optimal pH range ensures the stability of the process.

pH values measured during the experiment in Leeds using diluted primary sludge ranged from 6.9 to 7.4 for the influent and 6.5 to 7.1 for the effluent. These pH values are well within the optimal range for anaerobic digestion which is 6.3 – 7.8.

The effluent pH gradually increased towards the end of the experiment and in the latter stages the effluent pH showed little variation and mostly stayed in the neutral range (Figure 4.13). From the results it is clear that during the period of the experiment there was never the risk of souring of the reactor. van Haandel and Lettinga (1994) reported similar results for pH and concluded that, in general, souring of the contents of a UASB reactor is not a problem in anaerobic sewage treatment and that there is no need for chemical pH adjustment. In view of the results obtained in the Leeds experiment and the conclusion above by van Haandel and Lettinga (1994), the pH was not monitored during the experiment in Kumasi.

5.5.4 Nitrogen removal efficiencies

In the Leeds experiment free ammonia concentrations were measured in the influent and effluent samples. The influent ammonia concentrations ranged from 30 to 54 mg/l and had a mean of 41 mg/l. The effluent concentrations varied from 37 to 69 mg/l with an average of 53 mg/l. The percentage increase in the ammonia concentrations varied between 22% and 49% with a mean of 30%. The results show that the anaerobic treatment resulted in an increase in the ammonia concentration. These increases can be attributed to conversion of organic nitrogen present in the soluble organic compounds into ammonia nitrogen. van Haandel and Lettinga (1994) report similar results.

High concentrations of free ammonia can be inhibitory to the anaerobic treatment processes as revealed by several investigations. van Velsen (1979) and Hashimoto (1986) observed ammonia inhibition for unadapted methanogenic cultures to commence at concentrations of 1500 – 2500 mg-N/l. Koster and Lettinga (1994) reported ammonia inhibition to occur at 1700 mg-N/l at pH 7.5. Hashimoto (1986), and Angelidaki and Ahring (1993) demonstrated that by adaptation of the anaerobic process to ammonia, ammonia concentrations of up to 4000 mg-N/l can be tolerated. Many lower free ammonia concentrations (100 – 150 mg-N/l) have also been reported for initial inhibition of an unadapted process (Braun *et al.*, 1981; De Baere *et al.*, 1984).

The results indicate that once the primary and faecal sludges have been diluted the ammonia concentrations were lower than the inhibitory concentrations reported in the literature and therefore should not pose any problem to the anaerobic digestion process.

In the Kumasi experiment, organic nitrogen was measured for the influent and effluent. The influent organic nitrogen concentrations ranged from 205 to 520 mg-N/l and had a mean of 300 N-mg/l while the effluent organic nitrogen concentration varied between 130 to 325 mg-N/l with a mean of 185 mg/l. The results indicated that organic concentration decreased as a result of the anaerobic treatment. This is to be expected as in anaerobic digestion, organic compounds containing organic nitrogen are consumed by bacteria and converted into ammonia and amino acids. Organic nitrogen removal during the anaerobic digestion in the UASB reactor ranged from 27% to 45% with an average of 37%. The ammonia nitrogen produced from the conversion of the organic nitrogen is consumed by bacteria as they grow (Leslie Grady Jr. 1999) and the only measurable event with regards to ammonia production is the net accumulation or loss of ammonia in the medium. The reduction in the concentration of organic nitrogen ranged between 60 and 195 N-mg/l. This is amount that would have been converted into ammonia, part of which would be used by growing bacteria and the rest released. These concentrations are well below the inhibitory levels and free ammonia concentrations should not pose any problems to the anaerobic digestion process.

5.5.5 Biogas Production

Anaerobic digestion of organic material produces biogas, which consists mainly of methane and carbon dioxide. The biogas is produced during methanogenesis which is the last stage of the anaerobic digestion process. Of the four identified stages of the anaerobic digestion process (Figure 2-11), methanogenesis has been identified as the most critical for the whole process (Henze and Harremoes, 1983). The production of biogas can therefore be used to ascertain the progress of anaerobic digestion as was done in this research.

In this research, with the main focus being on determining the suitability of the UASB reactor in the treatment of faecal sludges, biogas production was observed and used as an indication of the progress of the digestion process. Volumes of biogas produced were noted. Due to the experimental design, although volumes of biogas collected were noted, the experimental set-ups were not checked for gas leakages and also the biogas collected were not analysed for gas composition. Much more biogas could have been collected in the UASB reactors if they had been properly sealed to prevent any biogas leakages. Furthermore as highlighted in section 2.10.7, a considerable portion of the biogas produced remains dissolved in the liquid phase and leaves the system in the effluent.

The calculated volumes of methane in the collected biogas and its COD equivalence are shown in Figure 4-14 and Table 4-9 for the experiments conducted in Leeds. Figures 4-15 and 4-16 also show the ratio of COD equivalence of the methane to the total COD removed, the volume of methane produced per kg COD removed and the volume of methane produced per kg TVS removed. The volume of methane in the collected biogas was assumed to be 70% of the volume measured. While the volume of methane in the collected biogas increased from the start of the experiment and tended to stabilised in the latter half of the experimental period, the ratio of the COD equivalence of methane to total COD removed was fairly constant throughout the experimental period with a mean of 14% and standard deviation of 2%. In terms of the volume of methane per kg COD removed and volume of methane per kg TVS removed, it averaged 5.4 l with a standard deviation of 0.8 l and 9.3 l with a standard deviation of 1.8 l respectively. The variation in the calculated volume of methane was higher for TVS than COD removed (Figure 4-16).

Figure 4-27 and Table 4-15 show the calculated volumes of methane in the collected biogas and its COD equivalence for the experiments conducted in Kumasi. Figures 4-28 and 4-29 also show the ratio of COD equivalence of the methane to the total COD removed, the volume of methane produced per kg COD removed and the volume of methane produced per kg TVS removed. The calculated volume of methane in the collected biogas increased first sharply at the beginning and then gradually in the latter half of the experiment. Just as in the case for primary sludge the ratio of the COD equivalence of methane to total COD removed was fairly constant throughout the experimental period with the same mean and standard deviation. This suggests that the COD associated with the methane produced may be a fairly constant ratio to the total COD removed. In terms of the volume of methane per kg COD removed and volume of methane per kg TVS removed, it averaged 5.5 l with a standard deviation of 0.6 l and 9.6 l with a standard deviation of 2.0 l respectively. Once again the variation in the calculated volume of methane was higher for TVS than COD removed (Figure 4-29).

The calculated volume of methane in the biogas collected were substantially less that the expected values. As shown in the example in section 2.10.7, at a temperature of 23 °C, the methane yield per kg COD is about 340 l/kg COD. Practically, van Haandel and Lettinga (1994) reports that the loses could be up to 50%. In this experiment, gas leakages were not checked. Biogas lost due to leakages were very substantial if the calculated volume of methane in the biogas collected is compared to the expected value.

Although substantial amount of biogas produced was lost, with a properly built UASB biogas leakages can be eliminated. The potential of biogas production from the UASB reactor is presented in detail in the design of a typical UASB plant to treat the faecal sludge from the city of Kumasi. In Ghana, the use of liquid petroleum gas (LPG) by households and motorists is on the increase as people try to cut down their energy bills. There will therefore be a commercial market for the gas produced if faecal sludge treatment systems are chosen for their biogas production potential.

5.6 Comparison of FSTP and UASB Reactor

In the following sections FSTP is compared to the UASB in a typical design for the city of Kumasi for faecal sludge (detail designs of the two systems are presented in the Appendix 1). The design is based on the population of Kumasi from the census conducted in 2000.

5.6.1 Land Area Requirements

Excluding the land area required for either sludge drying or co-composting of sludge with suitable organic bulking material, the total area required for the FSTP is 12 hectares, while the land area required for the UASB system is 7 hectares. That is, if the UASB system is used, this could result in a land savings of 5 hectares (about 42%). In an urban setting where land is scarce and expensive, this could represent a substantial financial saving in capital expenditure.

5.6.2 Detention Time and Effluent Quality

The total detention time for the FSTP is 35 days and the BOD of the effluent is 127 mg/l. With regards to the treatment system using UASB, the total detention time is 10.5 days and it produces an effluent BOD concentration of 37 mg/l. The UASB system therefore achieves a much higher effluent quality (70% better) than the FSTP in a much shorter time.

5.6.3 Methane Production

In addition to the above advantages, the methane produced in the UASB reactor can be collected and this would have commercial value. In the design presented in the Annex, the daily production of methane is between 330 m^3/d and 376 m^3/d. These value are very conservative and substantially lower as they were calculated using the methane production per kg COD and TVS removed obtained in the Kumasi experiments respectively. As mentioned in section 5.5.5 these values are very low due to the loss of biogas from the UASB reactor used for the experiment. Thus in full-scale UASB where the collection of biogas is well designed the amount of biogas collected could be substantial.

5.7 General

5.7.1 Possible Sources of Error

As with any experiment, there are potential sources of error that would be associated with the experiment. These errors could be inherent with the experimental procedures adopted or due to the planning of the experiment.

As mentioned in earlier sections, the production of biogas was observed and primarily used as an indication of the progress of the experiment. The reactor was not checked for any gas leakages and it is possible that some of the biogas produced was lost and therefore unaccounted for. This would undoubtedly affect the volume of biogas collected and consequently affect the calculated volume of methane used in the data analysis.

The volume of gas is affected by temperature at normal atmospheric pressures. The temperature of the biogas collected was assumed to be the same as that of the ambient air temperature. Although the biogas temperatures were not measured directly, it may not be much different from the ambient air temperatures and it will not significantly affect the volume of biogas collected.

There was no equipment for sampling and analysing the composition of the biogas collected. The percentage of methane in the collected biogas was thus assumed and used in all the calculations. In the estimation of the volume of methane, it was assumed that this was 70%. Obviously, different assumed values would yield different results.

5.7.2 Mass balance calculations

As mentioned in section 2.10.6, the mass balance for organic material within the UASB reactor is given by:

$$MS_{ti} = MS_{te} + MS_{tx} + MS_{tm} + MS_{to} + \Delta$$

where

MS_{ti} = daily mass of influent COD
MS_{te} = daily mass of effluent COD
MS_{tx} = daily mass of COD in the discharged sludge
MS_{tm} = daily mass of COD in produced methane
MS_{to} = daily mass of oxidised COD
Δ = net rate of generation in the control volume.

In a steady state condition, the net rate of generation in the control volume becomes zero. In the experimental determinations, the influent COD, effluent COD and the volume of biogas produced were measured. Thus the terms daily mass of influent COD, daily mass of effluent COD, and the daily mass of COD in produced methane in the mass balance equation can be calculated. The COD of discharged sludge was not determined except in one instance and also there was no equipment to analyse the composition of the biogas collected. Thus it was not possible to estimate the daily mass of COD in the discharged sludge and daily mass of oxidised COD due to inadequate and lack of data. With two terms not accounted for in the above equation, it is be extremely difficult to conduct any meaningful mass balance. The omissions of the two terms were partly due to the experimental design and partly due to the lack of equipment to analyse the composition of the biogas. Calculations involving methane production were therefore limited to production per kg COD and TVS removed in the UASB.

Chapter Six

6 Conclusions

The conclusions set out in this section are the result of the experimental study of the physico-chemical characteristics and anaerobic treatment of primary and faecal sludges using the upflow anaerobic sludge blanket (UASB) reactor.

1. The study indicates that the UASB process can be used for the anaerobic treatment of primary and faecal sludges under tropical climatic conditions employing an organic loading rate of up to 21.5 kg COD/m^3.d and hydraulic retention times of 10-12 hours with treatment efficiencies of over 70% on the basis of total COD reduction.

2. The data from the experimental study clearly show that a UASB reactor is able to handle varying influent loads and yet produce an effluent of fairly constant characteristics.

3. The removal efficiencies for total chemical oxygen demand (COD), total solids (TS), total volatile solids (TVS), total suspended solids (TSS) were high and are comparable with those reported in the literature for anaerobic treatment of sewage using the UASB reactor.

4. UASB reactors treating primary and faecal sludges will reach steady state conditions much earlier than similar reactors treating domestic sewage due to the high influent solids concentration from the outset of the treatment.

5. In general, the trend of the treatment performance of UASB reactors treating primary and faecal sludges increases steadily from the outset, unlike the reported experiences with UASB reactors treating domestic sewage in which the treatment efficiency first dipped before rising steadily due to the time required for the development of the sludge blanket.

6. Although the experimental study has demonstrated that high removal efficiencies of COD, TS, TVS, and TSS can be achieved at short retention times, in both experiments the effluent quality was not good enough to allow direct discharge into receiving water bodies. It will be necessary to apply some form of post-treatment to the effluent from the UASB reactor.

7. The concentrated nature of untreated primary and faecal sludges requires that it be diluted before the UASB can treat it effectively. In this experimental study a dilution ratio of 1:10 to 1:6 was found to be adequate.

8. Although ammonia and ammonium nitrogen concentrations are high in the raw faecal sludge, the concentrations decrease after dilution to levels that are not inhibitory to the performance of the USAB reactor in the anaerobic treatment of the faecal sludges.

9. From the design example presented in the Appendix 1, using the UASB followed by treatment ponds produces effluent of better quality in a much shorter time and using less land area than using solely the FSTP.

10. From the analysis of the gas data and the design example presented in Appendix 1, if the biogas loses are taken into account, then substantial volumes of methane can be produced from the UASB reactor and if properly harnessed, could represent a potential source of energy.

11. Some form of preliminary treatment is required before the anaerobic treatment of the faecal sludges using the UASB reactor. The preliminary treatment should include:

    a) **Screening:** Initial screening to remove all coarse solids such as sticks, rags, carrier bags and other large objects. These solids are predominantly non-putrescible and would lead to a sludge built up with the reactor resulting in decrease in the effective volume of the UASB rector.

b) **Equalisation and Storage Tank:** Due to the variability of the characteristics of faecal sludges and the manner of collection and transportation, an equalisation storage tank to equalise the variations in the pollutional strength of the faecal sludges and also ensure a continuous flow to the UASB reactor is essential.

c) **Grit Removal:** The washing and cleaning of public toilets result in grit being deposited in the toilets, especially aqua-privies and KVIPs. In order to prevent grit accumulation in the reactor it is essential to have a grit removal system after the dilution process.

Chapter Seven

7 Suggestions for further work

The present study has clearly demonstrated that it is possible to use the UASB reactor to anaerobically treat primary and faecal sludges when diluted appropriately. This holds an enormous potential for faecal wastes treatment in developing countries like Ghana where majority of sanitation facilities are on-site systems. However before the system can be adopted on a scale similar to its acceptance for the treatment of domestic sewage in tropical and sub-tropical climatic conditions, it will be essential to carry out more experimental studies. In this regard the following suggestions for further research work are made:

- Experiments similar to the one undertaken in this research should be undertaken over a much longer period to establish steady state conditions that will enable correct assessments of long-term treatment efficiencies, optimal loading rates, optimal hydraulic retention times, suitable dilution ratio, and gas production potentials among others.
- During such an experiment, efforts should be made to minimised, if not completely eliminate, the loss of biogas due to leakages and also an analysis of the biogas produced must be carried out to determine the percentages of methane and carbon dioxide in the biogas.
- Such long term experiments should provide the required information for developing design and operation/maintenance guidelines when using the UASB for the treatment of faecal sludges.
- Research efforts to find improvements in the dewatering properties of faecal sludges when subjected to anaerobic treatment in the UASB should be undertaken.

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APPENDIX 1: DESIGN OF FAECAL SLUDGE TREATMENT PLANT AND UPFLOW ANAEROBIC SLUDGE BLANKET REACTOR

In this section, a faecal sludge treatment plant (FSTP) and an upflow anaerobic sludge blanket (UASB) reactor are designed for the city of Kumasi, Ghana, using the current population. The purpose of the design is to compare the systems and find the merits and demerits of the USAB reactor. The design of the FSTP is based on the recommendations for preliminary design guidelines proposed by Heinss *et al* (1998).

1. Design Assumptions and Requirements

Population[1]		1,020,000
Sanitation service coverage[2]	%	Population
Public Toilets	38	387,600
Bucket Latrines	15	153,000
WCs with septic tanks	25	255,000
Sewerage	7	71,400
Pit latrines (KVIPs/Traditional pits)	7	71,400
"Free range" (no facility)	8	81,600

Per capita waste contribution[3]	
Public toilet and bucket latrine sludge	2 l/cap.d
Septage	1 l/cap.d
COD	
Public toilet and bucket latrine sludge[4]	86,700 mg/l
Septage[5]	8,500 mg/l
Total Solids	

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[1] Population of Kumasi - Year 2000 population census

[2] Kumasi Metropolitan Assembly development plan 1996 - 2000

[3] Table 2.1

[4] Table 5.1

[5] Table2.2

Public toilet and bucket latrine sludge[6]	86,700 mg/l
Septage[7]	8,500 mg/l
Mean minimum monthly temperature[8]	25 °C

2. Design Calculations for FSTP

Volumetric Load

Public toilet and bucket latrine sludge = population x per capita contribution

= (387,600 + 153,000) x 2 l/cap.d

= 1,081,200 l/d

= 1,081.2 m^3/d, **say 1,100 m^3/d**

Septage = 255,000 x 1 l/cap.d

= 255,000 l/d

= 255 m^3/d, **say 260 m^3/d**

V_{load} = 1,100 m^3/d + 260 m^3/d

= 1,360 m^3/d

COD Load

Public toilet and bucket latrine sludge = COD concentration x volumetric load

= 86,700 mg/l x 1,100 m^3/d

= 86.7 kg/m^3 x 1,100 m^3/d

= 95,370 kg COD/d

Septage = 8,500 mg/l x 260 m^3/d

= 8.5 kg/m^3 x 260 m^3/d

= 2,210 kg COD/d

Total COD_{Load} = 95,370 kg COD/d + 2,210 kg COD/d

= 97,580 kg COD/d

Total Solids Load

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[6] Table 5.1

[7] Table2.2

[8] Temperature values of 30 year average (1961-1990) in Kumasi from Ghana Meteorological Services Department Regional Office, Kumasi.

Public toilet and bucket latrine sludge = TS concentration x volumetric load
= 55,700 mg/l x 1,100 m^3/d
= 55.7 kg/m^3 x 1,100 m^3/d
= 61,270 kg TS/d

Septage = 14,000 mg/l x 260 m^3/d
= 14 kg/m^3 x 260 m^3/d
= 3,640 kg TS/d

TS_{Load} = 61,270 kg TS/d + 3,640 kg TS/d
= 64,910 kg TS/d

Size of Sedimentation/Thickening Tanks (Figure I)

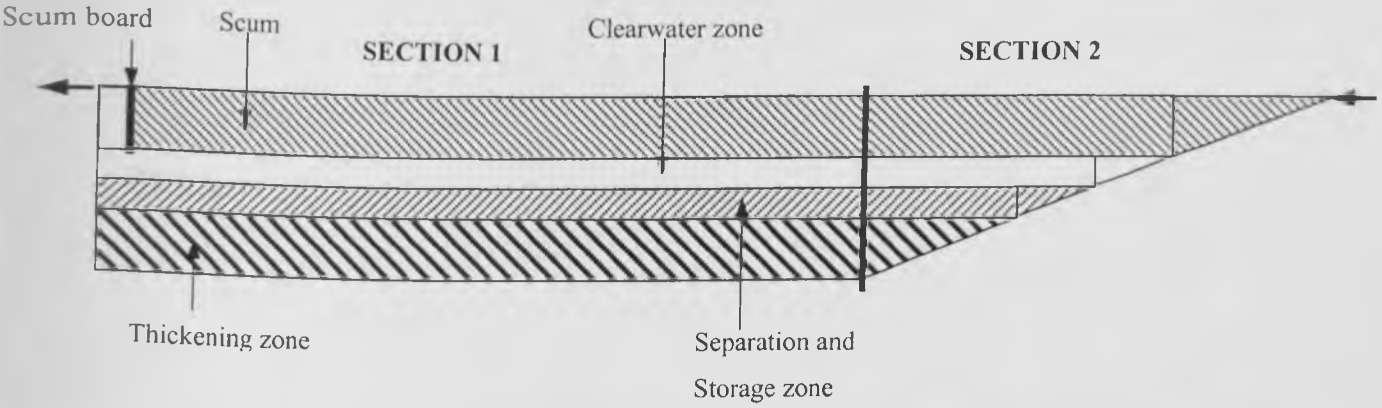

Figure I: **Sedimentation/Thickening Tank with Four Distinct Layers of Separated Solids** (Source: Heinss *et al,* 1998)

Table I: **Solids Concentration Attained in Full-Scale Settling/Thickening Tanks in Accra, Ghana** (Source: Heinss *et al*, 1998)

Zone	**Depth from the surface (m)**	**Solids concentration (kg/m^3)**
Scum	0 – 0.8	160
Clearwater zone	0.8 – 1.3	4
Separation and storage zone	1.3 – 1.8	60
Thickening zone	> 1.8	140

Assume minimal nominal liquid retention time = 4 h

V_{load} = 1,360 m^3/d

Assume 2 tanks are to be built in parallel

Hourly influent = (1,360/2) m^3 / 8 hours

= 85 m^3/h

Volume required for clearwater zone = 85 m^3/h x 4 h

= 340 m^3

Assume tank width = 10 m

Length of Clearwater zone (section 1) = 340 m^3 /(10 m x 0.5 m)

= 68 m **say 70 *m***

Assume slope of ramp = 20 °

Length of section 2 = 8.8 m, **say 10 m.**

Assume effective tank depth, d = 4 m

Volume of section 1 of tank:

V_{S1} = 70 m x 10 m x 4.0 m

= 2,800 m^3

Volume of section 2 of tank:

V_{S2} = (10 m x 10 m x 4.0 m) x 0.5 = 200 m^3

Volume of tank:

V = 2,800 m^3 + 200 m^3

= 3,000 m^3

V_{S1} accounts for 93% of the total volume of tank.

Sludge mass stored in section 1:

V_{S1} = (Sludge mass to stored/solids concentration per m^2) x depth

2,800 m^3 = [(Sludge mass stored) kg/(0.8m x 160 kg/m^3 + 0.5m x 60 kg/m^3 + 2.2m x 140 kg/m^3)] x 4.0m

Sludge mass stored = [2,800 m^3 / 4.0 m] x (0.8m x 160 kg/m^3 + 0.5m x 60 kg/m^3 + 2.2m x 140 kg/m^3)]

= 326,200 kg

Total sludge mass = 326,200 kg / 0.93
= 350,750 kg

Assume solids removal = 70%

Tank loading period = Sludge mass stored / (percent removal x TS load)
= 350,750 kg / (0.7 x 0.5 x 64,910 kg/d)
= 15.4 days, **say 14 days, i.e. 2 weeks**

Heinss *et al.* (1998) recommends 4 weeks resting period before emptying tank of accumulated sludge and reloading. Thus operationally, 6 tanks would be constructed with two in use at the same time with a loading period of 14 days and a resting period of 28 days. Hence once every 4 weeks a volume of approximately 6,000 m^3 of separated solids with TS ≥ 14% must be removed and further treated. Currently this is done by co-composting with suitable organic bulking material such as domestic refuse, sawdust or woodchips.

Total land area required for sedimentation/thickening tanks:
A_{total} = (70 m + 10 m) x 10 m x 6 = 4,800 m^2

<u>Anaerobic Ponds</u>

Assume COD removal in tank = 50%
COD load to anaerobic pond = 0.5 x 97,580 kg COD/d
= 48,790 kg COD/d

BOD[9] load to anaerobic pond = 13,186 kg BOD/d, **say 13,200 kg BOD/d.**

Assume volumetric BOD loading rate = 350 g BOD/m^3.d

<u>Check for Ammonia Toxicity</u>

Ammonium nitrogen concentration[10]
Septage = 200 mg/l
Public toilet and bucket latrine sludge = 2,500 mg/l

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[9] Table 4.1: COD/BOD = 3.7:1

[10] Heinss *et al* (1998)

Average influent concentration = {(255,000 l/d x 200 mg/l) +
(1,081,200 l/d x 2,500 mg/l)}/
(255,000 l/d + 1,081,200 l/d)
= 2060 mg/l

Assume ammonium nitrogen losses in tank = 5%
Ammonium nitrogen influent to pond system = 0.95 x 2,060 mg/l
= 1,960 mg/l

Using Figure 2-14, for a maximum pH = 8.5 and average temperature = 25 °C
Ammonia nitrogen concentration = 18% of 1,960 mg/l
= 353 mg/l.

This concentration is above the threshold for ammonia toxicity to anaerobic bacteria (100 mg/l). Hence the effluent from the pond is to be diluted 1:4 to ensure that the threshold concentration is not exceeded.

Sizing of anaerobic pond

Assume 5% loss in settling-thickening tank, thus volumetric load after dilution
V^{*}_{load} = 0.95 x 1,360 m^3/d x 4
= 5,168 m^3/d, **say 5,200 m^3/d**

BOD load = 13,200,000 g BOD/d
BOD volumetric load = 350 g/m^3.d
$V_{an\ p}$ = [13,200,000 g BOD/d] / 350 g/m^3.d
= 37,714 m^3

Retention time = (37,714 m^3) / (5,200 m^3/d)
= 7.2 d, **OK**

Assume anaerobic pond depth = 3 m
Anaerobic pond area = 37,714 m^3 / 3m
= 12,571 m^2

Choosing a width to length ratio of 3:1, the pond will have the following dimensions:
Length = 195 m
Width = 65 m

Assume 4 anaerobic ponds in parallel, then pond dimensions would be:
Length = 99 m
Width = 33 m

Total pond area = 4 x 33 m x 99 m = 13,068 m^2

Assume BOD removal in anaerobic pond = 75%

BOD load in anaerobic pond effluent = [0.25 x 13,200 kg BOD/d]

= 3,300 kg BOD/d

BOD concentration = [3,300 kg BOD/d] / 5,200 m^3/d = 635 mg/l

Facultative Ponds

Using Mara's (1987) tentative global equation:

λ_s = 350 (1.107 – 0.002T)$^{T-25}$

where λ_s = BOD surface area loading (kg/ha.d)

T = temperature (°C)

At T = 25 °C, λ_s = 350 kg/ha.d

$A_{\text{fac. pond}}$ = [3,300 kg BOD/d] / 350 kg/ha.d

= 9.428 ha

= 94, 280 m^2

Assume a width to length ratio of 1:6 and that 8 ponds are to be constructed, the pond dimensions would be:

Length = 270 m

Width = 45 m

Total pond area = 8 x 270 m x 45 m = 97,200 m^2

Using a depth of 1.5 m

Volume of pond, $V_{\text{fac. pond}}$ = 1.5 m x 97,200 m^2 = 145,800 m^3

Retention time = 145,800 m^3 / 5,200 m^3/d = 28 days,

Assume 80% BOD removal in facultative pond

Effluent BOD = 0.2 x 635 mg/l = 127 mg/l

Total land area required = 4,800 m^2 + 13,068 m^2 + 97,200 m^2

= 115, 068 m^2

≈ 12 ha

(Land area excludes area required for pond embankments, access roads and drying of sludge)

3. Design Calculations for UASB

The design of the UASB is based on organic loading rather than hydraulic loading because of the concentrated nature of the wastes. The necessary design equations used are:

$$l_o = \frac{L_o}{V_r} = \frac{Q_i S_{ti}}{V_r} = \frac{S_{ti}}{(HRT)}$$

where

l_o = applied specific COD load (kg COD m^{-3} d^{-1})
L_o = organic (COD) load (kg COD d^{-1})
V_r = volume of the reactor (m^3);
Q_i = average wastewater flow (m^3/d);
S_{ti} = influent organic material (COD) concentration (kg/m^3)
(HRT) = hydraulic retention time (d)

Volumetric Load

Public toilet and bucket latrine sludge = population x per capita contribution
= (387,600 + 153,000) x 2 l/cap.d
= 1,081,200 l/d
= 1,081.2 m^3/d, **say 1,100 m^3/d**

Septage = 255,000 x 1 l/cap.d
= 255,000 l/d
= 255 m^3/d, **say 260 m^3/d**

Total volume of faecal sludge = 1,100 m^3/d + 260 m^3/d
= 1,360 m^3/d

Assume dilution ratio of 1:8

Volumetric load, V_{load} = 10,880 m^3/d

COD Load

Public toilet and bucket latrine sludge = COD concentration x volumetric load
= 86,700 mg/l x 1,100 m^3/d
= 86.7 kg/m^3 x 1,100 m^3/d
= 95,370 kg COD/d

Septage = 8,500 mg/l x 260 m^3/d
= 8.5 kg/m^3 x 260 m^3/d
= 2,210 kg COD/d

Total COD_{Load} = 95,370 kg COD/d + 2,210 kg COD/d
= 97,580 kg COD/d

Influent COD concentration = [97,580 kg COD/d] / 10,880 m^3/d
≈ 9.0 kg/m^3

Assume applied specific COD load l_o = 17 kg COD/m^3.d [11]
Volume of reactor, V_r = L_o / l_o
= [97,580 kg COD/d] / 17 kg COD/m^3.d
= 5,740 m^3

A number of reactors would have to be constructed to accommodate this volume. Rectangular section reactors would be used for reasons stated in section 2.10.3.2

Use a reactor of dimensions (length x width x height) 24m x 10m x 6m
Volume of reactor = 1,440 m^3.
Hence four (4) reactors would be used with total volume = 5,760 m^3

Daily flow to each reactor = [10,880 m^3/d] / 4
= 2,720 m^3/d

Organic load to each reactor = 9.0 kg/m^3 x 2,720 m^3/d

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[11] Average applied specific COD load in the Kumasi experiment.

= 24,480 kg COD/d

Applied specific COD load l_o = [24,480 kg COD/d] / 1,440 m^3.

= 17 kg COD/m^3.d (< 20 kg COD/m^3.d) **OK.**

Total area cross-sectional area of 4 reactors = 4 x 24m x 10m

= 960 m^2

Hydraulic retention time, *HRT*, = S_{ti} / l_o

= [9.0 kg/m^3] / [17 kg COD/m^3.d]

= 0.53 d

= **12.7 h**

Upflow velocity, v_i (m/h) = *H / (HRT)*

= 6m /12.7h

= **0.47 m/h (< 1 m/h) OK.**

Storage/Flow Equalisation Tank

Due to the variability of the characteristics of faecal sludges and the intermittent manner of collection and transportation, an equalisation/storage tank is required to even out the variations in the pollutional strength of the faecal sludges and also ensure a continuous flow to the UASB reactor.

Volumetric Load = 10,880 m^3/d (delivered in 8 hours)

i.e. Influent flowrate into tank = 1360 m^3/h

Flowrate from storage tank = 454 m^3/h

Daily storage volume required = 10,880 m^3 - [454 m^3/h x 8 h]

= 7,248 m^3, **say 7,250 m^3**

Assume depth of tank = 2.5 m

Area of tank = 2,900 m^2

Using a width to length ratio of 1:4, the tank would have dimensions 27m x 108m.

Table: II Assumed Treatment Performance for UASB Reactor [12]

Parameter	Treatment Performance
COD removal	70%
TS removal	60%
TVS removal	75%
Organic nitrogen removal	35%
Volume of methane per g COD removal	5.5 ml
Volume of methane per g TVS removal	9.6 ml

Influent Characteristics after dilution

COD

COD = 9 kg/m^3

Total Solids

Public toilet and bucket latrine sludge = TS concentration x volumetric load
= 55,700 mg/l x 1,100 m^3/d
= 55.7 kg/m^3 x 1,100 m^3/d
= 61,270 kg TS/d

Septage = 14,000 mg/l x 260 m^3/d
= 14 kg/m^3 x 260 m^3/d
= 3,640 kg TS/d

TS_{Load} = 61,270 kg TS/d + 3,640 kg TS/d
= 64,910 kg TS/d

TS concentration = [64,910 kg TS/d] / [10,880 m^3/d]
≈ 6.0 kg TS/m^3, **i.e. 6,000 g/m^3 or mg/l**

Total Volatile Solids

Public toilet and bucket latrine sludge = TVS concentration x volumetric load
= 39,700 mg/l x 1,100 m^3/d
= 39.7 kg/m^3 x 1,100 m^3/d
= 43,670 kg TVS/d

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[12] Values from treatment performance of UASB reactor during the Kumasi experiments

Septage = 8,400 mg/l x 260 m^3/d
= 8.4 kg/m^3 x 260 m^3/d
= 2,184 kg TVS/d

TVS $_{Load}$ = 43,670 kg TVS/d + 2,184 kg TVS/d
= 45,854 kg TVS/d

TVS concentration = [45,854 kg TVS/d] / [10,880 m^3/d]
≈ 4.2 kg TVS/m^3, **i.e. 4,200 g/m^3 or mg/l**

Organic nitrogen

Ammonium nitrogen concentration before dilution [13]

Septage = 200 mg/l
Public toilet and bucket latrine sludge = 2,500 mg/l

Average influent concentration = {260 m^3/d x 0.2 kg/m^3 + 1,100 m^3/d x 2.5 kg/m^3} / 10,880 m^3/d
= 0.26 kg/m^3; **i.e. 260 g/m^3 or mg/l**

Using the above treatment performances (Table II), the effluent characteristics from the UASB reactor would be:

Effluent COD = 0.3 x 9,000 g/m^3
= 2,700 g/m^3
Effluent BOD[14] = 730 g/m^3

Effluent TS = 0.4 x 6,000 g/m^3
= 2,400 g/m^3

Effluent TVS = 0.25 x 4,200 g/m^3
= 1,050 g/m^3

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[13] Heinss *et al* (1998)

[14] Table 4.1 – COD:BOD = 3.7:1

Sizing of anaerobic pond

V_{load} = **10,880 m^3/d**

BOD load = 730 g/m^3 x 10,880 m^3/d

= 7,942,400 g BOD/d

BOD volumetric load = 350 $g/m^3.d$

$V_{an.\,p}$ = [7,942,400 g BOD/d] / 350 $g/m^3.d$

= 22,693 m^3

Retention time = (22,693 m^3) / (10,880 m^3/d)

= 2.0 d, **OK**

Assume anaerobic pond depth = 3 m

Anaerobic pond area = 22,693 m^3 / 3m

= 7,565 m^2

Choosing a width to length ratio of 3:1, and having 3 ponds in parallel will have the following dimensions:

Length = 87 m

Width = 29 m

Total pond area = 3 x 29 m x 87 m = 7,569 m^2

Assume BOD removal in anaerobic pond = 75%

BOD load in anaerobic pond effluent = [0.25 x 7,942,400 g BOD/d]

= 1,990 kg BOD/d

BOD concentration = [1,990,000 g BOD/d] / 10,880 m^3/d = 183 g/m^3

Facultative Ponds

BOD load = 1,990 kg BOD/d

Using Mara's (1987) tentative global equation:

$\lambda_s = 350\,(1.107 - 0.002T)^{T-25}$

where λ_s = BOD surface area loading (kg/ha.d)

T = temperature (°C)

At T = 25 °C, λ_s = 350 kg/ha.d

$A_{fac.\,pond}$ = [1,990 kg BOD/d] / 350 kg/ha.d

= 5.685 ha

= 56, 850 m^2

Assume a width to length ratio of 1:5 and that 6 ponds are to be constructed, the pond dimensions would be:

Length = 220 m

Width = 44 m

Total pond area = 6 x 220 m x 44 m = 58,080 m^2

Using a depth of 1.5 m

Volume of pond, $V_{fac.\ pond}$ = 1.5 m x 58,080 m^2 = 87,120 m^3

Retention time = 87,120 m^3 / 10,880 m^3/d = 8 days,

Assume 80% BOD removal in facultative pond

Effluent BOD = 0.2 x 183 mg/l = 37 mg/l

Total land area required = 960 + 2,900 + 7,569 + 58,080

= 69,509 m^2

≈ 7 ha.

(Land area excludes area required for pond embankments, access roads and sludge drying beds)

Methane Production

COD removed = 97,580 kg COD/d x 0.7

= 68,306 kg COD/d

= 68, 306,000 g COD/d

Volume of methane = 5.5 ml/g COD x 68,306,000 g COD/d

= 375,683,000ml/d

= 375,683 l/d

≈ **376 m^3/d**

Using the total volatile solids, the volume of methane produced is:

TVS removed = 45,854 kg TVS/d x 0.75
= 34,390 kg TVS/d
= 34, 390,000 g TVS/d

Volume of methane = 9.6 ml/g TVS x 34,390,000 g COD/d
= 330,114,000ml/d
= 330,114 l/d
≈ **330 m^3/d**