ANALYSIS OF A PILOT-SCALE ANAEROBIC BAFFLED REACTOR TREATING DOMESTIC WASTEWATER

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For Chris, Chris and Chris, but not necessarily in that order

and

my four boys

DECLARATIONS

I, Katherine Maria Foxon, declare that unless indicated, this thesis is my own work and that it has not been submitted, in whole or in part, for a degree at another University or Institution.

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Disclaimer

Use of quotations from Mr Donald Rumsfeld, US Secretary of Defence (2001 to 2006) does not indicate agreement with his views or those of any agency of the US state government... *I merely found his comments related to wars with controversial origins to be equally applicable to the undertaking of PhD research, and the writing of this thesis!*

ANALYSIS OF A PILOT-SCALE ANAEROBIC BAFFLED REACTOR TREATING DOMESTIC WASTEWATER

ABSTRACT

This thesis presents a chemical, microbiological and mathematical analysis of an anaerobic baffled reactor (ABR) treating domestic wastewater. The purpose of this study was to gain an understanding of the mechanisms of treatment of domestic wastewater in an ABR at pilot-scale, and to use this understanding to develop some guidelines for the design of ABR technology for the anaerobic treatment or pre-treatment of domestic wastewater. Previous research has been undertaken on ABR technology, but no detailed studies of the performance of an ABR on domestic wastewater at pilot-scale have been reported.

In this thesis, operating data from a 3 000 ℓ pilot-scale ABR are presented and analysed. Two hypotheses were proposed: that (i) the baffled design of the reactor would facilitate phase separation whereby acidogenic and methanogenic processes predominate in different physical locations in the reactor; and (ii) the critical design parameter is the applied hydraulic retention time.

The principle findings of this research were:

- The pilot-scale ABR functioned as a solids retention device. Particulate material was retained through settling in the first compartment, forming a gel-like matrix. Reduction of solids occurred through anaerobic conversion to CH_4 and CO_2.

- Partial phase separation of acidogenic and methanogenic communities was observed.

- The major factor that controlled biomass washout rate and therefore reactor performance was upflow velocity in each compartment. At higher upflow velocities, slow growing micro-organisms failed to establish, resulting in increased solids accumulation rates, while at lower upflow velocities, stable digestion proceeded.

- Relatively poor treatment rates were obtained due to the low inherent alkalinity of waters in eThekwini municipality resulting in low operating pH values.

- Insufficient pathogen reduction was observed indicating that post-treatment of effluent would be required.

It was concluded that the benefit of the baffled design was related to the system's solids retention characteristics and that the critical design parameters for an ABR domestic wastewater treatment unit were compartment upflow velocity and applied hydraulic retention time.

ANALYSIS OF A PILOT-SCALE ANAEROBIC BAFFLED REACTOR TREATING DOMESTIC WASTEWATER

LIST OF TABLES

LIST OF FIGURES

LIST OF ABBREVIATIONS

A-HRT	Applied hydraulic retention time
ABR	Anaerobic baffled reactor
ADM1	Anaerobic digestion model No 1
ADMI	American Dye Manufacturers Institute (measures colour intensity of dye)
BOD	Biological oxygen demand
BORDA	Bremen Overseas Research and Development Association
CFD	Computational fluid dynamics
cfu	Colony forming units
COD	Chemical oxygen demand
CSTR	Continuous stirred tank reactor
DAPI	4'6-diamidino-2-phenylindole
DEWATS	Decentralised wastewater treatment systems
DNA	Deoxy ribonucleic acid
EGSB	Expanded granular sludge bed reactor
FISH	Fluorescent in-situ hybridisation
HPLC	High performance liquid chromatograph
HRT	Hydraulic retention time
PCR	Polymerase chain reaction
PE	population equivalent
pfu	Plaque forming units
PI	Proportional + integral (control)
PLC	Programmable logic controller
PVC	Poly vinyl chloride
RBCOD	Readily biodegradable COD
RNA	Ribonucleic acid
SEM	Scanning electron microscopy
SS	Suspended solids
T-HRT	Target hydraulic retention time
TKN	Total Kjeldahl nitrogen
TP	Total phosphorous
TSS	Total suspended solids
TS	Total solids
UASB	Upflow anaerobic sludge blanket (reactor)
USEPA	United States Environmental Protection Agency

VFA	Volatile fatty acids
WRC	Water Research Commission
WWTP	wastewater treatment plant

1 INTRODUCTION

The work presented in this PhD thesis was part of a Water Research Commission (WRC) Project (K5/1248 ***The anaerobic baffled reactor for sanitation in dense peri-urban areas***) that sought to provide a detailed characterisation of the performance of an anaerobic baffled reactor (ABR) treating domestic wastewater. The purpose of the research was to assess the applicability of ABR technology in the provision of sanitation. This chapter describes the world-wide need for improved sanitation, background to the project, the objectives of the project and the project methodology.

1.1 SANITATION IN SOUTH AFRICA

In 2000, the United Nations Millennium Summit adopted the Millennium Development Goals which aim to improve the quality of life of the world's poorest and most vulnerable people and to address issues relating to environmental sustainability (United Nations, 2007b). The goals provide a framework for development and target dates for their achievement. These have become the reference against which improvements in the human condition in developing countries are measured and tracked. Goal 7 focuses on environmental sustainability, while Target 10, within Goal 7, aims … *to halve by 2015, the proportion of people without sustainable access to safe drinking water and* ***basic sanitation*** ... (United Nations, 2007b)

According to the 2006 Millennium Development Goals Report, 32% of Sub-Saharan Africans had access to improved sanitation in 2004. The United Nations has constructed a list of sanitation systems/services that may be regarded as improved or alternatively, not improved (**Table 1.1**)

Table 1.1: List of sanitation systems categorised as improved or unimproved (United Nations, 2007a)

Improved Sanitation	Unimproved Sanitation
Flush or pour-flush toilet piped to sewer	Flush or pour-flush toilet piped to road/ground/river
Septic tank	Pit latrine without slab or open pit
Ventilated improved pit latrine	Bucket toilet
Pit latrine with slab	Hanging toilet or hanging latrine
Composting Toilet	No facilities (bush/ field)

Improved Sanitation is roughly defined as a sanitation facility that hygienically separates human excreta from people, animals and insects at a household level (United Nations Development Group, 2003). The 2005 Millennium Development Project report (United Nations, 2005) reported that 26% of rural dwellers and 55% of urban dwellers in Sub-Saharan Africa had access to acceptable levels of sanitation in 2004. In 2001, the Statistics South Africa 2001 Census (StatsSA, 2005) reported that 13.6% of households did not have a toilet (chemical, flush or pit toilet). The South African Minister of Water Affairs and Forestry in her 2007 budget speech in May 2007 stated that 27% South Africans still do not have basic sanitation and must be serviced by March 2010, and that the building of 3.2

million toilets was required in the intervening years (Hendriks, 2007). South Africa is committed to eradicating its water supply and sanitation backlogs by 2008 and 2010 respectively, both in keeping with (and in fact exceeding) the requirements of the Millennium Development Goals, and in the upholding of the South African constitution, which states that *...Everyone has the right - to ... an environment that is not harmful to their health or well-being;...Everyone has the right to have access to - ... sufficient food and water*.

The primary function of a sanitation system is to create a physical barrier between humans and human excrement to prevent the transmission of pathogens via the faecal-oral route (Goldstein, 1999). In densely populated communities, a further and equally important objective is to prevent contamination of the environment with large amounts of pollutants including organics, nitrogen and phosphorus, which lead to eutrophication of water resources and disruption of natural eco-systems. Water-borne sanitation where diluted toilet contents are collected and transported to an activated sludge treatment system is a high-technology solution, however, it has large infrastructural, operational, maintenance and environmental costs which are economically, environmentally and socially unsustainable in many communities (Rockstrom et al., 2005; Foxon et al., 2006). In a South African context, the provision of waterborne sanitation systems, and full pressure water supply required for the operation thereof, is unsustainable in terms of the availability and cost of the water and infrastructure required for a universal waterborne sanitation system, the necessary financial resources, and the human capacity to operate and maintain these systems on a continuous basis (Hanekom, 2006). On an international level, there is an increasing realisation that, particularly in water-scarce and arid countries, the implementation of centralised wastewater treatment facilities supplied by a comprehensive sewer system through which wastes are transported by large volumes of drinking water is illogical and environmentally unsound (van Lier and Lettinga, 1999).

There is thus both a national and international drive to provide sustainable alternative water and sanitation services to millions of South Africans in the course of this decade, and an unsurpassed opportunity for innovation in the sanitation sector.

1.2 THE ANAEROBIC BAFFLED REACTOR PROJECT

The motivation for this project was that, in certain instances, communities may be supplied with piped potable water, but be located geographically in such a way that sewerage could not be removed by trunk sewer to central wastewater treatment facilities. Because of the cost of water to low-income householders, household water usage may be low and therefore the wastewater produced from these areas may be concentrated. It was proposed that research should be undertaken to identify potential decentralised waterborne wastewater treatment technologies that could function within this context (Foxon et al., 2006).

Anaerobic digestion is an area of biotechnology that is growing at a rapid rate. For many waste treatment applications it is becoming the technology of choice through its ability to produce renewable energy in the form of biogas, by reduction of greenhouse gases produced by other treatment processes and by the use of non-renewable fuels that biogas can replace; and by diverting organic waste from landfill and incineration (DEFRA, 2007).

Anaerobic digestion is a reasonable choice of bioprocess for on-site and decentralised sanitation since energy is not required for aeration and most importantly because of the low excess sludge production

obtained from anaerobic systems relative to aerobic processes (Speece, 1996 p. 11). In fact, these advantages apply equally to well-operated septic tank and pit latrine systems.

In 1999, the ABR was identified as a possible on-site treatment option in peri-urban settlements (WRC project K5/1248 The anaerobic baffled reactor for sanitation in dense peri-urban areas, Foxon et al., 2006).

The ABR may be described as a baffled septic tank in which internal baffles divide the reactor into different compartments and flow zones (**Figure 1.1**).

ABR technology has been used in the treatment of a variety of wastewater types; the hydraulic design ensures good solids retention and good contact between biomass and organic substrate in the wastewater and therefore good organic removal rates (Barber and Stuckey, 1999). The Pollution Research Group (University of Natal / University of KwaZulu-Natal) had undertaken an earlier project investigating anaerobic baffled reactor technology in another WRC project no. 853 *The assessment of a baffled compartmentalised anaerobic digester for the treatment of high-strength or toxic organic industrial effluents* and had found the technology to provide good treatment rates, and higher tolerance of hydraulic and organic shock loads than unbaffled anaerobic reactors for high strength applications (Bell, 2000; Bell and Buckley, 2003). Work by Bell (2002) indicated that the exceptional performance of the ABR was due, in part, to the development of specialised micro-organism populations in each compartment of the ABR.

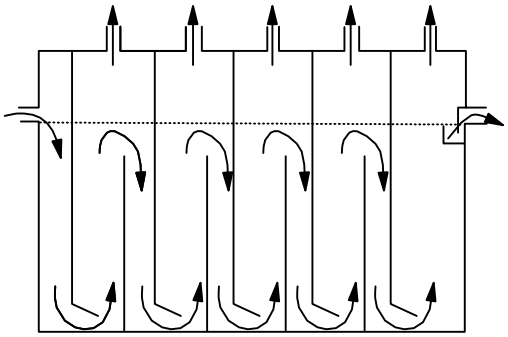

Figure 1.1: Graphical representation of a generic 5-compartment anaerobic baffled reactor

Domestic wastewater is regarded as low strength (relative to other anaerobic applications) and contains biodegradable material that may be particulate, colloidal or soluble. Most previous ABR research had been undertaken on high strength wastewaters with soluble biodegradable material (Barber and Stuckey, 1999). Research on domestic wastewater has been limited with most work undertaken on a laboratory scale using artificial wastewater, or at full scale with limited investigation of internal reactor dynamics; thus there was a need for further research into the performance of ABR technology on real domestic wastewater.

The WRC project 1248 identified the following advantages of using ABR technology for sanitation (Foxon et al., 2006):

- There is potential for suitably disinfected effluent from an anaerobic system to be recovered and reused in selected activities, thereby ensuring water conservation on a household or community level.

- From an environmental life cycle assessment perspective, no electricity is expended on removing nutrients as in conventional wastewater treatment. Further, these nutrients could enhance the reuse value of the effluent in agriculture. This translates to a double advantage: firstly there are no adverse environmental impacts associated with the use of electricity in treatment processes; secondly, the generation of a nutrient-rich effluent reduces the requirement for chemical fertilisers in crop production and the environmental damage resulting from the extraction, refinement, use and dissipation of chemical fertilisers.

- Anaerobic conversion of organic matter from wastewater results in the production of CH_4 gas which could be harvested as an energy source.

- The baffled configuration of the ABR has certain hydraulic and biochemical advantages over other anaerobic digester designs (**Section 2.5**).

However, anaerobic technologies have limited ability to remove nutrients and pathogens from wastewater. Consequently, it is necessary to carefully characterise the treatment performance of the ABR used in the treatment of domestic wastewater, to identify critical design characteristics and to identify appropriate applications for the technology before implementation.

The performance of the ABR in various applications has been studied on a laboratory-, pilot- and in a few cases, full-scale. A detailed study combining chemical, biochemical and microbiological performance on a pilot- or full-scale ABR treating real domestic wastewater has not previously been undertaken.

1.3 HISTORY OF THE ABR PROJECT AT UKZN

The Pollution Research Group in the School of Chemical Engineering of the former University of Natal, and now of the University of KwaZulu-Natal has been involved in research into anaerobic baffled reactor technology for more than 10 years.

Some of the research undertaken in the study of the pilot-scale ABR system formed part of Master of Science and Master of Science in Engineering dissertations. These are listed below. The main conclusions of these projects are described in **Section** 2.5.3.

- Bell (2007) used preliminary results from the pilot-scale ABR in the final report for WRC project K5/853 *The assessment of a baffled compartmentalised anaerobic digester for the treatment of high-strength or toxic organic industrial effluents*

- Dama (MScEng, (in preparation)) undertook a computational fluid dynamics study on water flow patterns around hanging and standing baffles in an ABR for different baffle configurations. This study assisted in the design of the pilot-scale ABR that was used in all the UKZN research reported hereafter.

- Mtembu (MScEng, 2005) monitored the operation of the pilot-scale ABR for a period of two years ending in mid 2003 devising practical solutions to operating problems with the pilot rig, and measuring reactor performance indicators.

- Lalbahadur (MTech, 2005) performed microbiological analyses on samples from the pilot-scale ABR in 2003 to characterise the microbial communities for each compartment.

- Pillay (MSc, 2006) made scanning electron microscopic examinations of compartment samples, particularly of sludge granules in 2003 and 2004, and measured levels of pathogen indicator organisms in feed, outflow and within compartments of the pilot-scale ABR.

This thesis collates and analyses all the data gathered from the pilot-scale ABR project including these sub-projects in order to develop a basis for reactor design and operating guidelines. **Figure 1.2** shows involvement of research students and assistants in research into ABR technology. It also shows the change in identity of the host institution from the University of Natal to the University of KwaZulu-Natal following an institutional merger of the former Universities of Natal and Durban Westville at the beginning of 2004.

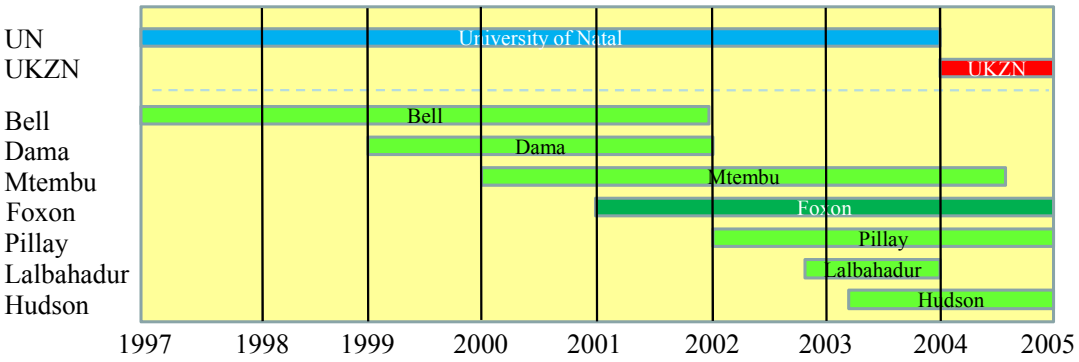

Figure 1.2: Participation of researchers in research into ABR technology. The former University of Natal merged with the University of Durban Westville to become the University of KwaZulu-Natal part way through this research.

1.4 OBJECTIVES OF THIS STUDY

The aims of this research were:

- To investigate the performance of a pilot-scale ABR in the treatment of wastewater of domestic origin and understand the mechanisms of treatment therein

- To identify the critical parameters in the design of an ABR sanitation system

- To determine whether the baffled design has any significant benefits over other anaerobic technologies in the treatment of domestic wastewater

- To develop a dynamic mathematical model of the biochemical processes in the ABR to assist in reactor design

1.5 HYPOTHESES

Two hypotheses were proposed:

- It is widely reported that division of anaerobic systems into an initial acidogenic zone and subsequent methanogenic zone (stage or phase separation) improves digester stability and overall treatment rates. Phase separation has been observed in baffled reactors(Bell, 2002) and it was hypothesised that phase separation in an ABR treating sewage is a benefit of the design over a single phase system, by allowing development of acidogenic and methanogenic zones.

- Hydraulic retention time (HRT) describes the amount of time fluid spends inside a reactor, and is a function of the reactor volume and fluid flow rate. It was expected that the critical parameter controlling effluent quality and sludge digestion rates in an ABR treating sewage was the applied hydraulic retention time (A-HRT), defined as reactor volume/flow rate i.e. the relationship between wastewater flow rate and reactor volume. The hypothesis was that by controlling the flow rate to an ABR of fixed volume, it would be possible to achieve specific COD reduction targets, with a target A-HRT time in the region of 20 h.

It was proposed that operation and analysis of chemical and microbiological data from a pilot-scale ABR treating domestic wastewater at a municipal wastewater treatment plant would allow these hypotheses to be tested. Specifically, examination of pH profiles in an operating ABR would be employed to understand the extent and significance of phase separation, while the effect of variations in feed flow rate on system performance indicators such as COD removal and solids accumulation would be investigated to determine the effect of A-HRT.

1.6 PROJECT TIME LINE

This thesis considers results obtained from operation of a 3 000 ℓ pilot-scale ABR operated at Umbilo wastewater treatment plant (WWTP) and Kingsburgh WWTP. **Figure 1.3** is a Gantt chart showing the timing of the different phases of operation of the pilot-scale ABR that were considered in this thesis.

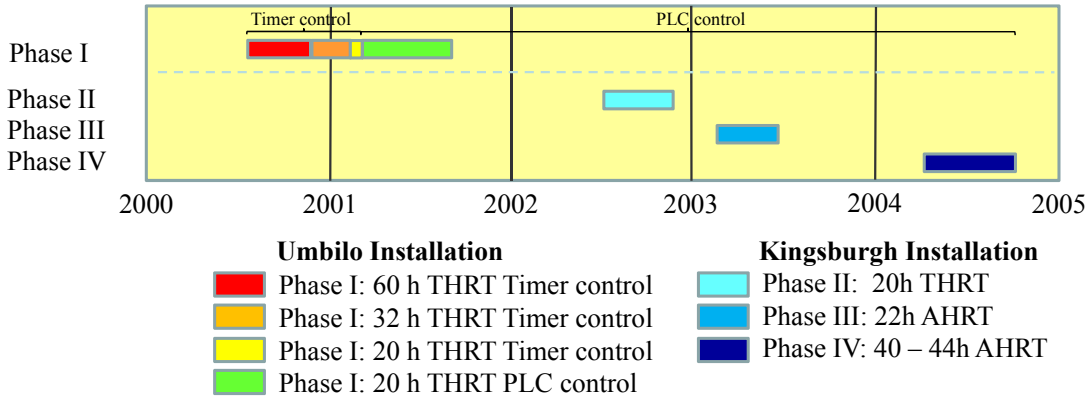

Figure 1.3: Pilot-scale ABR project time line showing different phases of operation included in this study at Umbilo and Kingsburgh WWTP. T-HRT: Target hydraulic retention time; A-HRT: Applied hydraulic retention time (average calculated value for period).

Four periods, together making up one essentially continuous phase of operation are shown during the time in which the pilot-scale ABR was installed at head of works at Umbilo WWTP from July 2000 to September 2001. Three distinct phases of operation were considered from the data obtained during operation at Kingsburgh WWTP in 2002, 2003 and 2004. Details of operation for each of the phases are described in detail in **Chapters 4** and **5**.

1.7 ORGANISATION OF THE THESIS

This study documents the performance of the pilot-scale ABR and considers its application in the treatment of domestic wastewater. **Figure 1.4** shows the logical flow of this thesis.

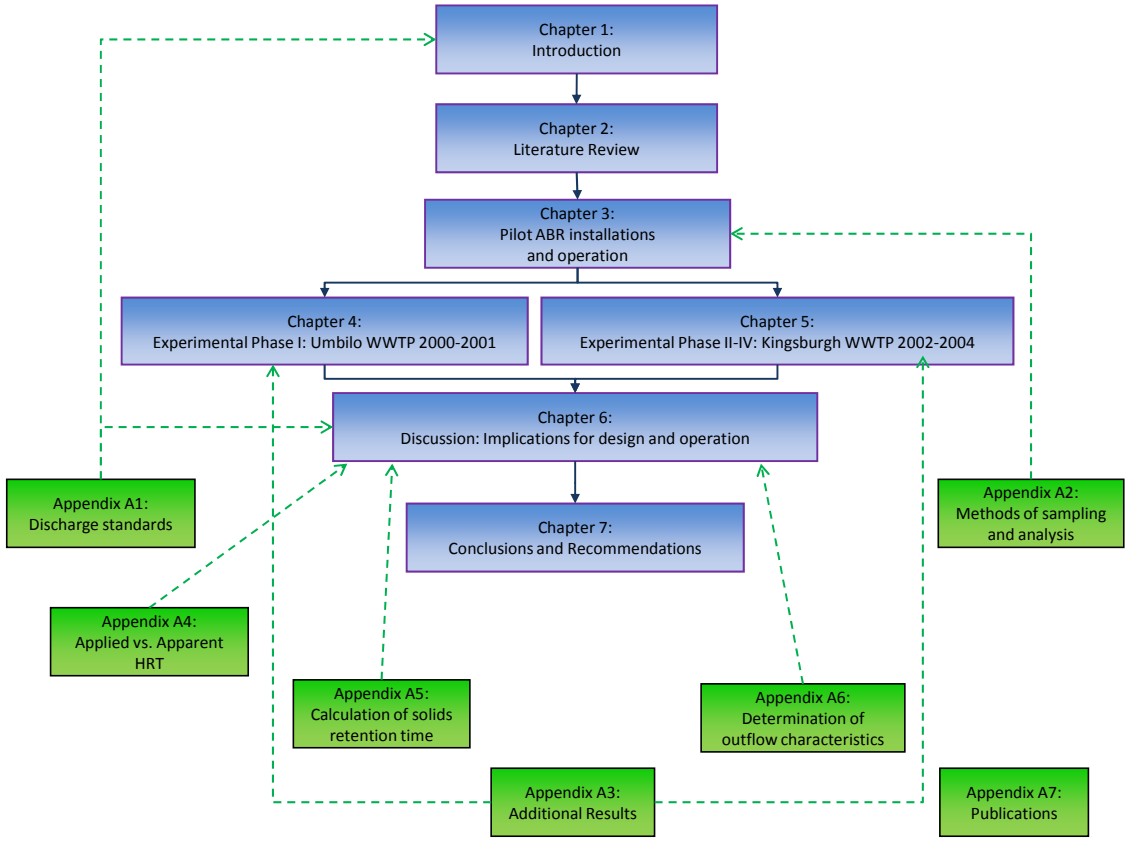

Figure 1.4: Organisation of the thesis

Chapter 1 introduces low cost sanitation in South Africa and the ABR project. Objectives of the study and the hypotheses of the research are presented.

Chapter 2 presents an overview of relevant literature including general theory of anaerobic digestion, anaerobic digester technology, anaerobic digestion of sewage and research into the anaerobic baffled reactor.

Chapter 3 is the materials and methods chapter of the thesis, and describes the construction of the pilot-scale ABR, installations at Umbilo and Kingsburgh WWTP, and sampling and analytical methods used in this study.

Chapter 4 and **Chapter 5** present and analyse the results of operation of the pilot-scale ABR at Umbilo and Kingsburgh WWTP.

Chapter 6 is the central chapter of the thesis in which the findings of the previous two chapters are considered in terms of what they reveal about design requirements of an ABR for domestic wastewater treatment. Gaps in the information are identified and mass balance and steady-state models are employed to supply the missing data, where possible.

Chapter 7 presents the conclusions and recommendations for future research.

Appendices A1 to **A6** provide additional data and analysis to support material presented in **Chapters 3** to **7**, as depicted in **Figure 1.4**. **Appendix 7** lists all publications that have arisen from this research.

2 LITERATURE REVIEW

A detailed review of on-site sanitation and decentralised wastewater treatment systems is presented in Foxon et al. (2006). For the purposes of this thesis, the literature review will concentrate on microbiology and biochemistry of anaerobic digestion systems, and previous research into ABR technology and other anaerobic decentralised waterborne sewerage treatment systems.

2.1 ANAEROBIC DIGESTION: MICROBIOLOGY AND BIOCHEMISTRY

Anaerobic digestion converts organic matter to inorganic end products along a series of interrelated biochemical pathways (Bailey and Ollis, 1986). Traditionally, anaerobic digestion has been used for passive treatment of domestic wastewater in septic tanks, but it is best understood as a process in the pre-treatment of high strength industrial effluents or in the disposal of waste activated sludge from aerobic wastewater treatment (Speece, 1996). Consequently anaerobic digestion of domestic wastewater is considered to be a *low-strength* application. This classification is the source of some confusion: in industrial anaerobic digestion applications, inlet chemical oxygen demand (COD) concentrations may exceed 5 000 mgCOD/ℓ (Speece, 1996). However, domestic wastewater with a COD value of 1 000 mgCOD/ ℓ is considered to be a high-strength domestic wastewater (Henze et al., 1997). Therefore even a concentrated domestic wastewater is considered a low-strength feed in an anaerobic process.

In aerobic respiration, molecular oxygen serves as an external electron acceptor, and there is a large flow of electrons and energy associated with these conversions (Bailey and Ollis, 1986). In the absence of an external oxygen supply, some carbon atoms associated with organic substrates are reduced (ultimately to CH_4) by accepting electrons from other compounds that are oxidised to carbon dioxide (CO_2). The conversions are therefore characterised by much smaller energy and electron fluxes resulting in a smaller driving force for the reactions (Speece, 1996 pp. 40-43).

Anaerobic conversion to CH_4 gas therefore provides relatively little energy to micro-organisms, resulting in a slow growth rate and only a small portion of the waste being converted to new biomass (i.e. low sludge yields). The conversion of organic material to CH_4 removes COD from the liquid phase.[1]

As much as 80 to 90 % of the degradable organic portion of a waste can be stabilised in anaerobic treatment, even in highly loaded systems (Speece, 1996 p. 11).

[1] Production of CO_2 gas does not indicate COD reduction; in anaerobic digestion, where there is no external oxygen supply; CO_2 production depends on internally available oxygen in the substrate (such as in the acid group of organic acids i.e. $CH_3(CH_2)_xCOOH$) and therefore does not contribute to the *oxygen demand* of the wastewater measured by the COD analysis. (Speece, 1996)

2.1.1 Sub-processes within anaerobic digestion

Figure 2.1 presents a conceptual flow of COD in catabolic anaerobic digestion (i.e. ignoring COD converted to biomass), from a hypothetical substrate containing 30% each of carbohydrate, protein and lipid and 10% inert material.

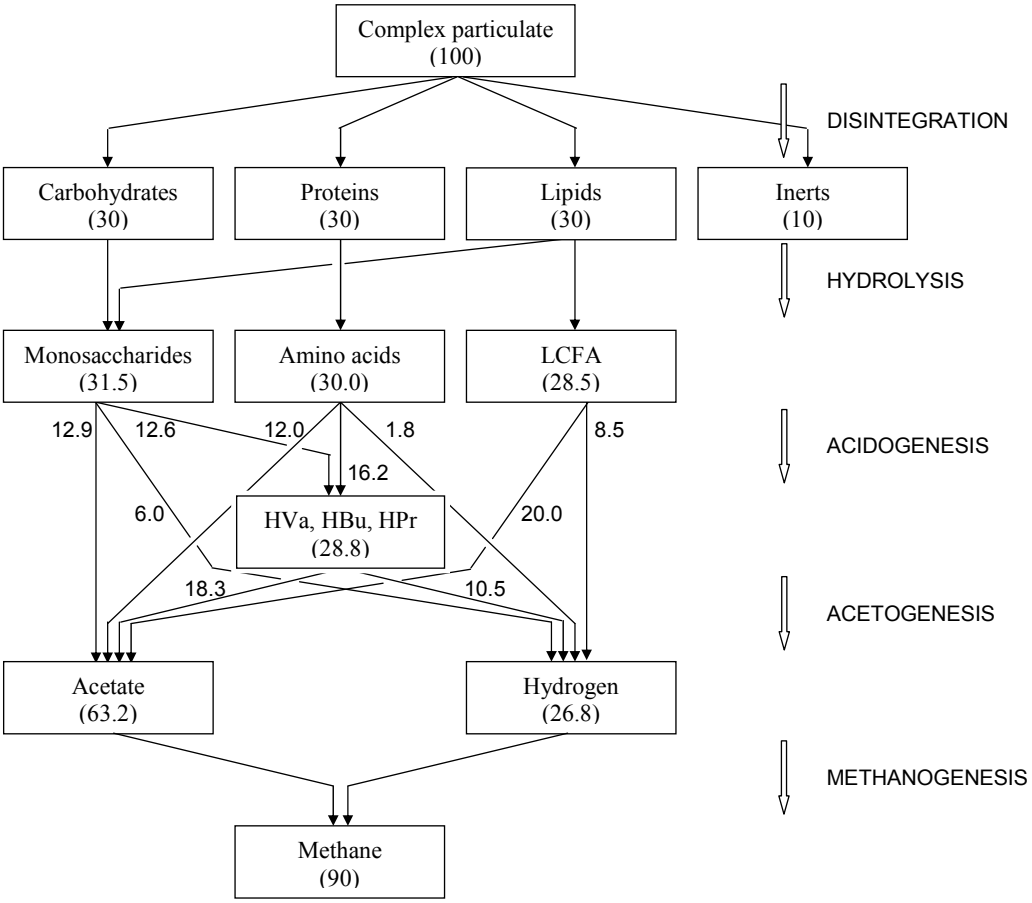

Figure 2.1: Flow-diagram for the anaerobic degradation of a composite particulate material, as implemented in ADM1 (adapted from: Batstone et al., 2002). Valerate (HVa), Butyrate (HBu) and Propionate (HPr) are grouped for simplicity. LCFA are long chain fatty acids. Figures in brackets indicate COD fractions

For complete digestion of the biodegradable COD (complete stabilisation) all COD is recovered as CH_4 gas.

There are 5 major sub-processes within the overall process of anaerobic conversion of complex organic substrates to CO_2, elemental hydrogen (H_2) and CH_4: disintegration, hydrolysis, acidogenesis, acetogenesis and methanogenesis. Each sub-process is mediated by one or more different microbial groups. **Sections** 2.1.1.1 to 2.1.1.7 summarise the main features of the sub-processes primarily responsible for conversion of COD to H_2 and CH_4 endproducts in anaerobic digestion, while **Sections** 2.1.1.8 and 2.1.1.9 describe additional processes that may occur in anaerobic digestion under appropriate conditions. This summary of anaerobic digestion processes is drawn from a review of the development of anaerobic digestion modelling by Remigi and Foxon (2004).

2.1.1.1 Disintegration and hydrolysis

The process of breaking down complex feed components into simple substances that can be assimilated into micro-organisms can be divided into two steps:

- Disintegration of composite particulate material into smaller carbohydrate, protein and lipid fractions. This is usually achieved by mechanical means including agitation or as a result of dissolution of binding agents in a complex particulate structure.

- Extracellular enzymatic hydrolysis of large molecular weight compounds to long chain fatty acids, simple sugars and amino acids. Hydrolysis is accompanied by some release of energy (Batstone et al., 2002).

2.1.1.2 Anaerobic oxidation

Anaerobic oxidation is the process by which long chain fatty acids are oxidised to simple organic acids (otherwise called volatile fatty acids or VFA). The long carbon chain is sequentially shortened by the removal of two carbon atoms (an acetate molecule) at each step. The final product of anaerobic oxidation of fatty acids with an even number of carbon atoms is acetate only; when the fatty acid has an odd number of carbon atoms, one mole of propionate is produced per mole of substrate.

$$CH_3(CH_2)_nCOOH + n \cdot H_2O \rightarrow \left(\frac{n+2}{2}\right)CH_3COOH + n \cdot H_2 \quad \text{(n is even)}$$

$$CH_3(CH_2)_nCOOH + (n-1) \cdot H_2O \rightarrow CH_3CH_2COOH + \left(\frac{n-1}{2}\right)CH_3COOH + (n-1) \cdot H_2 \quad \text{(n is odd)}$$

Eq. 2-1

Anaerobic oxidation requires the use of hydrogen ions (H^+) as electron carriers. These may be assumed to be derived from the dissociation of water as shown in Eq. 2-1. Consequently relatively large amounts of dissolved elemental hydrogen (H_2) are released in this process (Batstone et al., 2002).

2.1.1.3 Acidogenesis

Amino acids and simple sugars are fermented by acidogens or *acid formers* that produce VFA including acetic, propionic, butyric and lactic acid. Acidogenesis can occur without an additional electron donor or acceptor (Gujer and Zehnder, 1983). The VFA end-product of acidogenesis is determined by the environmental conditions (Mosey, 1983). Different amounts of H_2 are produced during acidogenesis depending on the acidogenesis end-product. Eq. 2-2 shows the production of acetic acid from glucose by acidogenesis with H^+ ions acting as electron acceptor.

$$C_6H_{12}O_6 + 2H_2O \rightarrow 2CH_3COOH + 2CO_2 + 4H_2 \qquad \textbf{Eq. 2-2}$$

This is the thermodynamically preferred reaction since it provides acid-forming bacteria with the biggest energy yield. Other VFA including propionic, butyric and lactic acids or a combination thereof may also be produced by acidogenesis, according to the type of micro-organisms present and solution thermodynamics, largely governed by dissolved H_2 concentration (Mosey, 1983). Production of butyric and propionic acid occurs in response to accumulation of dissolved H_2 following high organic loading incidents, since the H_2 yield of these reactions is less (or in the case of propionic acid, reversed) than the reaction that produces acetic acid; through this self-regulated microbial switching of biochemical pathways, micro-organisms are able to play an active role in controlling redox potential (Mosey, 1983).

Acidogenesis of amino acids usually occurs via Stickland oxidation where different amino acids are fermented pairwise, with one amino acid in a pair acting as electron donor, while the other acts as electron acceptor (Batstone et al., 2002). There is typically a 10% shortfall of electron acceptor and

therefore approximately 10% of amino acid acidogenesis occurs by uncoupled oxidation with H^+ ions or CO_2 acting as an electron acceptor (Nagase and Matsuo, 1982). Thus the products of amino acid digestion will be VFA, CO_2, NH_3, H_2 and reduced sulphur, depending on which of the 20 amino acids have been degraded (Batstone et al., 2002). Significant amounts of NH_3/NH_4^+ may be released depending on the nitrogen content of the digested organic material (Speece, 1996).

2.1.1.4 Acetogenesis

A further category of bacteria (acetogens) ferment propionic, butyric and lactic acids to acetic acid. In most cases, each group of acetogens can only ferment one type of VFA. This is considered a separate step to acidogenesis since no large pH effect is associated with the conversion of higher acids to acetic acid, and there is no internal electron acceptor (Batstone et al., 2002). Eq. 2-3 shows acetogenesis from butyric acid using H^+ ions as an electron acceptor to produce acetic acid and H_2.

$$CH_3 - CH_2 - CH_2 - COOH + 2H_2O \rightarrow CH_3COOH + 2H_2$$ **Eq. 2-3**

2.1.1.5 Acetoclastic methanogenesis

The final stage in anaerobic digestion in which reduced carbon is removed from the reaction liquors is the conversion of acetic acid to CH_4 and CO_2 by a group of Archaea known as acetoclastic methanogens (Eq. 2-4). The conversion to CH_4 is the only strictly anaerobic step that results in the removal of COD originating from reduced carbon to the gas phase.

$$CH_3COOH \rightarrow CH_4 + CO_2$$ **Eq. 2-4**

Acetoclastic methanogens all belong to the Kingdom Archaea and are found in only two genera, *Methanosarcina* and *Methanosaeta* (Madigan et al., 1996).

Methanogenic micro-organisms may compete with sulphate-reducing micro-organisms if sulphate is present at sufficiently high concentrations (Speece, 1996).

2.1.1.6 Hydrogenotrophic methanogenesis

Hydrogenotrophic methanogenesis is the production of CH_4 from dissolved H_2 and CO_2 by a select group of slow-growing methanogens (Eq. 2-5). This process can account for up to 30 % of the CH_4 produced by anaerobic digestion of an organic waste, and plays a key role in controlling dissolved H_2 concentration in the reaction liquor. Removal of COD associated with hydrogen is achieved via this route.

$$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O$$ **Eq. 2-5**

The Gibbs free energy of reaction ($\Delta G'$) is only favourable (i.e. < 0) for hydrogenotrophic methanogenesis in a narrow concentration range of HCO_3^- and H_2. Therefore changes in reaction conditions can easily disrupt this process (Madigan et al., 1996; Batstone et al., 2002).

2.1.1.7 Homoacetogenesis

Homoacetogenesis is the generation of acetic acid from dissolved H_2 and CO_2 (Eq. 2-6).

$$4H_2 + 2CO_2 \rightarrow CH_3COOH + 2H_2O$$ **Eq. 2-6**

Homoacetogens are one of the most versatile physiological groups among the anaerobic bacteria. They can use one-carbon compounds as substrate and can carry out partial oxidation of reduced fermentation products released by other fermenting bacteria. Homoacetogens can use various substrates sequentially or simultaneously and may constitute an energy link from hydrogen, via acetate to acetoclastic methanogens (Madigan et al., 1996).

2.1.1.8 Sulphate reduction

The presence of any sulphate in the feed to an anaerobic digestion process will result in sulphidogenesis, the generation of sulphide (S^{2-}) from sulphate (SO_4^{2-}). As with methanogenesis, this process can be either acetoclastic or hydrogenotrophic. Organisms reducing sulphur can obtain the electrons directly by oxidising VFA, or by oxidising the H_2 produced by acetogens. Additionally, VFA are used as a carbon source, and as a result, organisms reducing sulphur compete with the majority of groups in anaerobic digestion (Kalyuzhnyi and Fedorovich, 1998). Eq. 2-7 shows two routes for the production of H_2S from H_2SO_4 using either acetate or H_2 as electron donor. The effect of sulphate reduction on anaerobic systems is further complicated by the fact that the reduced product, sulphide, has an inhibitory effect on almost all microbial groups (Batstone et al., 2002).

$$CH_3COOH + H_2SO_4 \rightarrow H_2S + 2CO_2 + 2H_2O$$
$$4H_2 + H_2SO_4 \rightarrow H_2S + 4H_2O \qquad \textbf{Eq. 2-7}$$

2.1.1.9 Denitrification

Denitrification, or dissimilatory nitrate reduction, is the reduction of nitrate (NO_3^-) to nitrogen oxides (NO_2^-, NO, N_2O) and N_2 under anoxic conditions. Denitrification results in the removal of nitrogen from the liquid phase to N_2 in the headspace. Denitrifying micro-organisms have a higher cell growth yield per unit substrate consumed than methanogenic micro-organisms and compete for the same carbon source and electron source (e.g. acetate or H_2). Eq. 2-8 shows the overall reduction of nitrate by acetic acid to produce N_2.

$$5CH_3COOH + 8NO_3^- + 8H^+ \rightarrow 4N_2 + 10CO_2 + 19H_2O \qquad \textbf{Eq. 2-8}$$

In most cases, nitrate reduction takes place in a number of steps (i.e. $NO_3^- \rightarrow NO_2^- \rightarrow NO \rightarrow N_2O \rightarrow N_2$), and is usually mediated by at least two different types of bacteria (Madigan et al., 1996). In addition to competition for the same substrates as methanogenesis, intermediate species in denitrification have also been found to be inhibitory to methanogenic micro-organisms. These two effects result in a decrease in CH_4 production and an increase in alkalinity. Thus in anaerobic digestion, the presence of nitrate can have significant impact on carbon and electron flow and thereby CH_4 production, on microbial competition and inhibition, and on gas composition (Batstone et al., 2002).

2.1.2 Physico-chemical processes in anaerobic digestion

There are three broad types of non-biological chemical processes (Remigi and Foxon, 2004):

- *Acid-base equilibrium processes*: There are a number of important acid-base equilibrium systems, which have dissociation constants close to the operating pH of anaerobic systems.

- *Gas-liquid processes* (i.e., gas-liquid transfer of CO_2, CH_4, H_2, hydrogen sulphide (H_2S) and nitrogen gases).

- *Liquid-solid processes* (i.e., precipitation/solubilisation of ions).

Important physico-chemical subsystems include inorganic carbon, VFA dissociation, sulphate, sulphite, sulphide, free and saline ammonia (NH_3+NH_4^+), oxidised nitrogen, phosphate, and the gas-liquid interaction of H_2 and CH_4 gases. These processes impact on the availability of inorganic species in solution and may profoundly affect the pH and alkalinity in an anaerobic process. The role of alkalinity in particular is discussed further in **Section** 2.1.6.

2.1.3 Interaction of sub-processes in anaerobic digestion

Redox potential and acidity/alkalinity of the liquid phase are determined by intermediates and by-products of anaerobic digestion, H_2, CO_2 and VFA, and affect the available energy of many of the sub-processes (See e.g. Smith and Van Ness, 1987 for calculations of Gibbs free energy of reaction under different redox/pH conditions). Acetoclastic methanogenesis is particularly vulnerable to low pH conditions and quickly becomes inhibited if the pH drops below a value of 6.5. The overall anaerobic digestion process is therefore balanced between acid-producing acidogenesis and acid-consuming methanogenesis; any event that causes an increase in acid production rate (e.g. high organic load) or decrease in rate of acid removal (due to e.g. a decrease in buffering and therefore pH) can cause *souring* i.e. where low pH causes complete inhibition of methanogenesis and the ability of the system to remove acid fails (Speece, 1996 p. 184). Although it is possible to recover from souring, the overall rate of anaerobic digestion/stabilisation decreases considerably. Other anaerobic digestion products and intermediates (e.g. H_2, NH_3) can also cause inhibition of different sub-processes, with implications on micro-organism selection and overall rate of stabilisation.

From a thermodynamic perspective, the energy change (per unit mole) of substrate fermented to H_2 or CH_4 in anaerobic digestion is small; moreover, most organic compounds are fermented to these end products in a step-wise process by different groups of micro-organisms; therefore each of the different micro-organism groups involved is only able to harvest a fraction of the energy yield from the overall conversion. The extent to which bacterial growth occurs is a function of the energy released by the electron transfer and the efficiency of the energy utilisation by the micro-organism mediating the transfer (McCarty, 2006). The small amount of energy released in anaerobic transformations results in a low growth yield for all types of anaerobic micro-organisms, although the individual growth yield for each type of micro-organism depends on the energy release by the reaction that they mediate and the conditions under which they operate. The dominant micro-organisms in a consortium will be those that can harvest the available energy in any substrate most rapidly and efficiently in a given environment. These species will out-compete other micro-organisms that utilise the same substrate (Mosey, 1983; Bailey and Ollis, 1986; Batstone et al., 2002).

This has consequences for process stability since upsets in environmental conditions can reduce the Gibbs free energy of a certain sub-process to the point where it is no longer thermodynamically feasible, or only at significantly reduced rates. This in turn interrupts the supply of substrate and therefore energy to all subsequent processes in the metabolic chain. For example, the balance between propionate oxidation, acetate decarboxylation and H_2 oxidation is critical for a stable anaerobic digestion process. The optimal range for all three reactions (i.e., they must all be exergonic) is very narrow, and is mainly controlled by free propionate, H_2 and acetate concentrations (Gujer and Zehnder, 1983; Batstone et al., 2002).

Most of the control in anaerobic digestion is undertaken directly by the micro-organisms themselves: i.e., pH regulation by the formation and removal of short chain fatty acids and regulation of redox potential by formation and removal of trace concentrations of dissolved H_2 (Mosey, 1983). The specifics of pH and hydrogen regulation have been shown to be complicated, and are not fully understood: product and substrate inhibition of several of the anaerobic sub-processes have been observed, resulting in feedback loops which assist in maintaining pH and redox conditions in ranges which allow anaerobic digestion to proceed (Batstone et al., 2005; Rodríguez et al., 2005). It is possible to increase the range of operating conditions and stability to a certain extent by a physical separation of acidogenic and methanogenic bacteria in serial reactors since they are sufficiently different with respect to their physiology and nutritional requirements (Cohen et al., 1980). This spatial separation, often termed *phase separation* or *stage separation* is discussed in greater detail in **Section** 2.2.3.

2.1.4 Stoichiometry and rates of anaerobic digestion processes

An understanding of the stoichiometry of the individual processes is necessary to construct mass balance relationships between substrates and products, cell yields and to calculate pH changes. The rate of the process will control the amount of substrate consumed or endproduct (including biomass) produced within a specified time

2.1.4.1 Stoichiometry of dissimilatory (energy generating) processes

The biochemical pathways of anaerobic digestion are complex and the most energetically favourable route to convert a substrate to an endproduct may change as a result of changing operating conditions, available substrate/intermediate concentrations and micro-organisms. Standard stoichiometry to describe the sub-processes in anaerobic digestion, expressed in Petersen matrix form are defined in the Anaerobic Digestion Model No. 1 (Batstone et al., 2002). These are necessarily a simplification of the complexity observed in actual systems, but are regarded as offering sufficient complexity to adequately simulate most real systems given the limited nature of most experimental data available. (Remigi and Foxon, 2004; Batstone et al., 2005).

2.1.4.2 Stoichiometry of assimilatory (growth) processes

The amount of growth that is associated with the consumption of a unit of substrate is described by the yield coefficient Y, which may be defined as the amount of new active cell mass (usually in gVSS or gCOD) that is generated per mass of substrate consumed (measured in g or mole of substrate, or gCOD) (Henze et al., 1997). (**Table 2.1**)

Each micro-organism that catalyses a process in anaerobic digestion has an independent yield coefficient that depends to a certain extent on the free energy of reaction associated with the conversions it mediates. However, different sub-processes with similar functions tend to have similar yields. Generally, the yield of CH_4-generating micro-organisms is lower than that of acid producing organisms (Pavlostathis and Giraldo-Gomez, 1991).

Substrate limitations can result in a considerably lower overall yields with values between 0.05 and 0.1 mgCOD/mgCOD (Henze et al., 1997).

Table 2.1: Typical values of cell growth yield for acidogenic and methanogenic processes in anaerobic digestion (from Pavlostathis and Giraldo-Gomez, 1991; Henze et al., 1997; and Batstone et al., 2002)

	Y (mgVSS/mgCOD) (Pavlostathis and Giraldo-Gomez, 1991)	Y (mgVSS/mgCOD) (Henze et al., 1997)	Y (mgCOD/mgCOD) (Henze et al., 1997)	Y (mgCOD/mgCOD) (Batstone et al., 2002)
Acidogenesis	0.15	0.15 – 0.20	0.2 – 0.3	0.04 - 0.10
Methanogenesis	0.03	0.03 - 0.04	0.04 – 0.05	0.05 – 0.06

2.1.4.3 Reaction rates in anaerobic digestion

Each of the five sub-processes described in **Figure 2.1** proceed at different rates, depending on operating conditions and substrate concentrations. The overall rate of stabilisation therefore will be limited by the slowest or rate-limiting step. The rate-limiting step will be different in different systems, and may even change from one sub-process to another with time within a system (McCarty and Mosey, 1991). Extracellular process kinetics tend to be slow, and are generally poorly characterised (Vavilin et al., 2001). For this reason, disintegration and hydrolysis are often lumped in a single process with first order or surface saturation-type kinetics (Eastman and Ferguson, 1981; Vavilin et al., 2001; Batstone et al., 2002). Acid and CH_4 producing steps usually exhibit a Monod-type relationship between reaction rate and substrate concentration. As a general rule, when the primary substrate is soluble or labile, the rate-limiting step will be methanogenesis, while extracellular processes will dominate the overall kinetics of digestion of particulate or refractory substrates (Pavlostathis and Giraldo-Gomez, 1991; Lyberatos and Skiadas, 1999).

2.1.5 Factors affecting the rate and extent of anaerobic digestion

In all sub-processes of anaerobic digestion, adverse conditions (e.g. low pH for methanogenesis or high dissolved H_2 concentration for acid production) will slow or halt the reaction in question. Overall environmental conditions such as pH value, temperature, essential trace nutrients and toxicants can play a major role in modifying the reaction rates of individual sub-processes in anaerobic digestion.

2.1.5.1 Types of inhibition

The IWA Task Group for Mathematical Modelling of Anaerobic Digestion Processes (Batstone et al., 2002) defined two general categories of inhibition for micro-organisms. The first, *biocidal inhibition* describes reactive toxicity experienced by the micro-organism due to a chemical or condition that is normally irreversible (i.e. the micro-organisms will not completely recover subsequent to the removal of the adverse condition / chemical)l. The second category is *biostatic inhibition*. In this case, growth related functions cease during exposure to inhibitory conditions, but may resume when growth-conducive conditions are re-established.

Anaerobic processes show similar patterns of inhibition to aerobic processes, and therefore cannot be regarded as being inherently sensitive to inhibition. As with all inhibition effects, the slowest processes will succumb first, and in the case of anaerobic digestion, this is usually methanogenesis. Sufficient inhibition of methanogenesis results in acid accumulation and failure of digestion, and

therefore the sensitivity of the overall stability of anaerobic processes to toxicants is greater than in aerobic processes (Henze et al., 1997).

2.1.5.2 pH inhibition

A pH range of 6.5 to 8.2 is generally considered acceptable for anaerobic digestion (Speece, 1996 p. 29), although the effect of pH is different for each of the subprocesses. Methanogenesis is particularly sensitive to pH values, exhibiting a rapid decrease in maximum reaction rate when the pH drops below a value of 6.5, or exceeds 8.5. pH inhibition occurs as a result of disruption of homeostasis, and increased levels of non-dissociated VFA (Batstone et al., 2002).

Table 2.2: Agents that cause biocidal inhibition and biostatic inhibition in anaerobic micro-organisms (from Batstone et al., 2002)

Biocidal inhibitors	Biostatic inhibitors
Long chain fatty acids (LCFA)	Product inhibition
detergents	weak acid/base inhibition (VFA, NH_3, H_2S)
aldehydes	pH inhibition
nitro-compounds	cation inhibition
cyanide	anything else that disrupts homeostasis[1]
azides	
antibiotics	
electrophiles	

Reactor failure or *souring* occurs through a vicious cycle of low pH values and methanogenesis inhibition: Low pH values inhibit removal of acid by methanogenesis, resulting in accumulation of volatile acids, which in turn lower the pH value. This problem is exacerbated by the fact that bicarbonate alkalinity does not buffer well at pH values below 6 and since decreasing pH is accompanied by increasing concentrations of un-ionised VFA that also have an inhibiting effect on methanogenesis (**Section** 2.1.5.3). This increases the extent of inhibition of methanogenesis until a point is reached where methanogenesis is completely inhibited. A common method of calculating pH inhibition for pH values below the optimum pH range is given in Eq. 2-9 (Batstone et al., 2002)

$$I = \exp\left[-3\cdot\left(\frac{pH - pH_{uL}}{pH_{UL} - pH_{LL}}\right)^2\right]_{pH<pH_{uL}} = 1\Big|_{pH>pH_{UL}} \qquad \textbf{Eq. 2-9}$$

where pH_{UL} is the lowest pH value where there are no pH inhibition effects (upper limit) and pH_{LL} is the highest pH value at which complete inhibition is observed (lower limit).

[1] Homeostasis is the sum of all biochemical processes within a cell that maintain the correct cytoplasm moisture, salt and other chemicals composition (Madigan et al., 1996).

The values of pH_{UL} and pH_{LL} differ for different trophic groups. This equation does not predict the effect of high pH values on reaction rates. **Figure 2.2** shows the shape of the pH inhibition function (Eq. 2-9) for pH_{UL} = 7 and pH_{LL} = 6. These are the recommended values for pH inhibition of acetoclastic methanogenesis (Batstone et al., 2002).

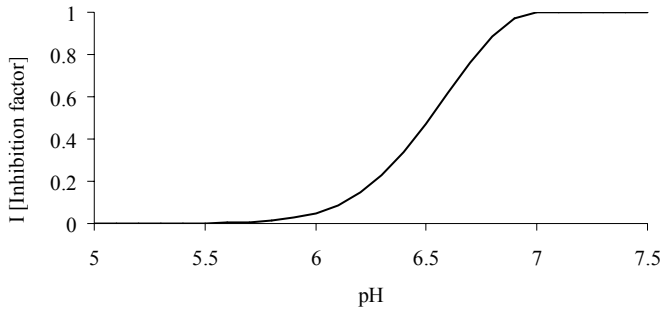

Figure 2.2: Shape of pH inhibition function calculated from Eq. 2-9 for acetoclastic methanogenesis with pH_{UL} = 7 and pH_{LL} = 6

Inhibition of biological functions at pH values above 8 may be related to the association of NH_4^+ to form NH_3, which is toxic at high concentrations to most anaerobic micro-organisms (Speece, 1996).

2.1.5.3 Weak acid/base inhibition

Undissociated acids and bases disrupt cell homeostasis by passively diffusing through the cell membrane and dissociating inside the cell cytoplasm. Since the amount of undissociated weak acid or base present is a function of pH, inhibition by undissociated acid or base is also pH dependent. It is difficult to observe these two inhibition effects independently; hence the empirical pH inhibition term often incorporates inhibition by weak acids or bases (Batstone et al., 2002). However Mösche and Jördening (1999) showed that both factors (low pH and undissociated acid inhibition) affected reaction rates in anaerobic digestion. Undissociated VFA that are understood to cause inhibition include acetic, propionic, butyric and valeric acids, which all have pKa values in the range 4.7 to 4.9 (at 25 ºC). NH_3 is the principle weak base that causes inhibition at higher pH values. The pKa value for NH_3 is 9.3 at 25 ºC (Mösche and Jördening, 1999; Batstone et al., 2002).

2.1.5.4 Dissolved H_2 inhibition

The ability of micro-organisms to complete a biological conversion is dependent on the free energy change of the conversion under reaction conditions (**Section** 2.1.3). The free energy change of conversion is strongly dependent on the concentration of dissolved hydrogen in solution since at higher concentrations of H_2, the pε, which measures the relative tendency of a solution to accept or donate electrons is low, resulting in a low tendency for oxidation and conversely a higher tendency for reduction (Stumm and Morgan, 1996 pp. 429, 473-476). Thus biological transformations, particularly those that produce H_2 from reduced organic compounds become thermodynamically impossible.

From **Figure 2.3**, conversion of propionate to acetate is only thermodynamically possible at H_2 concentrations less than 10^{-4} M, although conversion of ethanol to acetate continues at far higher concentrations. Thus, increasing concentrations of dissolved H_2 result in inhibition of processes responsible for the production of acetate, thereby disrupting metabolic processes that ultimately produce methane (CH_4). Dissolved H_2 concentration is controlled by H_2-utilising micro-organisms that scavenge H_2 produced by fermentation processes, thereby allowing fermentation processes dependent on dissolved H_2 concentration to proceed.

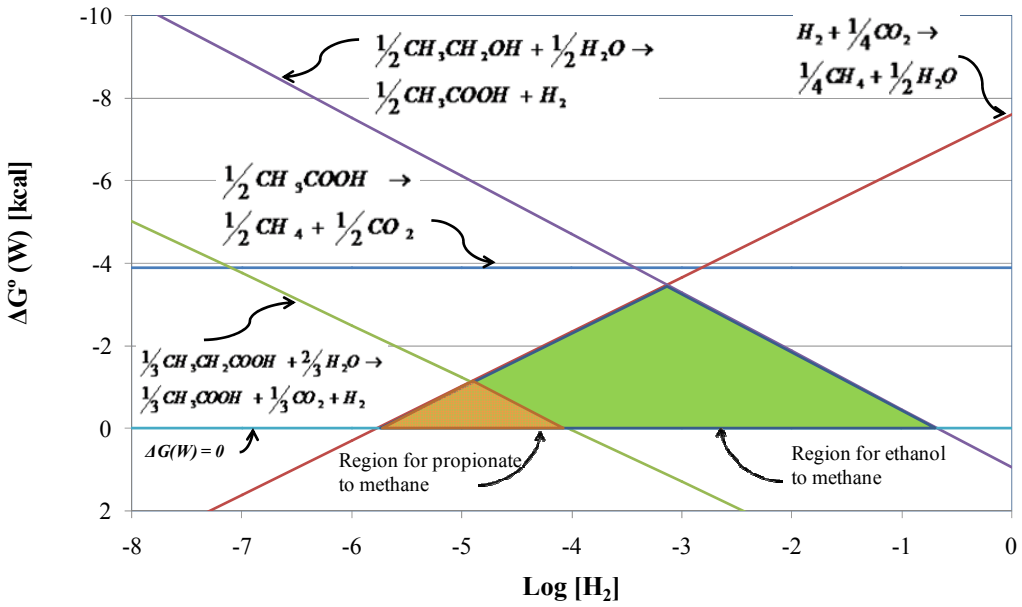

Figure 2.3: Change in free energy available for reaction vs. dissolved H_2 concentration showing H_2 concentrations at which conversion of propionate or ethanol to methane is possible (Reproduced from Speece, 1996 p. 43)

2.1.5.5 Temperature

For cryophilic (0 to 25ºC) and mesophilic (20 to 40ºC) temperature ranges, the change in reaction rates of anaerobic processes with temperature are commonly described by an equation of the following form (Eq. 2-10):

$$\mu_{max}(T) = \mu_{max}(20°)e^{\kappa(T-20)} \qquad \textbf{Eq. 2-10}$$

Each subprocess will have different temperature coefficients (κ)

Anaerobic digestion between 40ºC and 50ºC is unstable and prone to failure. At temperatures above 50ºC, thermophilic micro-organisms operate at higher rates than their mesophilic counterparts, but little or no activity occurs above 70ºC (Henze et al., 1997).

Traditionally, anaerobic digesters are operated at temperatures in the mesophilic range since good stability and high reaction rates may be obtained. However, many applications have been successfully tested at ambient temperatures. Some pertinent studies are summarised in **Table 2.5** (**Section** 2.4.3).

2.1.5.6 Nutrients

The nutrient requirements of anaerobic digestion are relatively small since nutrient requirements are essentially linked to growth and anaerobic processes are characterised by low growth yields (Speece, 1996 p. 59). As with all biological processes, nitrogen, phosphorus, sulphur and iron are required for growth, as well as a host of other micro-nutrients that are required in trace amounts.

2.1.6 Alkalinity

Bicarbonate alkalinity is a critical factor in controlling the condition of anaerobic digestion. It is a key factor in maintaining digester pH during the production of weak acids by the digestion processes. This is particularly important in anaerobic systems as compared to aerobic systems since one of the weak

acids produced by digestion, CO_2, is not stripped off by aeration as in aerobic systems, while the other class of weak acids, VFA, can be utilised and thus removed by relatively few classes of anaerobic micro-organisms.

2.1.6.1 Definition of alkalinity

Alkalinity measures the acid neutralising capacity of an aqueous system and is primarily determined by the concentration of the salts of weak acids present (Speece, 1996 p. 184; Stumm and Morgan, 1996 pp. 163-164). Alkalinity is measured by a titration of the solution in question with a strong acid to a pH endpoint determined by the equivalence point of carbonic acid, H_2CO_3. The amount of acid titrated describes the alkalinity and is reported in terms of $CaCO_3$ equivalence; i.e. number of equivalents of acid titrated divided by equivalent weight of $CaCO_3$ (50g/eq.) (Speece, 1996 p. 184).

2.1.6.2 Role of bicarbonate alkalinity in anaerobic digestion

The presence of acid-neutralising salts helps to buffer pH values in the digester, by reacting with CO_2 and VFA produced during digestion. If production of weak acids exceeds the available buffering capacity, the pH of the digester decreases, resulting in inhibition of microbial activity (**Section** 2.1.5.2 and 2.1.3, Speece, 1996 p.184). In most anaerobic systems, the predominant weak acid is CO_2. Thus the main purpose of maintaining a sufficiently high concentration of alkalinity in an anaerobic digester is to buffer carbon dioxide acidity. The most important contributors to the alkalinity measurement are the salts of inorganic carbon species (carbonate and bicarbonate). Since anaerobic applications usually operate at near neutral pH conditions, bicarbonate species dominate, and thus bicarbonate alkalinity is of principle interest. Salts of volatile acids, such as CH_3COONa (sodium acetate) also contribute to the measurement of alkalinity at neutral pH since they react with strong acid during the alkalinity titration, however they are not available for neutralising additional volatile acids produced during anaerobic digestion. Therefore a distinction is made between *total alkalinity* and *bicarbonate alkalinity*. Bicarbonate alkalinity refers to the total alkalinity measured by titration to a pH endpoint of around 4.5, *less* the volatile acid equivalent alkalinity, which is due to neutralisation of VFA (Speece, 1996 pp. 184, 187, 191).

2.1.6.3 Sources of bicarbonate alkalinity

There are two main sources of bicarbonate alkalinity in an anaerobic digester. The first is bicarbonate species present in the digester feed; the second is due to metabolism of organic compounds accompanied by the release of a cation. The latter source is termed *metabolism-generated alkalinity* (Speece, 1996 p. 190). Alkalinity is chiefly generated by degradation of proteinaceous organics through the production of NH_3, salts of organic acids or soaps through the release of the salt's cation, and sulphate or sulphite reduction. Conversely, biodegradation of organic compounds that do not result in the release of a cation; i.e. degradation of carbohydrates, organic acids, aldehydes, ketones and esters does not generate any alkalinity. The amount of metabolism-generated alkalinity in a system is proportional to the amount of cation released by the degradation of organics within the system as shown for the degradation of an amino acid in Eq. 2-11 (Speece, 1996 p. 190):

$$R-CH-NH_2-COOH+2H_2O \rightarrow RCOOH+NH_3+CO_2+2H_2$$
$$NH_3+H_2O+CO_2 \rightarrow NH_4^+ + HCO_3^- \qquad \textbf{Eq. 2-11}$$

For systems with low potential for generating alkalinity through metabolism (in particular, wastewaters with a low protein content) it may be necessary to add additional alkalinity in the form of lime (CaO), carbonate, hydroxide or bicarbonate for buffering digestion (Speece, 1996 p. 200).

2.1.7 Micro-organisms, biofilms and granules

This section provides a brief discussion of aspects relating to the microbiology of anaerobic systems including the types of micro-organisms, interaction between different classes of anaerobic micro-organism and their environment.

2.1.7.1 Micro-organisms

A range of different micro-organisms are involved in the anaerobic digestion of complex particulate material. In some cases, specific species may only mediate one biochemical conversion from a particular substrate to a particular product, and therefore are dependent on other micro-organisms to supply the substrate and consume the product. As a result, anaerobic micro-organisms work best in synergistic microbial communities to optimise substrate concentrations and minimise distances that substrates must diffuse.

All micro-organisms involved in anaerobic conversions in a conventional anaerobic digester are prokaryotic, i.e. they are simple single-celled organisms that do not possess a membrane-enclosed nucleus (Bailey and Ollis, 1986). Within the broader category of prokaryotic cells are two groups with distinct genetic, metabolic and morphological differences which are classified into different domains, true bacteria, or *Eubacteria,* and *Archaea* (Raskin et al., 1994).

Archaeal methanogens are responsible for conversion of acetic acid/acetate and dissolved H_2 to CH_4 and therefore play a central role in the removal of COD from wastewater. Hydrolytic and acidogenic conversions are undertaken by eubacterial species.

2.1.7.2 Anaerobic biomass communal synergism

Anaerobic sludge may be in the form of a dispersed sludge blanket consisting of micro-agglomerates (<50 μm diameter) and individual micro-organisms suspended in the reaction liquors, or a granular sludge consisting of distinct, almost spherical granules (Batstone et al., 2006). While it is possible for individual micro-organisms to live individually suspended in the reaction liquors, it has been found that aggregation of micro-organisms into a structured community such as in a biofilm or granule enables a close co-operation between micro-organisms successively involved in degradation of a substrate by reducing the distance over which intermediates produced by one organism must travel before being consumed by the next (Speece, 1996 p. 136). For example, hydrogenotrophic methanogens are often found near propionate-utilising acetogens, which produce significant amounts of H_2 as a by-product of propionate conversion (MacLeod et al., 1990).

Different classes of micro-organisms have been found to predominate in specific predictable locations of a biofilm or granule according to concentration gradients of different structures; thus micro-organisms that would be completely inhibited by conditions in the bulk liquor may thrive in the interior of a biofilm or granule; similarly, micro-organisms can metabolise at far higher rates than could be predicted from bulk phase substrate concentrations as a result of local increases in concentration of the substrate of interest within the structured anaerobic aggregate. McCarty and Smith (1986) present a model describing substrate concentration profile for ethanol, propionate, acetate and H_2 in the bulk liquid, mass transfer zone and within layers of an anaerobic biofilm that

considers the thermodynamic favourability of utilisation reactions at different locations. This model predicts where growth of particular micro-organisms will occur depending on whether the reaction they facilitate is thermodynamically favourable or not. Similar studies by other researchers have shown that the relative position of different micro-organisms within a biofilm or granule depends on thermodynamic and kinetic considerations and that the spatial location of micro-organisms plays an important role in the energetic and kinetics of the overall conversion of a substrate (Speece, 1996 p. 137; Batstone et al., 2006).

2.1.7.3 Granulation processes in anaerobic digestion

The formation of granular sludge has been observed in many anaerobic digestion applications. Granules are complex structures consisting of layers of micro-organisms associated with extracellular polymeric substances (EPS) and in some cases inorganic material that has become embedded in the granule, or that assisted in nucleus formation (Liu et al., 2003; Hulshoff Pol et al., 2004; Maximova and Dahl, 2006; Tiwari et al., 2006; Zhou et al., 2006). The presence of granular sludge in upflow anaerobic systems has ensured the success of these systems in high rate removal of organic material as a result of their superior settling characteristics and therefore high sludge retention rates (Hulshoff Pol et al., 2004; Zhou et al., 2006). A well-developed and stable granular sludge allows the establishment of high sludge loads in the reactor and maintains micro-organism diversity, even under conditions of high selection pressure[1] (Hulshoff Pol et al., 1983).

Granulation usually occurs spontaneously, although it may take several months from start-up of an anaerobic system before a stable granular sludge develops. There are a number of different theories that describe the mechanism of granulation. Most authors identify the filamentous acetoclastic methanogenic species *Methanosaeta concilii* (previously known as *Methanothrix soehngenii*) as playing a central role in the development of the granule through a number of possible mechanisms (MacLeod et al., 1990; Liu et al., 2003; Hulshoff Pol et al., 2004; Maximova and Dahl, 2006). Another critical component is the production of extracellular polymeric substances which are understood to affect the surface properties of bacterial flocs by supplying complexation sites on the surface of the micro-organisms and providing a structural support in which the colonies and individual micro-organisms become embedded (Zhou et al., 2007).

There is currently no accepted unifying theory that is able to describe all reported observations of granulation. However it is clear that substrate type and load and operating conditions all play a role in the structure of granules that form, the rate at which they form, and their behaviour under changing conditions (Tiwari et al., 2006).

2.2 ANAEROBIC DIGESTER TECHNOLOGY

Anaerobic digestion is carried out by a range of micro-organisms that may be largely or entirely self-regulated. However, the success of any system in treating a particular effluent will depend also on the

[1] Selection pressure is the continuous selection of sludge particles on the basis of size, density and structural integrity by the continuous application of hydraulic shear forces from upflow through the granular sludge bed. Upflow results in the washout of light and small granules or granule debris. (Hulshoff Pol et al., 1983).

design of the digester in which the micro-organisms are to perform the treatment. It is necessary to engineer the digester design to favour specific microbiological and biochemical conditions. This section presents a brief review of the main types of high rate anaerobic digestion reactor technologies.

2.2.1 Upflow Anaerobic Sludge Blanket (UASB) reactor

Probably the most widely used anaerobic digester technology is the upflow anaerobic sludge blanket (UASB) reactor. The concept of an upflow sludge blanket reactor was first proposed by Lettinga et al. (1980). The UASB reactor may be slowly stirred or not agitated and ensures contact between wastewater and anaerobic sludge by the introduction of the wastewater at the bottom of the reactor and retention of sludge by settling.

Wastewater flow is introduced at the bottom of the UASB reactor with treated effluent withdrawal at the top, resulting in upward vertical flow of liquid components and counter-current settling of settleable particulate components. Upflow velocities are typically around 1 m/h. The construction of the UASB reactor results in the development of four zones (**Figure 2.4**); (i) the sludge bed is a layer of settled anaerobic biomass that remains in the bottom of the reactor; (ii) the sludge blanket is a region of the reactor with a dense blanket of suspended or fluidised anaerobic sludge that remains suspended by the combined forces of the upflow of the liquid, settling of the solids and upward movement of biogas produced by anaerobic digestion; (iii) a gas-solid separator reduces the formation of a scum caused by attachment of solids to gas bubbles; and (iv) a settlement compartment creates a quiescent zone where solids are able to settle and return to the sludge blanket (Lin and Yang, 1991).

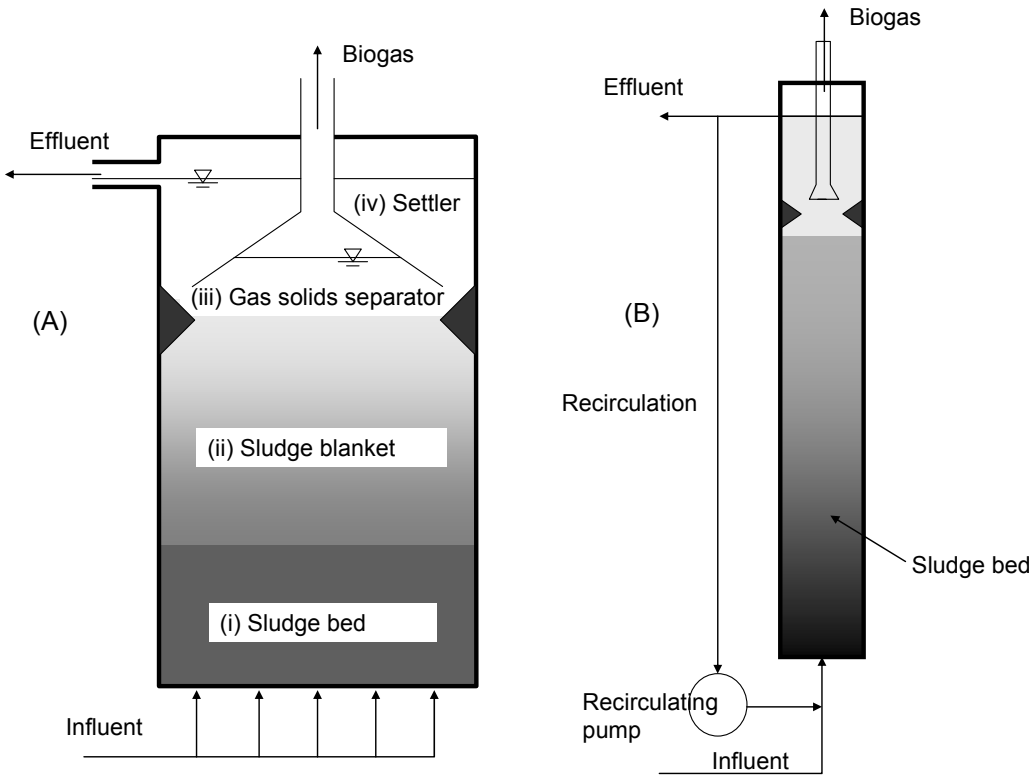

Figure 2.4: Basic design of (A) an upflow anaerobic sludge blanket (UASB) reactor (adapted from Lin and Yang, 1991) and (B) an expanded granular sludge bed (EGSB) reactor (adapted from Seghezzo et al., 1998).

The UASB has a number of significant advantages over earlier single stage anaerobic digesters (Seghezzo et al., 1998). Particulate components including active biomass granules and undegraded particulate material are able to settle and be retained in the reactor due to the relatively slow upward flow of liquid. Thus, good solids retention and high solids loading may be maintained. High rates of anaerobic digestion can be achieved in the dense sludge bed that collects at the bottom of the reactor. Particulate material with poor settling characteristics are entrained in the liquid flow and are not retained in the sludge bed or sludge blanket. Consequently, the system selects for well settling anaerobic biomass that is not susceptible to wash-out from the system (Hulshoff Pol et al., 1983). The high sludge load that can be achieved results in successful treatment at high organic loading rate (OLR). It has been found that agitation of the sludge bed results in attrition of sludge granules, resulting in poorer sludge settling characteristics (Lettinga et al., 1980). Thus natural agitation as a result of biogas generation in the sludge bed and the hydrodynamic forces of the incoming wastewater are often sufficient to ensure adequate contact between sludge and wastewater. Hence the operating costs of the technology are fairly low. Similarly, the simple design and high treatment rate per unit volume result in low capital cost for the construction of this type of system (Seghezzo et al., 1998).

However, it has been shown that internal mixing in UASB reactors is not ideal; significant zones of *dead space*[1] may be observed in the reactor (Wu and Hickey, 1997). Furthermore, use of a single-stage UASB does not permit separation of acidogenic and methanogenic processes (**Section** 2.2.3), an effect which has been shown to be beneficial in the digestion of certain wastewaters.

UASB technology has been successfully used to treat a wide variety of wastewaters, both domestic and industrial (Lin and Yang, 1991). Lin and Wang (1991) review many published applications of UASB technology and reported that UASB technology could successfully treat soluble wastewaters at OLR values of up to 30 kg COD/m^3/d, (although OLR values of between 5 and 20 kg COD/m^3.d were more common), while lower loading rates were required for partially soluble wastewaters (0.5 to 5 kg/m^3.d).

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[1] Danckwerts (1953) describes *dead water* as fluid which is trapped in eddies and therefore spends longer than the average hydraulic retention time inside the reactor. The remainder of the flow passes through more rapidly than the average hydraulic retention time as a result of the reduced passage for flow (reactor volume less eddy volume). *Dead space* is a region of the reactor volume that is not available for liquid flow due, for example, to the presence of grit, or internal reactor features that have non-negligible volume.

Dead space is often understood to be a stagnant area of fluid that reduces the volume of a reactor for fluid flow and thus resulting in short circuiting of fluid around the stagnant area. This is used to explain measurements of mean residence time that are shorter than would be predicted from the empty reactor volume. However, any volume filled with liquid cannot be completely *dead* since liquid and soluble components can diffuse into and out of the volume, resulting in a long tail in the exit concentration curve of a tracer test, the end of which may be below the detection limits of the tracer, and thus left out of the residence time distribution analysis.

Levenspiel (1999) simplifies the analysis by using *dead water* to describe a hypothetical inert region of liquid into and out of which no diffusion occurs. This is as an approximation for the early part of a tracer curve when there is a significant volume of stagnant water (or eddy volume). However, this approach does not completely describe the tail of the tracer curve, and may result in an under-recovery of tracer.

2.2.2 Expanded granular sludge bed (EGSB) reactor

The expanded granular sludge bed (EGSB, **Figure 2.4**) reactor is also an upflow reactor, but is operated at much higher superficial velocity then a UASB system (Van der Last and Lettinga, 1992). The EGSB is a tall reactor with large height : diameter ratio. Effluent is recirculated to the reactor feed resulting in upflow liquid velocities that may exceed 4 m/h. The high upflow velocity causes the sludge bed to expand, eliminating dead zones and resulting in improved contact between sludge and wastewater. Small sludge granules and dispersed biomass are washed out at these velocities, causing the system to select for well-structured, large biomass granules that are associated with high rate anaerobic treatment and stable digestion, since these also tend to have the best settling characteristics. Effluent recirculation results in dilution of the wastewater, and causes the sludge bed to behave as a completely mixed tank (Seghezzo et al., 1998). Dilution of the influent with effluent results in low in-reactor concentrations of influent components, resulting in enhanced ability to handle toxicants. (Seghezzo et al., 1998). Kato et al. (1997) showed that low strength wastewaters (COD concentration < 200 mg/ℓ) could be satisfactorily treated with EGSB reactors. Higher OLRs (up to 40 kg COD/m^3/d) may be achieved in the EGSB reactor relative to the UASB. Poor removal of suspended solids and colloidal components was observed in EGSB reactors (Van der Last and Lettinga, 1992).

2.2.3 Stage separation in anaerobic digestion

It has been shown that stage separation can result in improved efficiency in terms of suspended solids and COD removal in anaerobic digestion. Stage separation refers to the creation of an initial acid step in which acidogenic processes result in low pH values (pH 4 to 6) and a second methanogenic stage which is operated at pH values appropriate for methanogenesis (pH 6.5 to 8.2). Speece (1996 p. 22) states that the benefits of stage separation are most apparent in the treatment of pollutants that yield propionic acid and H_2 intermediates. The concept was first proposed by Pohland and Ghosh (1971). Since then, several authors have demonstrated the advantages of stage-separated anaerobic digestion over single stage digestion, specifically the improved suspended solids and COD volumetric removal efficiencies that result (Anderson et al., 1994; Siegrist et al., 2002).

However, the advantages of two-phase digestion are less evident when the substrate is a complex organic or particulate material (Hanaki et al., 1987) since complete acidification of the substrate cannot be achieved in the first stage unless effective separation of particulate and soluble components can be achieved. Leitão et al. (2006) conclude that in two-stage systems treating domestic wastewater in tropical climates, methanogenesis will always occur in the first step, since acidification rates are not sufficiently high that accumulation of VFA will result in complete inhibition of methanogenesis.

Zeeman and Lettinga (1999) indicated that the choice between a one or two phase UASB system should be determined by the required sludge retention time (SRT), i.e. if the required SRT cannot be achieved in a single-phase system, a two phase system should be considered.

2.3 HYDRODYNAMICS IN ANAEROBIC DIGESTERS

The performance of an anaerobic digester in the treatment of a particular wastewater strongly depends on the hydrodynamics of the system since this determines how long both solids and liquid or dissolved components spend inside the system, and thereby the extent of degradation of anaerobic substrates and the condition and composition of the microbial populations that digest them. This section presents an

introduction to characterisation of reactor hydrodynamics and how reactor hydrodynamics impact on anaerobic digestion processes.

2.3.1 Theory of flow modelling

A flow model of a reactor requires an appreciation of the flow patterns within the reactor; it is relatively easy to develop a flow model for a reactor design that may be approximated by mixed flow or plug flow when there is a single fluid phase present. Analytical solutions for a number of cases are readily available in the literature (Levenspiel, 1999 pp. 90-119). However in many cases, certainly in the case of multi-stage digesters, flow patterns can deviate widely from these ideal flow patterns due to channelling of fluid, recycling of fluid, existence of stagnant regions in the reactor, or the presence of a different phase (Levenspiel, 1999 pp. 257-277).

Although the advent of computational fluid dynamics software and ever-increasing computing speed and power allows for the generation of complete velocity maps for complex reactor analysis problems (given sufficient experimental data), it is often only necessary to know how long individual molecules stay within the reactor, i.e. the distribution of residence times of defined sub-divisions of the flowing fluid. This approach to incorporating the effects of flow on reactor dynamics is called ***residence time distribution modelling***. This information can be obtained using a stimulus-response experiment (Levenspiel, 1999).

2.3.1.1 Residence time distribution modelling

In a continuous reactor operated at (or approaching) steady state with non-ideal flow properties, the flow patterns cannot be satisfactorily described with either a completely mixed tank reactor or plug flow reactor model. The degree of deviation from an ideal model of flow can be determined by applying a disturbance to the input into the system, and measuring the response in the output of the system. A common method is to apply a tracer to the reactor feed in either a pulse signal or a step input signal. Levenspiel (1999) recommended that any material that can be detected and which does not disturb the normal flow pattern in the reactor can be used as a tracer. By measuring the concentration of tracer that appears in the outflow over time, C_{out}, it is possible to calculate the exit age distribution E_t to be equal to the normalised exit concentration curve:

$$E_t = \overline{C_t} = \frac{\int_0^t C_{out}\, dt}{\int_0^\infty C_{out}\, dt} \qquad \textbf{Eq. 2-12}$$

If the tracer used can be assumed to have the same exit age distribution, or residence time distribution (RTD) as other liquid or dissolved components, then the form of E_t calculated above applies equally to liquid and dissolved species. Thus residence time distribution of the tracer pulse in the reactor system provides information on residence time distribution of other liquid or soluble components in the system.

In a non-ideal reactor with a homogenous (single phase) conversion process (substrate → product), it is possible to predict the yield of an isothermal process from batch kinetic data and knowledge of the residence time distribution. For a first order reaction, the prediction is unique, i.e. a single solution will emerge irrespective of the flow patterns within the reactor that result in the observed RTD (Danckwerts, 1953). For other orders of reaction kinetics, bounds on the extent of conversion may be calculated by considering the cases of maximum mixedness (i.e. a continuous stirred tank reactor, CSTR) and maximum segregation (i.e. a plug flow reactor, PFR) (Zwietering, 1959). This is due to the

fact that for most non-ideal flow systems there are a range of possible reactor configurations that can give the same tracer response curve; when the reaction kinetics are not first order, different flow arrangements result in substantially different concentration vs. location maps, and hence different average reaction rate (Zwietering, 1959).

2.3.1.2 Contact time distribution modelling

In fluid-solid systems, where reaction occurs due to contact between the fluid and solid phases, there are two new complications. Firstly, the solid phase generally has a significantly different residence time distribution than the fluid phase; secondly, the extent of reaction depends on the distribution of times that packages of fluid spend in contact with solid, called the contact time distribution (Nauman and Collinge, 1968b; a). This depends on the residence time distribution of both the solid and the liquid phases, but cannot be adequately described by either. The contact time, t_c is defined as the time in which a fluid molecule is in direct contact with the solid phase such that reaction could occur. Thus the contact age is the sum of all contact time acquired by a molecule since entering the reactor. Finally the contact time distribution is the range of contact ages of all fluid molecules leaving the reactor (Nauman and Collinge, 1968b). Nauman and Collinge describe the contact time distribution as *the exact heterogeneous counterpart of the ordinary residence time distribution*.

Each of the authors cited above specify that the approaches they adopt for determining residence time and contact time distribution are valid for *steady-flow systems* with *single isothermal reactions*.

2.3.1.3 Nature of anaerobic solid-liquid interaction

Anaerobic sludge has the capability to form granules with a complex internal structure (**Section** 2.1.7). The size of each granule will affect the rate of diffusion of reactants into and out of the granule and the rate of conversion within the granule. The surface area of the granule will affect the overall rate of surface-based conversions (McCarty and Smith, 1986; Batstone et al., 2006). Further, the distribution of sizes and amounts of granules within the reactor may have a substantial effect on where within the reactor the bulk of the reaction occurs. In addition to granular sludge, there may also be a dispersed sludge phase where solid-liquid reactions may be better described by adsorption models rather than diffusion.

It follows that there is no simple method for predicting contact time distributions in anaerobic systems.

2.3.2 Flow modelling of anaerobic systems

Some work has been undertaken on measuring and understanding flow dynamics in anaerobic reactors. This section presents a brief review of some of the key findings of this work.

Smith and co-workers (Smith et al., 1993) used a range of methods for analysing tracer response curves from a pilot-scale contact process anaerobic digester. Tracer tests were performed on a large well mixed contact digester at different sludge concentrations and different impeller speeds. The aim of this work was to compare so-called *conventional* methods of analysing tracer response curves (single point indices, models describing the degree of dispersion and division of the tracer response curve into regions representing different flow regimes) to a simulation model. These authors

concluded that the pseudoplastic[1] nature of the anaerobic sludge had a significant effect on the mixing characteristics of the anaerobic digester; lower concentration and therefore less viscous sludge exhibited better mixing characteristics than higher concentration sludge. In this work, it was clear that using a simulation model to describe flow patterns allowed more reasonable predictions of increased mixing with increased impeller speed than other tracer response analysis methods.

Lin and Yang (1991) reviewed a number of different studies on UASB systems in which stimulus-response experiments using a lithium tracer indicated that the sludge bed and sludge blanket regions could be modelled as completely mixed zones. In addition, the sludge bed usually exhibited a dead zone[2]. Bolle et al., (1986) showed that inclusion of short-circuiting flows over each of the sludge bed and the sludge blanket improved the ability of the hydraulic model to reproduce experimental data.

Heertjes and Van der Meer (1978) provided a mechanistic description of fluid and solid movement in a UASB process: The fluid flow depends on a combination of gas production patterns, influent distribution at the bottom of the reactor and solid particles distribution. Uneven distribution of influent can cause channelling in the sludge bed, reducing the effective contact time between fluid and solid. These authors also stated that mixing in the sludge blanket can be expected to be very good as a result of free rising gas bubbles. These statements are supported by their experimental results which indicate that the sludge blanket behaves like a completely mixed reactor. However, mixing in the settled sludge bed was described as being less good than in the sludge blanket, with exchange occurring between the sludge bed and sludge blanket for both the solid and liquid phases. This last effect was attributed to gas bubble evolution and solid settling, which create rising and settling eddies between the sludge bed and sludge blanket phases.

2.4 ANAEROBIC TREATMENT OF SEWAGE

Waterborne sewage is conventionally treated in aerobic treatment processes such as activated sludge processes, trickling filters, ponds and oxidation ditches. However, anaerobic treatment is in fact both an older and more recent approach to domestic wastewater treatment. This section contains a review of common anaerobic domestic wastewater treatment systems and new research into the application of anaerobic technology in domestic wastewater management.

2.4.1 Sewage characteristics

Domestic wastewater usually has fairly low strength, with an organic content of between 250 and 1 000 mgCOD/ℓ (Raunkjear et al., 1994; Henze et al., 1997; Seghezzo et al., 1998). However, the wastewater is complex in nature, consisting of a mixture of proteinaceous, fatty and carbohydrate components in particulate and soluble forms, and inert and refractory components (Henze et al., 1987; Raunkjear et al., 1994). Hydrolysis has been identified as the rate-limiting step in the digestion of domestic wastewater because the complex particulate components require slow disintegration and

[1] A pseudoplastic material exhibits a decrease in viscosity with increase in shear rate. Such materials are also called shear-thinning (Perry and Green, 1997)

[2] See footnote on page 24

hydrolysis steps to convert them to simple sugars, amino acids, VFA and H_2 that undergo acidogenesis and methanogenesis. Thus, the management of suspended solids associated with the domestic wastewater is the critical factor in the design of a system for anaerobically treating domestic wastewater (Lettinga et al., 1993; Kalogo and Verstraete, 2001). Suspended solids are hydrolysed slowly relative to other degradable components of the wastewater, and therefore tend to accumulate in the reactor, thereby decreasing reactor volume available for active sludge (Kalogo and Verstraete, 2001).

The organic fraction of domestic wastewater, characterised by the COD measurement is conventionally described in terms of the fractions depicted in **Figure 2.5** (Wentzel et al., 1999):

- Unbiodegradable COD: This fraction contains all carbonaceous material that cannot be degraded by the treatment system in question, and may be further divided into particulate unbiodegradable COD ((UP)COD) and soluble unbiodegradable COD ((US)COD)

- Biodegradable COD: This category contains material that can be degraded in the treatment system and is further subdivided on the basis of the rate at which each fraction may be degraded, i.e. slowly biodegradable COD ((SB)COD) and readily biodegradable COD ((RB)COD). Many authors make a further subdivision in (RB)COD to differentiate between VFA and fermentable COD ((F-RB)COD), where the latter fraction are those components that are easily fermented to VFA before assimilation by micro-organisms.

- Active biomass: This category consists of all micro-organisms that may continue to be active after entering the treatment system

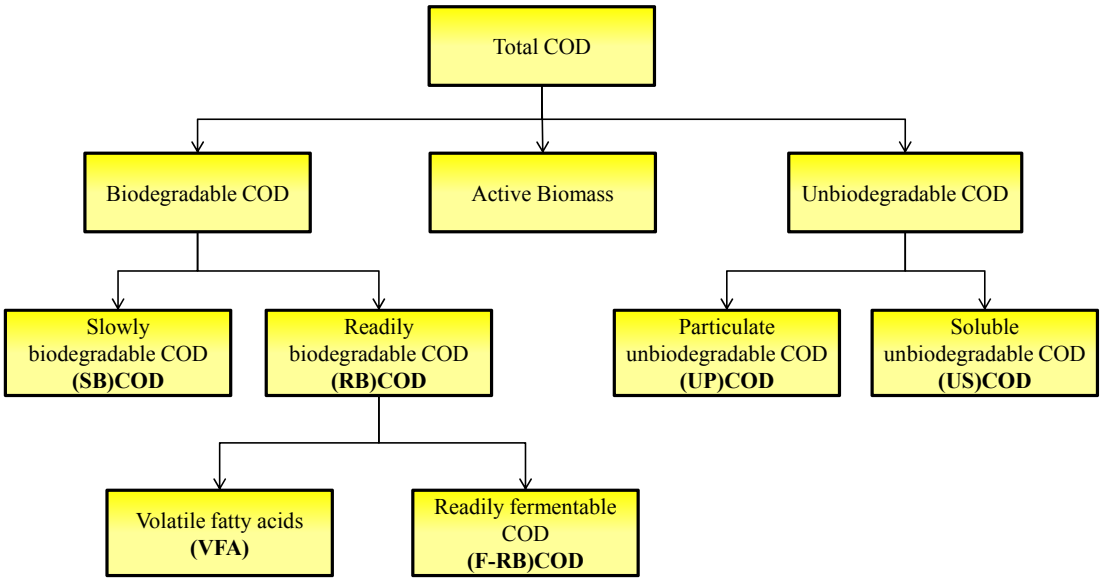

Figure 2.5: Division of wastewater COD into constituent fractions (From Wentzel et al., 1999)

This characterisation was developed for wastewater that was to be treated in a predominantly aerobic system. Classification of biodegradable and unbiodegradable components is therefore defined in terms of their ability to be biodegraded within the residence time of the aerobic system treating them. Ekama et al. (2006) concluded that the *ultimate* biodegradability measurements of a wastewater component made under aerobic and anaerobic conditions have not been shown to be statistically significantly

different. However, the length of the residence time in either system can affect *apparent* biodegradability in terms of how much degrades while in the system.

2.4.2 Septic tank and soak-away systems

There is little formal scientific literature relating to septic tank design and performance, but a large body of information on these subjects exists in the trade literature. Septic tanks are the most commonly used unit for pre-treatment of domestic wastewater in on-site applications. The United States Environmental Protection Agency (USEPA) has published an *Onsite Wastewater Treatment Systems Manual* (USEPA, 2002) which presents a thorough review and discussion of the subject of septic tank and soak-away design and performance. Unless otherwise stated, all further information presented in this section is taken from the USEPA document.

A septic tank system consists of two units. The first unit is the septic tank itself, which pre-treats wastewater by solids and scum retention and partial anaerobic digestion. The second part of treatment occurs in a subsurface wastewater infiltration system, a french drain or evapo-transpiration area where septic tank effluent is infiltrated into the ground via a series of gravel-filled trenches. From here water percolates through the ground, or is removed by evapo-transpiration. Micro-organisms associated with the soil and plant roots, as well as specific plants are able to effect significant nutrient removal from the wastewater. Effluent from these systems is rarely collected for reuse.

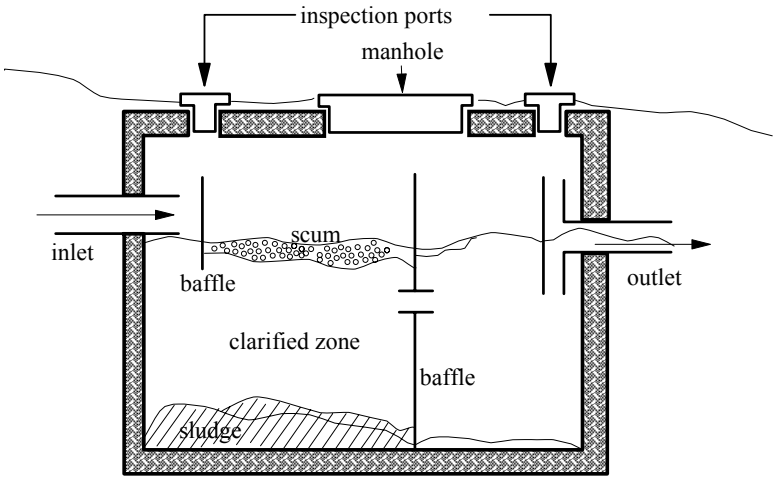

Figure 2.6: Example of septic tank construction showing internal baffle, inlet baffle, outlet tee piece, inspection ports and manhole (after USEPA, 2002)

Figure 2.6 shows an example of a septic tank design. The tank is a covered, watertight rectangular, oval or cylindrical vessel that is usually buried. Dimensions vary, but the tank should be longer than it is wide or high. Primary treatment in the tank is due to wastewater retention under quiescent conditions. Solids and scum from the influent wastewater are separated in the tank by settling or floating. A population of anaerobic micro-organisms develops in the tank which partially digest solids and scum, and to a lesser extent, suspended organic material in the liquid phase. Digestion of scum and solids can result in reduction of up to 40% of retained material, however a slow accumulation of sludge is observed in the tank over a period of between 2 and 20 years, depending on loading.

Anaerobic digestion in the tank generates CH_4 and CO_2 gases that are commonly vented. Wastewater inlet structures in the tank are designed to reduce short-circuiting of incoming wastewater across the tank to the outlet. Outlet structures are designed to retain sludge and scum layers by drawing effluent

from the clarified zone between the sludge and scum layers. The outlet is usually fitted with an effluent screen and/or a tee-piece to retain larger solids that would otherwise be carried out in the effluent to the soak-away, contributing to clogging and eventual system failure. Inspection ports and manholes are provided in the tank cover to allow access for the periodic removal of tank contents, including the accumulated scum and sludge.

Compartmentalised tanks such as that shown in **Figure 2.6**, or tanks placed in series are reported to provide better suspended solids removal than single-compartment tanks alone, although results from different studies vary.

Septic tanks are reported to remove 60 to 80% of non-soluble material in domestic wastewater. Solid and colloidal material is hydrolysed and acidified, producing volatile fatty acids that are only partially converted to CH_4, and exit in the effluent stream. Biological oxygen demand (BOD) removal is typically in the order of 30 to 50% for a septic tank operating at a 48 h retention time. Actual performance of the septic tank will depend on the ambient temperature, operating hydraulic retention time and presence of inert or micro-organism-inhibiting chemicals in the influent. **Table 2.3** presents septic tank effluent characteristics from a number of systems, before secondary treatment in a soak-away or other system.

Soak-away systems provide a degree of wastewater treatment and dispersal through soil purification processes and ground water recharge. The performance is dependent on the treatment efficiency of the septic tank, the method of wastewater distribution and loading to the soil infiltrative surface, and the properties of the vadose and saturated zones underlying the infiltrative surface. Considerable data on the treatment efficiency of soak-away systems are available in the trade literature (USEPA, 2002). High removal rates of BOD, suspended solids, faecal coliforms and surfactants have been observed within a few metres of unsaturated, aerobic soil. Phosphorus and metals are removed through adsorption, ion-exchange and precipitation reactions, although the retention capacity of the soil has a limit that depends on specific soil properties including soil mineralogy, organic content, pH, redox potential, and cation exchange capacity. Pathogen survival rates have been found to vary with a number of factors including initial pathogen load, temperature, humidity and solar radiation.

Table 2.4 shows typical pathogen survival times in fresh water, sewage and unsaturated soil at 20 ºC to 30 ºC (Feachem et al., 1983). Bacterial and protozoan pathogens have relatively short survival times (less than a month), but viruses and helminth eggs (e.g. *Ascaris* spp.) can survive for considerably longer periods.

Table 2.3 Effluent characteristics from septic tanks (before being discharged e.g. to soak-away)

Study Ref	Type	Location	No. of tanks/ homes	No. of samples	BOD mg/ℓ	COD mg/ℓ	TSS mg/ℓ	TKN mgN/ℓ	TP mgP/ℓ	Oil and grease mg/ℓ	*E. Coli* log (cfu/100 mℓ)	pH	Flow ℓ/d
1	cluster	Texas, USA	9			266		29.5	8.2		5.0	7.4	
2	domestic	Wisconsin, USA	7	150	138	327	49	45	13		3.6		
3	cluster	Wisconsin, USA	90		168	338	85	63.4	8.1		6.3	6.8-7.4	140
4	domestic	Wisconsin, USA	33	140-215	132	445	87	82	21.8		5.5		
5	domestic	Oregon, USA	8	56	217		146	57.1			5.4		
6a	cluster	Oregon, USA	11		157	276	36	41		65		6.4-7.2	150-230
6b	cluster	Oregon, USA	Small community		118	228	52	50		16		6.4-7.2	180
6c	cluster	California, USA	330		189	284	75			22		6.5-7.8	150-220
7	domestic	Florida	8	36	141		161	39	11	36	4.1-7.2		
8	domestic	Florida	1	3	179		59	66	17	37	6.0		
9	domestic	SW Cape, RSA			26 (DOC)				14.2		6.6	6.8	
10[1]	domestic	Australia			330		660	250	36		6		

1. Brown et al., 1977
2. University of Wisconsin 1978
3. Otis, 1978
4. Harkin et al., 1979
5. Ronayne et al., 1982
6. Bowne, 1982
7. Ayres Associates, 1993
8. Ayres Associates, 1996
9. Wright, 1999
10. Charles, 2004

[1] 80th percentile values are reported, i.e. 80% of systems sampled had measured values equal to or less than reported value.

Table 2.4: Typical pathogen survival times in water, sewage and soil at 20 ºC to 30 ºC (from Feachem et al., 1983)

	Typical survival times in days	
Pathogen	**in fresh water and sewage**	**in unsaturated soil**
Viruses		
Enteroviruses	<120 but usually <50	<100 but usually <20
Bacteria		
Faecal coliforms	<60 but usually <30	<70 but usually <20
Salmonella spp.	<60 but usually <30	<70 but usually <20
Shigella spp.	<30 but usually <10	
Protozoa		
Entamoeba histolitica cysts	<30 but usually <15	<20 but usually <10
Helminths		
Ascaris lumbricoides eggs	many months	many months

2.4.3 Treatment of domestic wastewater using conventional anaerobic digester technology

In the past, anaerobic digestion has not been considered appropriate as the core technology for the treatment of domestic wastewater since the technology was understood to only be cost effective for high strength applications (Seghezzo et al., 1998; Aiyuk et al., 2006). However, the development of high rate, high efficiency anaerobic reactors such as those described in **Section** 2.2 have allowed the successful treatment of low-strength wastewaters such as domestic sewage. Most success has been had under tropical climate conditions where ambient temperatures are sufficiently high that external heating is not essential for stable operation (Seghezzo et al., 1998; Verstraete and Vandeviviere, 1999; Aiyuk et al., 2006; Leitão et al., 2006).

In their review of anaerobic processes for the treatment of domestic wastewater, Aiyuk et al. (2006) conclude that the rate limitations imposed by the hydrolysis of solids can be overcome by pre-settling the wastewater, possibly with the addition of chemical precipitants or flocculants.

Lin and Wang (1991) reviewed UASB applications and concluded that the presence of finely dispersed solids and particles with poor flocculating characteristics, such as may be expected in domestic wastewater interfere in the granulation process. The proposed reason for poor granulation in these conditions was that micro-organisms preferentially attach to dispersed solids instead of consolidating into granules. Aiyuk et al. (2006) found conflicting reports about the development of granules in UASB reactors treating domestic waster, with a number of references indicating that granulation was observed, while most did not. In the same review, it was reported that granular sludge was formed in expanded granular sludge bed reactors treating domestic wastewater. The granules formed had a distinct layered appearance with coccoid bacteria predominating on the surface, and rod-shaped micro-organisms resembling *Methanosaeta spp.* comprising most of the interior (Kalogo, 2001, in Aiyuk et al., 2006).

Zeeman and Lettinga (1999) indicated that the ability of a digester to develop a stable methanogenic population is dependent on the sludge retention time in the reactor. At a certain temperature, the value calculated for sludge retention time (the average amount of time spent in the reactor by solids, irrespective of whether they are generated by growth of biomass, or originate from the feed) can be used to determine whether methanogenesis will occur or not. These values are used to calculate the required hydraulic retention time for a wastewater of a particular strength. For low sludge retention time, the washout rate of slow-growing methanogenic biomass will be greater than their growth rate; therefore a stable population will not develop. However, at longer sludge retention times, methanogenesis may be expected to occur (Zeeman and Lettinga, 1999).

Zeeman and Lettinga (1999) presented a model for calculating the hydraulic retention time required to give a certain SRT. This model is presented in **Appendix** A5**.**1. SRT values must be selected for a particular temperature to be large enough for methanogenesis to occur. In their review of available data, Zeeman and Lettinga (1999) reported that methanogenesis could be achieved during digestion of manure when the SRT was 100 days, but no methanogenesis was observed for SRT of 50 days.

Lettinga (2001) calculated that the maximum OLR for an anaerobic digester treating domestic sewage at 25 ºC was 1.5 kg $COD_{biodegradable}/m^3$.d. Based on experience with a 64 m^3 UASB treating domestic wastewater at 25ºC, Lettinga predicted that COD removal efficiencies exceeding 80% could be achieved when a hydraulic retention time of 4 hours or more is applied.

Mahmoud et al. (2004) studied the performance of a combined UASB-digester system for the treatment of domestic wastewater. Wasted UASB sludge was treated in a mesophilic digester parallel to the UASB unit and digested sludge was returned to the base of the thermophilic UASB. This system was shown to result in significantly better COD and solids removal than a single UASB with the same OLR.

Ruiz et al. (1998) found that the amount of anaerobic sludge present and the methanogenic activity of the sludge were the main factors limiting efficiency of a UASB treating domestic wastewater.

Treatment of domestic wastewater by anaerobic digestion results in an effluent that is unlikely to meet effluent standards (Verstraete and Vandeviviere, 1999) due to the presence of nitrogen and phosphorus compounds and pathogenic micro-organisms. Aiyuk et al. (2006) report that ion exchange using zeolites may be used to remove ammonium (NH_4^+-N) from the effluent, while pre-treatment by clarifying with ferric chloride, poly-electrolyte and/ or aluminium sulphate can result in substantially reduced phosphorus and pathogen measurements in the effluent.

Table 2.5 presents results from a range of studies on digestion of domestic wastewater in anaerobic digesters

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Table 2.5: Studies using anaerobic technology to treat domestic wastewater for applications with average temperature between 18 ºC and 30 ºC

Reference	Duration [months]	Temperature [ºC]	Reactor type	Reactor Volume [ℓ]	Upflow velocity [m/h]	HRT [h]	OLR [kg COD/m^3.d]	$COD_{in,total}$ [mgCOD/ℓ]	$COD_{in,soluble}$ [mgCOD/ℓ]	SS_{in} [mg/ℓ]	$COD_{out,total}$ [mgCOD/ℓ]	$COD_{out,soluble}$ [mgCOD/ℓ]	SS_{out} [mg/ℓ]
Barbosa & Sant'Anna (1989)	12	19-28	UASB	120	-	4	3.8 [1]	627	357	376	***74***[2]	***78***[2]	***72***[2]
Lettinga et al. (1983)	1	21	UASB	30	-	9	1.4 [1]	520-590	73-75		***57-79***[2]	***50-60***[2]	***30-70***[2]
Lettinga et al. (1983)	17	18-20	UASB	120	-	12	0.83 [1]	248-581	163-376		***72***[2]	***62***[2]	
Lettinga et al. (1987)	9	25	UASB	64 x 10^3	-	6-8	0.8-1.0 [1]	267	95		***75-82***[2]	***75-93***[2]	***70-80***[2]
Vieira & Garcia (1992)	24	18-28	UASB	120 x 10^3	-	5-15	0.5-1.6 [1]	188-459	104-244	67-236	***60***[2]	***70***[2]	***70***[2]
Schellinkhout &Osorio (1994)	>36	24	UASB	3.36 x 10^6	-	5	1.8 [1]	380	160	240	***45-60***[2]	***64-78***[2]	***60***[2]
Vieira et al. (1994)	14	16-23	UASB	67.5 x 10^3	-	3	3.2 [1]	402	515	379	***74***[2]	***80***[2]	***87***[2]
Sayed and Fergala (1995)		18-20	2 stage UASB	42 (4.6)[3]	-	8-4 (2)[3]	1.22-2.75 (1.7-6.2)[3]	200-700		90-385	***74-82***[2]	***73-100***[2]	***86-93***[2]
Behling et al. (1997)	7	30	UASB	55.5	-	7.7	1.21	423		162	59		
Tare et al. (1997)	13	18-32	UASB	12 x 10^6	-	8	3.5 [1]	1183	484	1000	***51-63***[2]	***53-69***[2]	***46-64***[2]
Tare et al. (1997)	11	18-32	UASB	6 x 10^6	-	8	1.2 [1]	404	205	362	***62-72***[2]	***65-71***[2]	***70-78***[2]
Ruiz et al. (1998)		20	UASB	2	-	5-24	0.6-2.9	220-985	63-523	116-424	90-300		20-60
											53-85[2]		***63-85***[2]
Kalogo (2001)	2.7	20-35	UASB	1.2	1.25	2.4	1.4	130	110	15	61		
											54[2]	***55***[2]	***51***[2]
Seghezzo et al. (2001)	24	20-35	UASB	500	0.43	7	0.52 [1]	153	74	79	69	42	27
											55	***45***	***66***
Aiyuk (2004)	15	20-35	UASB	2.1	1	6	0.7	140	115	30	53	45	35
											57[3]	***61***[3]	
Halalsheh et al. (2005)	6	25	UASB	60 x 10^3	0.19-0.22	23-27	1.4-1.6	1612	252	431	632	210	180
											62	***27***	***60***
Halalsheh et al. (2005)	6	18	UASB	60000	0.19-0.22	23-27	1.4-1.6	1419	292	352	816	219	182
											51[3]	***23***[3]	***55***[3]

[1] Calculated from published influent total COD and HRT data

[2] Values in italics are % removal

[3] Terms in brackets refer to the second stage UASB

2.5 THE ANAEROBIC BAFFLED REACTOR

The anaerobic baffled reactor is a high rate anaerobic digester that has been used to pre-treat or co-digest high strength or toxic industrial effluents. Its application in the treatment of low-strength wastewaters has been tested on a laboratory-scale and in a number of full-scale applications for the primary treatment of domestic wastewater. A modified ABR forms the central unit of the DEWATS (DEcentralised WAstewater Treatment Systems) plant that has been implemented to provide low cost domestic wastewater treatment in low income communities in South-East Asia, India, China and Africa (Sasse, 1998; BORDA, 2008).

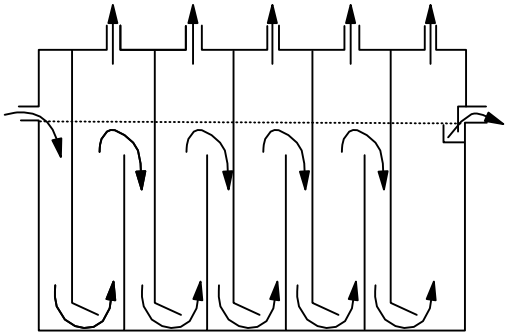

Figure 2.7: Diagram of an anaerobic baffled reactor (ABR) showing hanging and standing baffles. Curved arrows indicate liquid flow, while straight arrows represent gas production.

The ABR is similar in design and application to the upflow anaerobic sludge blanket (UASB) reactor, but it is reported that no special granule formation is necessary for its operation (Barber and Stuckey, 1999). The first baffled digesters to be called ABR were described by Bachmann et al. (1985).

The ABR has alternating hanging and standing baffles (**Figure 2.7**), which divide it into compartments, and direct liquid flow through a series of upward and downward passes. A sludge bed accumulates through the settling of solids in the bottom of each compartment, and the liquid flow is forced through this blanket as it passes under each hanging baffle. Good contact between wastewater flow and active biomass is ensured by this design. In principle, all phases of the anaerobic degradation process can proceed simultaneously in each compartment. However, the sludge in each compartment will differ depending on the specific environmental conditions prevailing and the compounds or intermediates to be degraded (Barber and Stuckey, 1999).

The hydrodynamics and degree of mixing that occur within a reactor of this design strongly influence the extent of contact between substrate and micro-organisms, thus controlling mass transfer and potential reactor performance. Micro-organisms within the reactor rise and settle depending on flow direction (up or down) and velocity, particle settling velocity and gas production. Their rate of movement along the reactor from compartment to compartment is generally slow. The main driving force behind reactor design has been to enhance the solids retention capacity. (Barber and Stuckey, 1999)

The reported advantages of the baffled reactor design are as follows (Barber and Stuckey, 1999):

- The reactor design is simple, with no moving parts or mechanical mixing.

- There is no requirement for biomass with unusual settling properties.
- Sludge generation is low and solids retention time (SRT) is high; this is achieved without the need for biomass to be fixed to media particles or for a solid-settling chamber, although the addition of a stationary phase such as plastic or ceramic packing has been investigated by a number of authors.
- Gas separation devices are not required.
- The ABR has been found to be stable to hydraulic and organic shock loads and the reactor configuration protects the biomass from toxic compounds in the influent.

2.5.1 Research on the performance of the ABR

This section provides a review of research on the ABR that may be relevant to designing systems for and interpreting data from an ABR treating domestic wastewater.

2.5.1.1 Start-up

The overall objective of a start-up protocol in anaerobic digestion is the development of the most appropriate microbial culture for the waste stream to be treated. Specifically, slow growing micro-organisms should not be organically over-loaded, and both gas and liquid upflow velocities should be low to facilitate flocculent and granular sludge growth. The recommended initial loading rate for anaerobic digesters is approximately 1.2 kg COD/m³.d (Speece, 1996). Barber and Stuckey (Barber and Stuckey, 1998) showed that by starting with a long hydraulic retention time (HRT) (80 h) and gradually reducing it in a stepwise fashion for a constant feed concentration, better reactor stability and treatment performance were achieved than in a similar system in which hydraulic retention time was maintained with stepwise increases in substrate concentration. This assessment was based on improved solids accumulation, promotion of methanogenic populations and faster recovery from hydraulic shocks.

2.5.1.2 Hydrodynamic studies of anaerobic baffled reactors

Grobicki and Stuckey (1992) conducted a series of residence time distribution studies in the ABR both under *clean* conditions (no biomass) and when in normal operation for ABRs with 4, 6 and 8 compartments. From visual observations, Grobicki and Stuckey (1992) concluded that at high flow rates (low hydraulic retention time) channelling in the biomass bed had a significant effect on fluid flow patterns. Similarly at high OLRs (low hydraulic retention time) increased substrate supply resulted in increased biological activity and thus increased mixing as a result of gas production. They found that the ABR could be characterised as a series of continuous stirred tank reactors (CSTRs) and that there was a low proportion of dead space[1] (8 to 18 % hydraulic dead space) in comparison with other anaerobic reactor designs. In a water-filled reactor, deadspace was <8 %, but increased to 18% when 8 gVSS/ℓ of sludge was added. Grobicki and Stuckey (1992) differentiated between hydraulic dead space and *biological dead space,* i.e. reactor volume that behaves in a similar way to hydraulic deadspace, but can be attributed to the presence of biological solids in the reactor. The effect of hydraulic retention time on dead space was considered, and it was observed that the biological dead space decreased with increasing hydraulic retention time, while the hydraulic dead space increased

[1] See footnote on page 24

with increasing hydraulic retention time. These authors found that hydraulic dead space was a function of number of baffles and flow rate, and that biological deadspace was related to biomass concentration, gas production rate and flow rate, but that there was no clear correlation between total dead space and hydraulic retention time.

Nachaiyasit and Stuckey (1997c) also undertook residence time distribution studies on laboratory-scale ABRs and confirmed the value of 18% dead space for normal operation, and further found that increasing the amount of reactor biomass by three-fold did not affect the value of dead space calculated in the residence time distribution studies. Langenhoff (2000) undertook tracer tests to determine residence time distribution in 10 ℓ 8-compartment ABRs, fed synthetic low strength soluble and colloidal wastewater. No clear trends were observed in the residence time distribution with changes in hydraulic retention time, feed type or biomass concentration. In all tests, the dead space did not exceed 37 % of the total reactor volume. In Langenhoff's study, no quantitative difference was observed in the amount of dead space determined from the tracer tests for identical reactors with different amounts of biomass, confirming the findings of Nachaiyasit and Stuckey (1997c). It was also concluded that the mixing characteristics of the 8-compartment ABR could be simulated by 8 continuous stirred tank reactors (CSTRs) in series.

Grobicki and Stuckey (1992) found that the number of tanks-in-series (N, the number of CSTRs in series in a hydrodynamic model that would result in the same exit concentration curve as the reactor tested in a tracer test) calculated from the tracer tests was similar to the number of compartments in the ABRs studied. **Table 2.6** reproduces the results of these tests for 8-compartment ABRs operated at different hydraulic retention times and different biomass concentrations. It was concluded that for low values of hydraulic retention time, the calculated number of tanks-in-series closely approximated the actual number of compartments of the ABR. It was also found that the amount of back-mixing inferred from a dispersion model decreased with increasing hydraulic residence time. The authors concluded that the baffles of the ABR inhibit back-mixing, but that there is a large degree of mixing within each upflow compartment. However, it was noted that the downflow section of each compartment was more likely to behave as a plug-flow reactor. The authors indicated that a reasonable approach to modelling the hydrodynamics of the ABR would be a tanks-in-series model with N (number of tanks) equal to the number of compartments, and that the effluent solutes concentration from each compartment should be the same as the average concentration within the compartment. However, low but significant dead spaces were observed, accounting for between 1 and 22% (mean = 9.8%, standard deviation = 8.2%) of the working volume of the ABR.

Table 2.6: Selected results of residence time distribution studies on 8 compartment anaerobic baffled reactors (ABR) from Grobicki and Stuckey (1992)

Run no.	Retention time [h]	Biomass [g/ ℓ]	Gas production [m ℓ /h]	Dead space [%]	Number of theoretical tanks - N
12	20	5.58	774	18.58	10.95
13	20	3.49	691	1.21	6.95
14	10	2.04	739	17.38	8.28
15	10	6.56	1485	7.69	7.13
16	5	6.16	1938	5.4	8.22
17	5	8.5	1804	9.55	8.03

Kennedy and Barriault (2005) performed tracer tests on a water-filled 4-compartment ABR and found that hydrodynamics could be described by 4 or 5 CSTRs in series. These results did not take the presence of solids in an operational reactor into account.

Skiadas and Lyberatos (1998) performed residence time distribution studies on a periodic anaerobic baffled reactor – a modified ABR in which four compartments are arranged to flow around an annulus. This reactor was equipped with many valves allowing flow to be directed in virtually any sequence through the four compartments. When operated in ABR mode (i.e. flow passed sequentially through the compartments), the periodic anaerobic baffled reactor was found to have a residence time distribution equivalent to four well-mixed reactors in series. These results might not be directly extrapolated to a simple ABR since flow around the annulus is likely to result in a greater degree of mixing than in a reactor where there is no enforced direction change in the horizontal plane.

Gopala (2007) performed lithium tracer tests on 8 compartment 10 ℓ bench-scale ABRs fed with low-strength synthetic wastewater. The data obtained were analysed using a two-phase dispersion model from which a total deadspace was calculated to lie between 18.5 and 25.5 % for HRTs of 6, 8 and 10 h, and in a second part of the study, between 22.8 and 34.2 % for HRTs of 6, 8, 10, 15 and 20 h. In the second study, the calculated dead space increased with increasing HRT for HRTs above 8 h.

In conclusion, previous studies have indicated that the baffled design of an ABR results in a residence time distribution that can be approximated by a number N of completely mixed tanks in series where N is the sufficiently close to the number of real compartments of the ABR. However, analysis of UASB reactors (**Section** 2.3.2) indicated that a model which considers two completely mixed zones followed by a plug-flow region is required to adequately describe the residence time distribution data. These studies have all used a completely soluble tracer that may diffuse in and out of biomass flocs. The residence time properties reported therefore apply only to the soluble phase.

2.5.1.3 Investigations into the effect of compartment number

Wanasen (2003) undertook a laboratory-scale comparison between a conventional septic tank design, and septic tanks modified with 1 and 2 internal baffles to create a 2 and 3 compartment ABR. These reactors were fed with a mixture of septage (matured septic tank sludge) and university wastewater (understood to have a lower organic load than typical domestic wastewater). At a hydraulic retention time of 48 h, the baffled septic tanks had approximately the same removal efficiencies (in terms of COD, BOD, TS, and TSS) as the septic tank. However, when operated with a hydraulic retention time of 24 h, the removal efficiency in the conventional septic tank was reduced by up to two-fold compared to the baffled reactors. The three-baffled septic tank removal efficiencies were 10 to 15% higher than observed in the conventional septic tank[1].

[1] Although not directly concluded by the author, these results imply that near maximum stabilisation of biodegradable material was achieved in all units when operated at a 48 h hydraulic retention time, but that the extent of stabilisation decreased more rapidly in the unbaffled units with reducing hydraulic retention time than in the baffled units

A total solids mass balance was undertaken which clearly showed that the baffled septic tanks retain much more solids than the conventional septic tank; 45 to 55 % of solids were retained by the baffled tanks at an hydraulic retention time of 48 h, while only 30% was retained in the conventional septic tank. With a hydraulic retention time of 24 h, and higher TS loading rates, the three-baffle septic tank was able to retain around 65% of the solids, the two-baffle septic tank retained about 40% of the solids, and the septic tank retained only about 15% of the solids (Wanasen, 2003).

Boopathy (1998) digested swine manure in laboratory-scale ABRs. Four laboratory-scale reactors where used which respectively had two, three, four and five compartments. At OLRs of between 4.0 and 8.0 gVS/ℓ.d (approximately 6 to 12 kg COD/m^3.d), it was found that the solids removal, COD removal and CH_4 production rates all increased with increasing number of compartments.

2.5.1.4 Phase separation

The most significant advantage of the compartmentalised structure of the ABR is understood to be the phenomenon of digestion phase separation longitudinally down the reactor with acidogenesis occurring to a greater extent in early compartments and methanogenesis occurring in later compartments. This phase-separation effect allows the reactor to behave as a two-phase system without the control problems and high costs usually associated with two-phase systems (Weiland and Rozzi, 1991).

Uyanik et al. (2001a; b) treated a mid-strength (6 200 mgCOD/ℓ) ice-cream wastewater in 4-compartment 100 ℓ ABRs. The reactor was seeded with granular sludge and the appearance and characteristics of the sludge were monitored over time. It was observed that the largest granules were found in the second compartment, and that CH_4 composition in the first compartment was around 40% but increased to 70% in the subsequent compartments. The pH value in the first compartment (a pH value of 6.4 was reported) was found to remain lower than that of subsequent compartments (neutral pH), while significantly higher VFA concentrations were observed in the first compartment than in later compartments. These results indicate that partial phase separation occurred with acidogenic processes dominating in the first compartment. However, CH_4 production was observed in all compartments indicating that true phase separation did not occur.

The same author (Uyanik et al., 2001b) used a Most Probable Number (MPN) technique to enumerate acidogens and methanogens in each of the compartments. There was no clear difference between the numbers of acidogens counted in different compartments, but it was apparent that the number of methanogens in compartment 1 decreased with time, while the number in the subsequent compartments remained constant. Scanning electron microscopy (SEM) of sludge granules from different compartments showed that hydrogenotrophic (and formate-utilising) methanogens were apparent in the first compartment, while acetoclastic methanogens dominated in later compartments. On dissection, granules were found to have a layered structure with short rod-shaped and coccoid micro-organisms on the outer layer of the granule and filamentous and long rod-shaped micro-organisms in the interior of the granule.

A number of similar studies using different wastewaters, number of compartments and OLRs reported similar results viz. early compartments (usually compartment 1, but also compartment 2 in some studies) had a higher fraction of acidogenic micro-organisms and few methanogenic ones, while granular sludge in later compartments was mostly made up of micro-organisms that were absolutely identified, or tentatively identified based on appearance under scanning electron microscopy as

methanogenic (Akunna and Clark, 2000; Uyanik et al., 2001b; Baloch and Akunna, 2003; Sallis and Uyanik, 2003; She et al., 2006).

Baloch and Akunna (2003) seeded an ABR with granular sludge and fed it with a synthetic glucose-based wastewater at OLRs from 1.25 to 20 kg COD/m^3/d. They observed floating and breaking up of granular sludge in the early compartments of the reactor, while methanogenic sludge in later compartments retained their structure. These authors reported that a *white sticky mass* formed in the sludge of the early compartments. This proved to be bacteria of the *Enterobacteriaceae* genus that can use glucose as a sole carbon source and are tolerant to low levels of dissolved oxygen.

Bell (2002) investigated microbial population characteristics through fluorescent in-situ hybridisation (FISH) analysis of samples drawn from different compartments of an 8-compartment ABR fed a soluble sucrose/ protein feed. Eubacteria were found to predominate in early compartments, while Archaea dominated in later compartments. There was also a distinct shift between H_2, CO_2 and formate utilising methanogenic Archaea in the early compartments and acetoclastic Archaea in the later compartments. Micro-organisms from the genus *Methanosarcina* were only occasionally observed in the first compartment. This genus is usually outcompeted by acetoclastic methanogens from the genus *Methanosaeta* and can only predominate at high acetate concentration (Speece, 1996).

2.5.1.5 Response to hydraulic and organic shock loads

The ABR has exhibited superior resilience to hydraulic and organic shock loads compared to other reactor configurations (Barber and Stuckey, 1999). Grobicki (1989) simulated a hydraulic shock by decreasing the hydraulic retention time from 20 h to 1 h, for a period of 3 h in an ABR operating at an OLR of 4.8 kg COD/m³ on a soluble synthetic carbohydrate / sucrose / protein feed. The reactor returned to its previous COD removal efficiency of in excess of 95 % within 24 h of resuming normal operating conditions. Less than 15 % of the active biomass was lost. In a similar experiment, the OLR was increased to 20 kg COD/m³ and, under these conditions a COD removal efficiency of 72 % was still achieved.

Nachaiyasit and Stuckey (1997b; c) investigated the effect of hydraulic and organic shock loads on a 10-ℓ 8 compartment ABR fed a synthetic sucrose/protein wastewater. Baseline conditions for these experiments were a hydraulic retention time of 20 h and an OLR of 4.8 kg COD/m^3.d (Feed concentration = 4 g COD/ℓ). When the feed concentration was doubled to 8 g COD/ℓ at constant hydraulic retention time, increases in compartment VFA, dissolved H_2 and COD concentrations and decreases in measured compartment pH values were observed in the early compartments, while no significant changes were observed in the last few compartments or outflow, and COD removal efficiency did not change. A further increase of feed concentration to 15 g COD/ℓ resulted in increases in VFA and COD concentration in all compartments and the outflow, but only saw significant increases in H_2 concentrations in the early compartments. Measured pH values actually increased as a result of increased alkalinity production at the higher OLR, indicating that the reactor was not at risk of going sour. These results clearly showed that the compartmentalised design of the ABR resulted in the overall process anaerobic digestion of the feed COD being staged to allow development of an acidic zone (and higher dissolved H_2 concentrations) in the early compartments and a neutral zone in later compartments. As a result, methanogenesis in later compartments was protected from high dissolved H_2 and low pH incidents due to increased organic load, thereby enhancing the stability of the reactor (Nachaiyasit and Stuckey, 1997b).

Experiments to investigate the effects of hydraulic shock loads (increased flow at constant feed strength) were undertaken on the same system (Nachaiyasit and Stuckey, 1997c) and it was found that the COD removal rate decreased from 97 % at a hydraulic retention time of 20 h to 90 % and 52 % when the hydraulic retention time decreased to 10 h and 5 h respectively. However, the COD removal efficiency returned to its baseline level of 97 % after only 9 h after the baseline hydraulic retention time of 20 h was restored. Substantial biomass loss was observed during hydraulic shock loads, and tracer tests indicated that the dead space in the reactor increased substantially (from 18 % to 39 %) during shock hydraulic loads. The authors inferred that significant channelling of fluid flow occurs through the sludge beds during shock hydraulic loads and concluded that this effect helps to reduce the amount of biomass washed out of the reactor and therefore reduce the recovery time after the shock load. Channelling was also understood to result in reduced exposure of the biomass to substrate during these high load incidents, resulting in high outflow COD values, but reduced impact of organic overload on the sludge (Nachaiyasit and Stuckey, 1997c).

Garuti et al. (2004) performed experiments on a 24.2 m^3 2-compartment hybrid ABR supported by laboratory-scale biomass transport experiments on a 9.4 ℓ UASB reactor. Both systems were fed with domestic wastewater at the Biancolina wastewater treatment facility near Bologna, Italy. Measurements of TSS concentrations at two heights on each of the ABR compartments and in the outflow of the UASB were obtained, and sludge bed height in the UASB was measured visually. A mathematical model of sludge bed expansion was developed by considering the sludge column to be divided into 6 height zones and modelling the TSS dynamics in each zone. Predictions of sludge bed height with upflow velocity dynamics were obtained, and it was concluded that short bursts of flow at high flow rates resulted in better overall sludge retention than longer periods of flow at a lower flow rate, (but overall equal average hydraulic load), since the maximum sludge bed expansion achieved during short bursts of flow was less than during sustained low flow periods.

2.5.1.6 Low-strength applications

Several authors have treated low-strength wastewaters effectively in the ABR (Barber and Stuckey, 1999). Treatment of low-strength wastewaters necessarily occurs at low OLRs, except when very high hydraulic loading rates are applied. Thus, dilute wastewaters inherently provide a low mass transfer driving force between the biomass and substrate, reducing biomass activities according to Monod kinetics. As a result, treatment of low-strength wastewaters has been found to encourage the dominance of scavenging micro-organisms, such as *Methanosaeta* species (Polprasert et al., 1992). Speece (1996) cautions that for dilute wastewater, greater attention is required for biomass immobilisation since lower growth rates will be achieved at the same hydraulic loading than for more concentrated systems and thus the sludge washout rate would be equivalently higher. However, other authors (e.g. Barber and Stuckey, 1999) indicate that biomass retention may be good for low-strength treatment due to the low gas production rates and reduced agitation of the sludge bed, suggesting that low hydraulic retention times are feasible during low-strength treatment. Witthauer and Stuckey (1982) (cited in Barber and Stuckey, 1999) observed that biogas mixing was greatly reduced and this resulted in minimal biomass/substrate mass transfer. The authors suggested that when treating dilute wastewaters, baffled reactors should be started-up with relatively high biomass concentrations in order to obtain a sufficiently high sludge blanket and good gas mixing

Langenhoff et al. (2000) studied the performance of the ABR on a dilute synthetic wastewater consisting of soluble and colloidal components (500 mgCOD/ℓ, milk, colloidal rice and dog food) at 35 ºC. High COD removal rates (>80%) were obtained at hydraulic retention times of between 80 h

and 6 h with no significant difference in COD removal observed between reactors fed with soluble and colloidal material.

2.5.1.7 Treatment of solids

Most research on ABRs have involved the use of soluble or colloidal feed material, and much of the research has been undertaken using synthetic substrates to ensure consistency of the feed flow and to facilitate laboratory-scale reactor operation. The behaviour of solids and their affect on ABR performance have been addressed in two separate studies:

- A research group at the University of Illinois (Chynoweth et al., 1980, cited in Barber and Stuckey, 1999) digested sea kelp in baffled reactors. It was reported that significant solids build-up occurred in the first compartment in two weeks of operation, and that this reduced the contact between substrate and micro-organisms, reducing rates of hydrolysis. Manual agitation of the contents resulted in improvements in performance. Accumulated solids were reported to displace biomass in the reactor.

- Boopathy (1998) studied treatment of swine waste (52 gTS/ℓ) in 2, 3, 4 and 5 compartment ABRs. Solids retention time was estimated using chromic oxide (Cr_2O_3) as a stable element marker. The 5-compartment ABR was found to have a solids retention time 6 days longer than the 4-compartment reactor and 12 days longer than the 3-compartment reactor, although the hydraulic retention time was the same for all reactors. The authors concluded that additional baffles resulted in improved particle retention. Better COD removal rates were also observed for reactors with more baffles (**Section** 2.5.1.3). Boopathy (1998) suggested that the improved solids bioconversion results obtained from an ABR treating swine waste compared to conventional digesters was partially attributable to the establishment of a natural filter as a result of addition of whole waste to the first compartment.

2.5.1.8 Effect of temperature

Generally, biochemical reactions double in relative activity for every 10 C° increase in temperature within the active range of the micro-organisms under consideration. Langenhoff and Stuckey (2000) studied the effect of temperature on the performance of a laboratory-scale ABR treated dilute semi-skimmed milk at a hydraulic retention time of 10 h. At 35 ºC, 80% COD removal was obtained, but this reduced to 70% at 20 ºC and 60% at 10 ºC. Nachaiyasit and Stuckey (1997a) found no significant reduction in overall COD removal efficiency when the temperature of an ABR was dropped from 35 °C to 25 °C[1]. Further reduction in temperature, to 15 °C, resulted in a 20 % decrease in COD removal. These studies did not consider long term effects of reduced temperature (such as acclimation or adaptation of anaerobic consortia).

2.5.1.9 Effect of influent alkalinity concentration on ABR performance

She et al. (2006) investigated the effect of influent alkalinity : COD ratio on the in-compartment pH values and overall treatment performance of a 4-compartment ABR treating a sucrose-based synthetic

[1] If the HRT was sufficiently long, then there would be no apparent reduction in COD removal, although the specific rate of COD conversion may have been reduced.

wastewater. Six different influent $NaHCO_3$:COD ratios were tested in the range 0.05 to 0.5 g $NaHCO_3$/g COD. The VFA to alkalinity ratio in each compartment changed from values of infinity (pH value of 4.88) to 0.55 in the first compartment with the lowest values corresponding to the highest influent $NaHCO_3$:COD ratio. Clearly low influent $NaHCO_3$:COD ratios resulted in poor process stability in the first compartment. However, the last compartment VFA : alkalinity ratio never exceeded 0.35, indicating that the reactor as a whole was not at risk of going completely sour. The authors reported that no significant differences were observed in overall COD reduction for operation at the different influent $NaHCO_3$:COD ratios.

Setiadi et al. (1996) experimented with recycle rates in palm oil mill effluent treatment system consisting of an 8-compartment ABR at system OLR of 15.6 gCOD/ℓ.d. These authors showed that, despite the fact that the palm oil mill effluent had a low alkalinity generation potential, the alkalinity supplementation requirements in the ABR could be reduced by recycling the effluent. At a recycle ratio of more than 15, it was possible to maintain the reactor pH at values above 6.8, thereby reducing inhibition of methanogenesis and eliminating the need for NaOH supplementation.

2.5.1.10 Nature of COD in outflow

Barker et al. (1999) analysed the soluble residual COD from two laboratory-scale ABRs treating (i) sucrose/protein/nitrate feed and (ii) dilute semi-skimmed milk. They found that ABR outflow contained a higher proportion of high molecular weight soluble COD than other anaerobic technologies. This was attributed to the relatively longer sludge age that is achieved in ABRs as a result of the excellent sequential sludge retention: long sludge ages result in a higher proportion of cell lysis products (especially cell wall components) appearing in the outflow than in processes with shorter sludge ages. However, low molecular weight organic material was the most abundant type of residual soluble COD for all reactor designs. Barker and Stuckey (1999) reviewed the available literature on soluble microbial products in wastewater treatment systems. These are defined as *the pool of organic compounds that are released into solution from substrate metabolism (usually with biomass growth) and biomass decay* (Barber and Stuckey, 1999), and consist of humic and fulvic acids, polysaccharides, proteins, nucleic acids, organic acids, amino acids, antibiotics, steroids, exocellular enzymes, siderophores[1], structural components of cells and products of energy metabolism. These are found to constitute most of the soluble organic material that is produced during biological treatment.

2.5.1.11 Recovery after inactive period

Manariotis et al. (2002) used a 14.7 ℓ, three-chamber anaerobic baffled reactor (ABR) to evaluate the treatment of low-strength synthetic wastewater (COD of 300 to 400 mg/ℓ) and assess process reactivation after a prolonged period of inactivity. The reactor was inoculated with anaerobic seed and start-up was reported to be immediate. At 26 °C and hydraulic retention times of 24 and 12 h, COD removal averaged 87.2 and 91.0%, respectively, and biogas yield for CH_4 was 0.184 and 0.102 m^3 CH_4/kg COD removed, respectively. The ABR was reactivated after two years without feeding. Response was prompt and removal averaged 85.3% even during the initial 10 d period.

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[1] Siderophores are usually non-ribosomal peptides that are iron-chelating compounds secreted by micro-organisms that scavenge Fe^{3+} from insoluble minerals and chelate it to make it biologically available to the micro-organism (Wikipedia, 2007)

2.5.1.12 Granule formation in ABRs

Akunna and Clark (1999) treated a high strength (9 500 mgCOD/ℓ) whisky distillery wastewater in a 10-compartment ABR and studied the dynamics of the granular sludge produced. They observed that the first two compartments were predominantly filled with a brown slurry-like sludge that spilled into downstream compartments. Poor granulation was observed in these compartments, and eventually no dark granular sludge was to be found in these compartments at all. Other data from this research indicated that acidogenesis was occurring in the early compartments, while methanogenesis was mostly observed in later compartments. These results suggest that the observed acidogens were predominantly non-granule forming, while the methanogenic sludge had a greater propensity for forming granules.

She et al. (She et al., 2006) found that the granules in compartment 1 were grey in colour, but those in later compartments were black. All of the granules observed were made up of a variety of bacilli, cocci and filaments. However, granules from earlier compartments had a higher proportion of short rod-shaped and filamentous bacteria, probably acidogens. Granules from later compartments appeared to consist mostly of micro-organisms similar in appearance to *Methanobacterium* and *Methanococcus spp.*

2.5.1.13 Other studies using ABR technology

Various other investigations have been made using ABR technology:

- Barber and Stuckey (2000a; b) and Bodík et al. (2003) introduced aeration in the latter stages of treatment with an ABR to achieve nitrification and denitrification of the wastewater.

- Wang et al. (2004) digested a glucose-based synthetic wastewater in a 5-compartment laboratory-scale ABR and found that acetate was the predominant intermediate in this process.

- Vossoughi et al. (2002) investigated the effect of different COD : SO_4 ratios on methanogenic activity, and found that methanogenesis could occur simultaneously with sulphate reduction in an ABR.

- Yu and Anderson (1996) used a modified ABR consisting of three sections (upflow – downflow – upflow) with the last two sections packed with plastic media to treat municipal wastewater at ambient temperatures. COD removal efficiencies of between 62 % and 84 % were obtained for hydraulic retention times ranging from 4 to 10 h and CH_4 conversion rates were between 0.09 and 0.12 ℓ CH4/g COD removed.

- Faisal and Unno (2001) successfully treated palm oil mill effluent in an anaerobic baffled reactor, achieving COD removal rates in excess of 90%.

- Bell and Buckley (2003) treated synthetic textile dye effluent in a laboratory-scale ABR. COD reduction efficiencies were consistently above 90% and no dye-breakthrough was observed in 160 days of operation.

2.5.2 Full-scale ABR installations

This section describes full-scale applications of the ABR treating domestic wastewater.

2.5.2.1 Tenjo, Colombia

Two 5-compartment ABR (called anaerobic plug flow reactors) were constructed in a Colombian town, Tenjo (population <2 500) to treat a combined stream consisting of industrial dairy waste and domestic wastewater (Orozco, 1997). These reactors were filled with a high porosity plastic supports to promote anaerobic biofilm growth. The reactors were constructed to be open to the air since the performance of an uncovered reactor in terms of COD removal had been marginally better than that of a covered reactor in pilot-scale experiments. The two full-scale 197 m^3 reactors were operated at a hydraulic retention time of approximately 10 h and removed an average of 70% of COD and 80% of suspended solids from the wastewater over a two month period at an OLR of between 0.45 and 1.96 kg BOD_5/m^3d (average approximately 0.90 kg $BOD_5/m^3.d$) and a design upflow velocity of 3.00 m/h. The reactors had been in operation for more than 3 years at the time of this publication, and the authors reported that granulation was in progress.

2.5.2.2 Biancolina, Italy

A hybrid *an*aerobic-*an*oxic-*ox*ic (ANANOX®) system was implemented at Biancolina WWTP near Bologna, Italy. This plant served the village of Biancolina (ca. 350 p.e.). The ANANOX® system consisted of a two-compartment ABR with a third anoxic compartment and a fourth compartment which operated as a sludge trap. Effluent from this unit passed into an aeration tank and then to a settling tank. A portion of the nitrified and therefore nitrate-bearing supernatant from the settling tank was returned to the anoxic compartment of the baffled reactor. Each of the ABR compartments had dimensions 2.80 m × 1.42 m × 2.05 m i.e. with a compartment volume of 8.15 m^3. Overflow between the compartments was carried by six PVC pipes directed to distribute the flow evenly over the bottom of the subsequent compartment. Waste anaerobic sludge was withdrawn from the bottom of the compartments and discharged to a thickening tank. Approximately 12.5 m^3 of screened degritted wastewater was fed to the plant daily (Garuti et al., 2001).

Biomass concentration in the anaerobic compartments was maintained at low values to prevent biomass washout. Feed with an average COD concentration of ca. 600 mg/ℓ was supplied intermittently to the ABR giving a maximum upflow velocity of around 2.5 m/h. Total COD and TSS removal across the ABR was 31.2% and 45% respectively at the end of a 4 month test period. The ABR in this system is a pre-treatment device and was not designed to achieve complete COD removal. The effluent from the entire ANANOX® plant showed 95% COD removal (Garuti et al., 2001).

2.5.2.3 DEWATS system

DEWATS (DEcentralised WAstewater Treatment Systems) consist of hybrid anaerobic/aerobic systems for community based sanitation. The actual configuration of the system varies according to wastewater quality and effluent quality requirements as well as locally available materials of construction. These systems should be easily managed and maintained under local conditions, and operate without energy input (BORDA, 2008). Four treatment steps are included (**Figure 2.8**):

- Sedimentation and primary treatment
- Secondary anaerobic treatment in fixed bed filters or *baffled septic tanks*
- Secondary and tertiary aerobic/anaerobic treatment in constructed wetlands (subsurface)
- Secondary and tertiary aerobic/anaerobic treatment in ponds

- BORDA reports that DEWATS systems serve over 250 000 inhabitants of over 150 cities in Africa and Asia (Kreutzer, 2008).

- A baffled reactor is described in the DEWATS design handbook as a suitable secondary treatment for *all kinds of wastewater* but preferably those with a high fraction of settlable solids and a small COD/BOD ratio (Sasse, 1998). BOD removals of 70 to 90% are expected in anaerobic filters or baffled reactors in a DEWATS system (Sasse, 1998). The baffled reactors implemented in DEWATS systems have a minimum of 4 compartments and are designed to have an upflow velocity not exceeding 2 m/h. The recommended OLR is less than 3.0 kg COD/m^3.d.

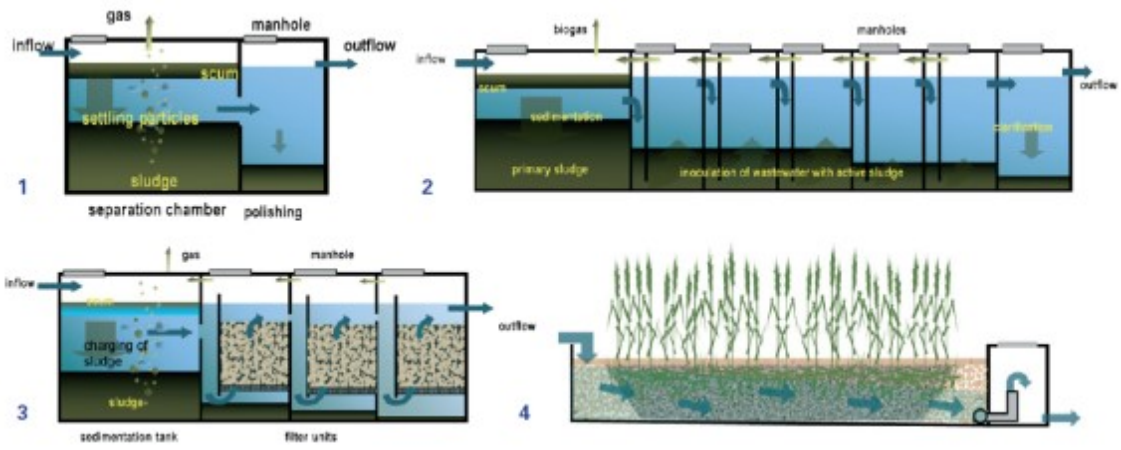


Figure 2.8: Main DEWATS units for wastewater treatment: (1) Settler; (2) Anaerobic baffled reactor; (3) Anaerobic filter and; (4) Planted gravel filter. (Reproduced from BORDA, 2008, with permission)

Hydraulic retention times are designed to be not less than 8 h. A settling compartment is implemented at the beginning of the DEWATS baffled reactor, with a submerged outlet to the next compartment so that scum is retained. These units are reported to require 3 months *maturation* (start-up period) and desludging at similar intervals to septic tanks. Sasse (1998) described the baffled reactor as *poorly known and little researched* and that the microbial dynamics are not well understood. The baffled reactor is usually followed by horizontal filters with constructed wetlands for pathogen and nitrogen removal.

2.5.3 The University of KwaZulu-Natal pilot-scale ABR

Four master level projects were undertaken on the design and performance of the UKZN pilot-scale ABR. Three of these have been finalised, while the last is currently still in preparation. The major findings and conclusions of these studies form the groundwork from which this thesis begins and are presented in this section.

2.5.3.1 Design and start-up of the pilot-scale ABR

The design of the pilot-scale ABR was the basis of an MScEng study undertaken by Dama in the School of Chemical Engineering, University of KwaZulu-Natal (Dama et al., 2002; Foxon et al., 2006; Dama, (in preparation)). The starting point for the design of pilot reactor design was laboratory-scale ABRs that had been used for co-digesting toxic and high strength effluents in WRC project K5/853. These reactors consisted of 8 compartments, and a total working volume of 10 ℓ. Flow between compartments was through a slot in each alternate baffle. As there was no experimental basis for deciding whether the design should be changed, two design parameters were investigated using *computational fluid dynamics* (CFD) modelling techniques to identify improvements for the pilot

reactor design. The CFD study indicated that positioning the hanging baffle to give an upflow-to-downflow area ratio of 2:1 and inclining the bottom edge of the hanging baffle reduced the volume of stagnant areas, and resulted in even flow patterns around the hanging baffle for a water-filled (solids-free) system.

These recommendations were implemented in the 3 000 ℓ pilot-scale ABR. The final design and construction of the pilot-scale ABR is presented in **Section** 3.1.1.

The pilot-scale ABR was initially commissioned at the Umbilo Wastewater Treatment Plant (WWTP) in July 2000 and was seeded with 10 ℓ of anaerobic sludge from the Umbilo anaerobic digesters and filled with screened and degritted wastewater from the inflow channel.

The pilot reactor was operated at Umbilo WWTP for a total of 409 d from 18 July 2000 to 31 August 2001. Since WWTP wastewater was the source of the ABR feed, the wastewater characteristics could not be controlled. The feed rate was controlled by an on-off timer switch on the feed pump, which was set to a target hydraulic retention time. This was stepped from an initial 60 h hydraulic retention time for the first 6 months, to 32 h and finally 20 h.

Dama et al. (2002) reported that outflow COD values of between 50 and 400 mgCOD/ℓ are achieved when the ABR approached steady state. The pH values in the feed, outflow and compartments were monitored, and it was reported that lower pH values were observed in the first compartment than in the last, although the data presented indicated that in the 32 h and 20 h hydraulic retention time periods, there was no clear trend relating the first compartment pH value to that in the last compartment. Alkalinity concentrations were also monitored, and a consistent increase in alkalinity from inflow to outflow was reported. It was further reported that low outlet alkalinity values coincided with poor COD reduction in the ABR.

2.5.3.2 Operation of the pilot-scale ABR

Mtembu operated the pilot-scale ABR from July 2002 to June 2003 (Mtembu, 2006). The focus of this project was to ensure smooth operation of the pilot-scale ABR by engineering improvements to auxiliary features including pump installation and outflow screening. COD, alkalinity, pH, total solids, volatile solids and sludge bed height were monitored. A target hydraulic retention time was set. COD removal of at least 42% was obtained and it was concluded that the reactor was hydraulically overloaded under these conditions. The outflow pH value was invariably lower than the inflow pH value. This observation, as well as worse than expected COD removal, was understood to be an indication of acidification lowering the pH value and inhibiting methanogenesis. However, fairly consistent COD reduction from an inflow value above 700 mgCOD/ℓ to around 200 mgCOD/ℓ was achieved, and a constant or decreasing soluble COD concentration after compartment 3 did not support the hypothesis presented that acetoclastic methanogenesis was significantly inhibited.

2.5.3.3 Microbiological analyses of pilot-scale ABR compartment samples

A study by Lalbahadur (2005) attempted to identify and quantify microbial species in the pilot-scale ABR. Samples from the 8 compartments obtained on 5 different sampling days were studied using 4'6-diamidino-2-phenylindole (DAPI) staining, Fluorescent *in situ* hybridisation and the Polymerase Chain Reaction (PCR) technique (Lalbahadur, 2005). Lalbahadur attempted to explain variations in observations of different micro-organism genera to available measurements of compartment conditions (pH, soluble COD concentration). However, there did not appear to be any correlatable relationship between the microbiological and physico-chemical measurements.

2.5.3.4 Scanning electron microscopy study of sludge samples

Pillay (2006) studied samples of sludge from the pilot-scale ABR operating at two different flow rates using scanning electron microscopy (SEM). He tentatively identified anaerobic micro-organisms based on morphology and concluded that greater concentration and greater diversity of micro-organisms were observed at lower flow rates. He was also not able to identify any *Methanosaeta spp*. in samples (*Methanosaeta spp*. are acetoclastic methanogens thought to be responsible for granule formation). These results are presented in some detail in **Section** 5.7.2.2.

2.5.3.5 Pathogen indicator organism removal

Pillay (Pillay, 2006) measured pathogen indicator organisms (*E. Coli*, total coliforms, coliphages and helminth eggs) in inflow and outflow streams of the pilot-scale ABR. Statistically significant removals of all of these indicators were observed, with highest removal rates for helminth eggs. However, the outflow stream still had unacceptably high pathogen loads, and it was concluded that further treatment would be required before effluent could be safely discharged to the receiving environment.

These results are also presented in Appendix A3.3.2

2.6 SUMMARY

This chapter has presented a review of general theory of anaerobic digestion, anaerobic digester technology, anaerobic digestion of sewage and research into the anaerobic baffled reactor. Much research has been undertaken on baffled reactors treating soluble or synthetic wastewaters, but limited work has been performed on the treatment of real domestic wastewater. Further, most studies were have been undertaken at laboratory scale. Therefore experimental work is required to develop an understanding of the application of ABR technology for large-scale treatment of domestic wastewaters in order that they may be adequately designed and maintained for this purpose.

3 PILOT-SCALE ABR INSTALLATIONS AND OPERATION

"As you know, you go to war with the army you have, not the army you might want or wish to have at a later time." – Donald Rumsfeld

This project undertook an experimental study using a 3 000 ℓ pilot-scale ABR treating wastewater at Umbilo and Kingsburgh WWTPs. This chapter is presented in two parts; in the first, the design of the pilot-scale ABR, the construction of the reactor and details of installation configurations during experimentation are reported. In the second part, methods used in the analysis of pilot-scale ABR performance are described.

3.1 THE PILOT-SCALE ABR

All experimental data presented in this thesis were obtained from a 3 000 ℓ pilot-scale ABR treating municipal wastewater.

3.1.1 Pilot-scale ABR design

The design of the pilot-scale ABR was the basis of an MSc Eng study undertaken by Dama in the School of Chemical Engineering, University of Natal. The design of the pilot reactor used in this study was laboratory-scale ABRs that had been used for co-digesting toxic and high strength effluents in a previous WRC project (WRC project K5/853, Bell et al., 2007). These reactors consisted of 8 compartments, and a total working volume of 10 ℓ. Flow between compartments was through a slot in each alternate baffle.

Figure 3.1: 10 ℓ Perspex laboratory-scale ABRs showing inlet, internal baffles, gas vents and sampling ports ((Bell, 2002))

Figure 3.1 is a photograph of the laboratory-scale reactors.

As there was no experimental basis for deciding whether the design should be changed, two design parameters were investigated using *computational fluid dynamics* (CFD) modelling techniques to identify improvements for the pilot reactor design (Dama et al., 2001)). Firstly, the effect of baffle

spacing on flow patterns in a single liquid phase was modelled, and an upflow-to-downflow area ratio (**Figure 3.2**) was selected to achieve uniform low upflow velocities without large dead volumes.

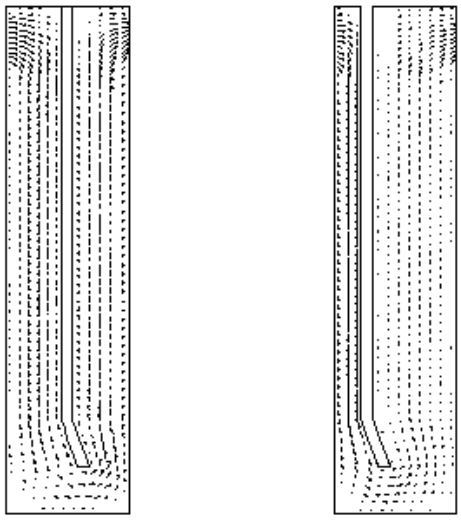


Figure 3.2: Velocity vector profiles obtained for a 20 h HRT using CFD software *FLUENT* for hanging baffle positioning. Profiles for 1:1 (left) and 2:1 (right) upflow-to-downflow area ratios are shown. (Dama et al., 2001)

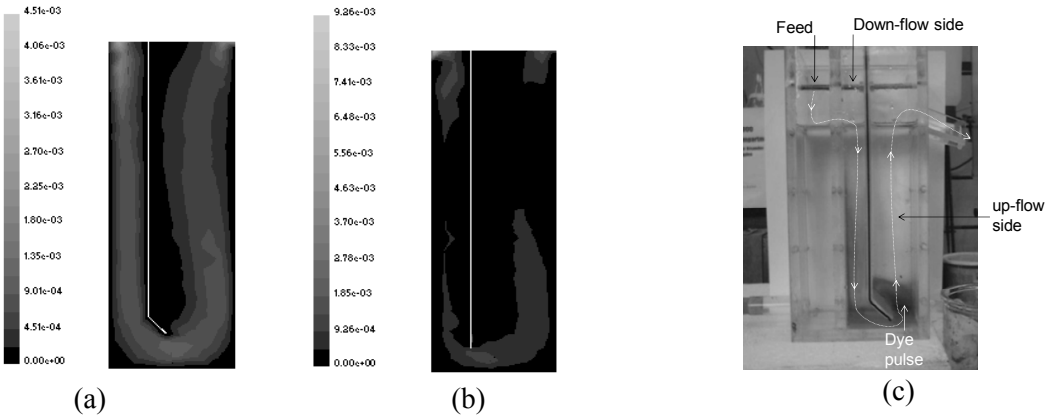


Figure 3.3: Longitudinal section through an ABR compartment illustrating the CFD velocity contours for the two different baffle configurations: (a) angled baffle, (b) straight baffle. Darker colours represent low flow rates. (c) Laboratory scale verification of CFD results using a dye tracer (Foxon et al., 2006) in a single ABR compartment constructed of perspex. The image shows a dye pulse moving around the bottom of an angled hanging baffle. Photographs of the progress of the dye pulse were similar to dye pulse trajectories simulated in *FLUENT*.

The velocity vector profiles along a transverse plane for the two baffle positions are presented in **Figure 3.2**. The magnitude of the velocity is indicated by the length of the velocity vector i.e., the longer the arrow, the greater the velocity. An upflow-to-downflow area ratio of 2:1 resulted in a fairly uniform distribution of flow relative to a ratio of 1:1. Increasing the upflow area resulted in a further increase in channelling and dead-space in the upflow region. These results are valid for a single liquid phase without gas or solids effects, and were not expected to be the same as in a reactor with all three phases present.

A CFD model showing the effect of angling the bottom of the hanging baffle was attempted. **Figure 3.3** shows the flow contours around an angled and straight baffle. It was found that the angled baffle

resulted in more even flow distribution and reduced dead-space. CFD tests were visually reproduced using a single compartment laboratory-scale ABR and dye in water (**Figure 3.3 c**).

The Fluent simulations were used to assist in selection of design features of the baffles. They were not intended to be used in prediction of flow patterns in an ABR treating domestic wastewater since the presence of two additional phases (solid and gas) can substantially alter flow patterns, as compared to clean water flow patterns.

3.1.1.1 Construction of reactor

The pilot-scale ABR was designed to have a total working volume of 3 000 ℓ. The hanging baffles were attached to the top of the reactor to separate the headspaces of subsequent compartments. The heights of the standing baffles were reduced across the reactor so that each subsequent compartment had a slightly lower level then the previous one. Diagrams of the pilot-scale reactor are shown in **Figure 3.4** and **Figure 3.5**. The pilot reactor was constructed of mild steel.

The walls and baffles were laser cut from mild steel sheets and welded together to form gas-tight compartments.

The following sampling ports were included in the reactor design and construction (**Figure 3.5**) (Dama, (in preparation)):

- 4 or 5 × 25 mm diameter ports on the upflow side of each compartment on one side of the reactor. Galvanised ball valves were attached to the top and bottom port of each compartment for sampling. Galvanised plugs were used to seal the other ports.

- A 75 mm diameter port at the bottom of each compartment to facilitate emptying of the compartment. These ports were closed using galvanised 75 mm plugs.

- 75 mm diameter ports on the top of both the upflow and downflow compartments for sampling, fitted with PVC plugs.

- 6 mm diameter gas vents above the upflow area of each compartment for venting and collecting biogas.

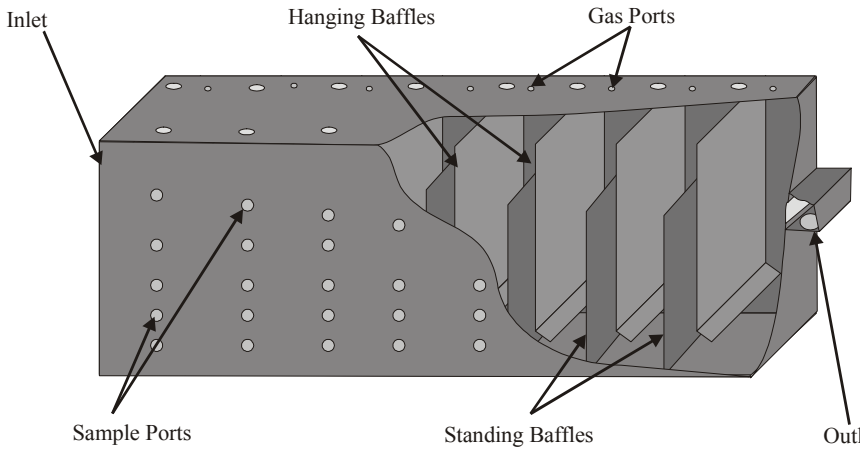


Figure 3.4: Diagram of the pilot-scale ABR with a cut-away to give an indication of the baffle configuration. (Dama et al., 2001)

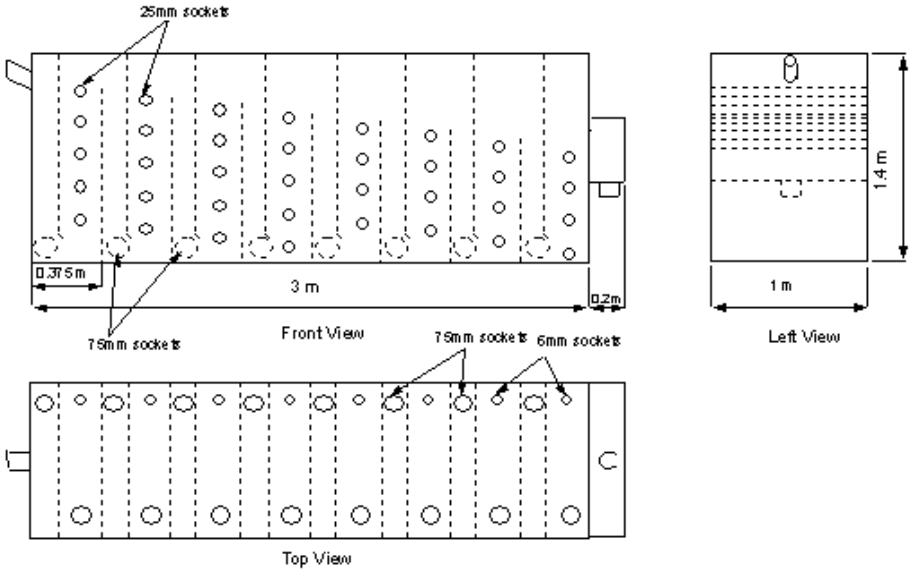


Figure 3.5: Orthographic projection of the pilot-scale ABR (Dama et al., 2001)

3.1.1.2 Construction of feed box

Wastewater feed was fed to the reactor via a feed splitter box. The splitter box consisted of 3 chambers (**Figure 3.6**). Wastewater was delivered to the middle chamber. Approximately 90 % of the flow supplied by the pump was bypassed (left compartment in **Figure 3.6** (a); right compartment in **Figure 3.6** (b)). The rest overflowed into the feed chamber (right compartment in **Figure 3.6** (a); left compartment in **Figure 3.6** (b)). The feed chamber had 3 outlets. A butterfly control valve (FC1) was fitted on the lowest outlet. This valve was used to periodically drain the feed chamber in order to control flow into the reactor. When the control valve was closed, the level in the feed chamber rose until wastewater overflowed through the feed pipe into the ABR. A third (highest) outlet on the feed chamber was supplied to collect overflow in the event of a blockage to the ABR feed line (emergency overflow, **Figure 3.6** (b)). (Dama, (in preparation)).

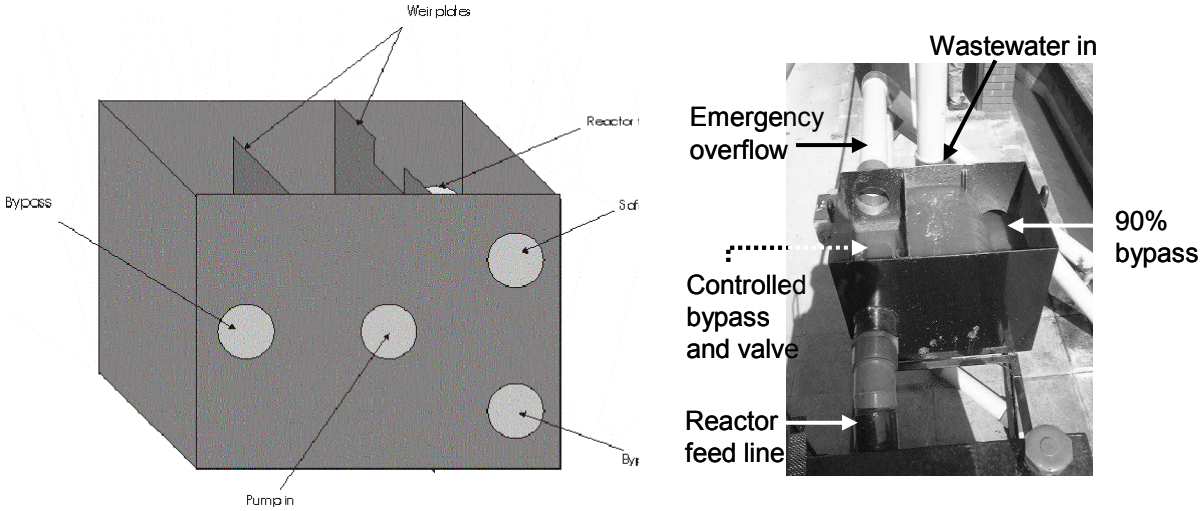


Figure 3.6: Schematic diagram of the feed splitter box installed at the inlet of the pilot-scale ABR (left); and reverse view of the splitter box installed on the ABR (right)

3.1.1.3 Auxiliary equipment

The pilot-scale ABR study aimed to investigate microbiological and chemical performance of the ABR configuration when fed domestic wastewater under controlled feed conditions. All of the

additional equipment used in the pilot-scale ABR installations was there for the purpose of sampling wastewater from a much larger flow than could be handled by the ABR, and feeding it to the reactor in a controlled and quantifiable manner. These included:

- A submersible pump to deliver municipal wastewater to the pilot-scale ABR.
- A pneumatic valve to control air supply to the by-pass valve.
- A compressor to supply air to the pneumatic valve.
- A magnetic flow meter (FI1) to measure and transmit flow rate at the outlet of the last compartment and cumulative flow.
- A programmable logic controller (PLC) to capture flow rate data, calculate feeding/by-passing requirements and control the by-pass valve.
- A timer control switch to control the by-pass valve when the PLC was off-line.

3.1.1.4 Principle of flow control

The reactor outlet passed through a magnetic flowmeter (FI1) which produced a signal that was recorded by a programmable logic controller (PLC). A number of different control algorithms were implemented to achieve a fixed and relatively steady flowrate. The measured flow at the outlet was used to increase or decrease the flow at the inlet by adjusting the timing of the bypass valve (FC1) opening.

During the experimental studies, there were three control regimes vis. timer control, bang-bang control and Proportional Integral (PI) control.

- **Timer control:** Before the PLC was correctly programmed a timer switch was used to open and close the by-pass valve for fixed times in a fixed control cycle. For example, the pump would be set to turn on for 10 seconds in every minute. The timer control system had no mechanism for adapting when the pump delivery rate changed. Pump delivery was erratic due to the heterogeneous nature of the wastewater, particularly the presence of rags that would jam or block the pump impellor. There was thus little control over the amount of wastewater delivered to the reactor. Further, turning the pump on and off so frequently resulted in damage to the pump motor and electrical circuits.
- **On-off control to flow setpoint:** A bang-bang control algorithm was implemented on the PLC. This aimed to control the flow rate to *not* exceed a specified flow rate. This target flow was determined as the flow required to achieve a target hydraulic retention time (T-HRT). This method of control did not allow compensation for periods of high or low flow, and no record was made by the PLC of the actual amount of flow through the reactor. Therefore, the average applied hydraulic retention time (A-HRT) cannot be accurately calculated when bang-bang flow control was implemented. A sample of flow data for bang-bang control is presented in Section 5.2.1.
- **PI control of hydraulic retention time:** A Proportional-Integral (PI) controller was programmed into the PLC using a time-slicing algorithm where, for a fixed cycle of 1 min, the

PLC calculated the fraction of that minute that the bypass valve should be closed in order to achieve a target flow rate and T-HRT. This control regime allowed less variable flow rates to the reactor than had been experienced using timer or bang-bang control, and ensured that the overall flow through the reactor was known. This program also included high and low flow warnings and emergency shut-down loops in the event of excessively high flows through the reactor being recorded. A sample of flow data for PI control is presented in Section 5.2.

3.1.2 Installations

The pilot-scale ABR was installed initially at Umbilo and subsequently at Kingsburgh Wastewater Treatment Plants (WWTP), where it was fed screened and degritted wastewater obtained from the head of works at each plant. Details of installations and operation procedures are included in **Chapters 4** and **5**

3.1.3 Alkalinity in South African waters

South Africa is characterised by many different types of rock structure. The underlying geology of a region will have a profound effect on the water quality, particularly in terms of hardness, pH, metals concentration and alkalinity (Tordiffe et al., 1985). Many areas in the High Veld region are characterised by hard surface waters with high concentrations of calcium, magnesium and bicarbonate. A rapid review of data published by DWAF shows that the total alkalinity in the Upper Vaal region measured in water resources and WWTP final effluent can exceed 250 meq HCO_3^-/ℓ, while sites in coastal areas may have total alkalinity concentrations around 30 meq HCO_3^-/ℓ, i.e. nearly an order of magnitude less acid neutralising capacity (DWAF, 2008). As a comparison, the mean total alkalinity measurement in the potable water produced by Durban Heights Water Works (a potable water plant that serves much of eThekwini Municipality) was 41.0 meq HCO_3^-/ℓ for the months of April and June 2008 (eThekwini Water Services, 2008).

These numbers indicate that water in eThekwini Municipality has a low alkalinity concentration relative to other parts of the country. Thus the influent wastewater to the Kingsburgh and Umbilo WWTP was expected to have low alkalinity concentrations relative to similar facilities in other parts of the country.

3.2 MATERIALS AND METHODS

This section describes the methods used to measure characteristics of operation of the pilot-scale ABR. The pilot-scale ABR was operated over a period of more than 4 years; during this time, the people responsible for sampling, measuring, analysing and collating data changed a number of times. Therefore, the methods used to measure certain properties were not consistent throughout operation of the pilot-scale ABR. Measurement techniques and procedures are presented in this section. Details of the methods are presented in **Appendix A2**. **Appendix A3** presents a table of analyses performed, listing people responsible for supervising operation, sampling, analysing and interpreting data for each analysis and operating period. This information is also included in the data presented in **Annexure 1** in the CD enclosed in the back cover of this thesis.

3.2.1 Sampling

The timing, quantity, method and representivity of a sample have a significant effect on how results of analyses on that sample are interpreted.

3.2.1.1 ABR Inlet and outlet samples

Samples of inlet and outlet flow to and from the ABR were obtained and analysed to determine the extent of treatment achieved by the anaerobic digestion in the ABR on wastewater components. Grab samples were obtained from the feed side of the feed splitter box (**Figure 3.6**) and the outlet pipe just before treated effluent was discharged back to the wastewater channel. For a 5 month period in Phase III, outflow samples were collected approximately one retention time after inflow samples so that a direct comparison between inlet and outlet characteristics could be made. The objective of this scheme was to reduce the influence of time variation of inflow characteristics on appraisal of reactor performance in terms of changes in properties of the flow such as COD, alkalinity and pH.

Samples were collected in 500 mℓ brown glass sample bottles without any headspace.

3.2.1.2 Head of works wastewater measurements

Operators at both Umbilo WWTP and Kingsburgh WWTP obtained hourly samples of wastewater at the head of works that were mixed to produce a time-average composite sample. This composite sample was analysed for key components at municipal laboratories. Samples were composed of sub-samples obtained every hour for 8 h on most working week days. The wastewater sampled was essentially the same as that which entered the ABR, except for the fact that the ABR feed passed through a submersible pump after passing the sampling point at the head of works and before entering the ABR in each case. These measurements were made completely independently of this research project, but were made available to the research team by the municipality.

3.2.1.3 Compartment samples

Liquid or soluble phase components or measurements within compartments provide an indication of how these components or measurements change as wastewater flows through the reactor. They illustrate the spatial distribution of conditions across the ABR, as well as temporal variations of conditions associated with a particular *package* of flow as it passes through the reactor.

It is assumed that most solid components remain within a compartment as a result of settling; therefore, measurements of solid phase components within a compartment are not directly related to the inflow or outflow characteristics at the time of measurement, but rather represent the overall condition of the ABR, particularly in terms of biomass load in each compartment.

During experimentation at Umbilo WWTP, compartment samples were drawn from the sample valves on the side of the reactor. The initial 100 mℓ drawn from each valve was discarded and the subsequent volume collected and stored for analysis.

The relative amounts of sludge and liquid in each compartment were measured using a sampling stick or *core sampler* (**Figure 3.7**). This consisted of a Perspex outer tube with a 50 mm internal diameter, roughly calibrated for height, and fitted with a rubber bung attached to an internal steel rod. The rubber bung was loosened from the outer tube and the internal rod was dropped into the ABR via the 75 mm port on the top of the compartment. The perspex tube was then dropped over the steel rod to land on the bung, capturing a core sample that would be withdrawn from the reactor. Initial sludge and liquid levels were recorded. A 5 min settling time was allowed before settled sludge levels were measured.

During the Kingsburgh experimentation, samples of compartment contents were not obtained from the valves on the side of the reactor as it was believed that conditions near the wall of the reactor did not

represent bulk conditions. Compartment samples were obtained using the core sampler. Once sludge levels had been recorded, the core sampler was balanced in a bucket and the bung worked loose so that the core sample flushed out into the bucket. Bucket contents were vigorously stirred and a sample withdrawn for storage and analysis.

Figure 3.7: Core Sampler filled with compartment 1 sludge (left) and compartment 8 sludge and supernatant (right)

Samples were collected in 500 mℓ brown glass sample bottles without gas headspace.

3.2.2 Sample storage and preparation

Wherever possible, samples were transported immediately to a laboratory for analysis. Samples were stored in a cold room at the University of KwaZulu-Natal (Temperature varying between 4 and 10 ºC) or a refrigerator at Durban Institute of Technology. Where appropriate, samples were coarse filtered through Whatman No. 1 filter paper and micro-filtered through 0.45 µm acetate filter cartridges on-site to reduce biological activity during transport and storage. For VFA measurements, samples were acidified by adding 3 drops of concentrated HCl to 20mℓ of sample. Samples for unstable analytes were transported in a cooler box filled with ice or ice-bricks.

3.2.3 Analytical methods

Where possible all analyses were conducted according to Standard Methods (APHA, 1998). Data on microbiological pathogens, and SEM work were performed by Pillay (2006) as part of his MSc research. DAPI staining and FISH work were undertaken by Lalbahadur (2005) as part of her MTech work.

3.2.3.1 COD

Inflow and outflow total COD concentrations were measured by the open reflux method; filtered or soluble COD concentrations were obtained by filtering samples through 0.45 µm acetate filters and using the titrimetric closed reflux COD method (APHA, 1998).

3.2.3.2 Alkalinity

Alkalinity was determined by potentiometric titration using HCl to an end-point pH value of 4.5.

3.2.3.3 Volatile Fatty Acids

Two methods were employed to measure VFA in samples:

Method 1-HPLC: Small samples (5 mℓ) were obtained from the inflow and compartments 1 to 4, and filtered on-site through 0.45µm acetate filter cartridges. These samples were transported on ice. A sample volume of 1 mℓ was passed through solid phase extraction cation exchange cartridges to extract organic acids, and eluted with a sodium carbonate solution. Pretreated samples were analysed using high performance liquid chromatography for acetic, propionic, butyric, iso-butyric, valeric and iso-valeric acids.

Method 2-Titrimetric: VFA were determined as acetic acid in samples that were analysed titrimetrically for alkalinity using a five point titration according to Moosbrugger (1992).

3.2.3.4 Sulphate

Spectrophotometric measurements of sulphate were obtained on inflow and outflow samples by an accredited municipal laboratory.

3.2.3.5 Phosphate

Spectrophotometric measurements of (ortho-) phosphate were obtained on inflow and outflow samples by an accredited municipal laboratory.

3.2.4 Statistical analyses

All statistical analyses were performed in Excel using methods outlined in Davies and Goldsmith (1977). Details of all statistical tests used are presented in **Appendix** A2.

4 EXPERIMENTAL PHASE I: UMBILO WWTP 2000-2001

The reactor was initially commissioned at the Umbilo WWTP. This location was chosen because of its proximity to the University of KwaZulu-Natal and because it had a well equipped laboratory where routine analyses could be carried out.

In this section, the start-up data of the pilot-scale ABR are presented. The data collection cannot be claimed to have been undertaken directly as part of this study as overall management, sampling, analyses and data collection were mostly performed by people who left the project team in 2001 (**Figure 1.2**). During this phase of the experimentation, First Bell and then Dama supervised the operation of the pilot plant with the assistance of Mtembu. Samples for analysis were taken and analysed by members of the Umbilo WWTP staff.

Much of the data that are presented in this chapter have been presented elsewhere (Dama et al., 2002; Foxon et al., 2006) They are presented again here for completeness. The data have been reanalysed and the interpretation and conclusions may be considered to be original for the purposes of this study.

4.1 UMBILO WWTP INSTALLATION

Figure 4.1 shows the installation of the pilot-scale ABR at the head of works at Umbilo WWTP. The compressor that supplied air to the valve was housed on top of the reactor.

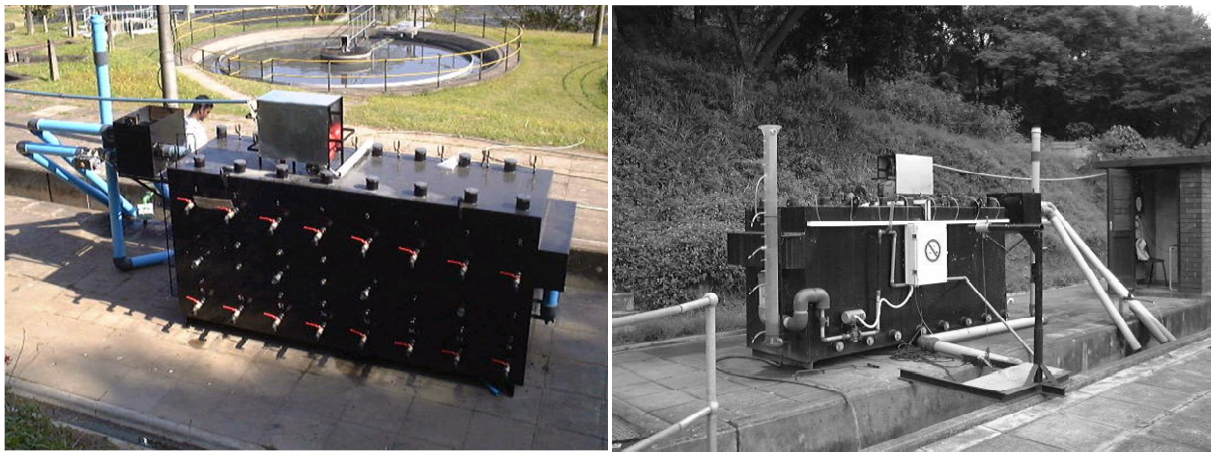


Figure 4.1: Photographs of the front and back of the pilot-scale ABR installed at head of works, Umbilo WWTP

The submersible pump was lowered into the influent channel next to the reactor at the head of the works. The PLC was housed in an enclosure in a control room seen in the right of the right-hand picture in **Figure 4.1**. The pilot-scale ABR was filled with screened and degritted wastewater from the inflow channel. 10 ℓ of dispersed anaerobic sludge from the Umbilo anaerobic digesters was added to the first compartment.

Umbilo WWTP treats a combined industrial and domestic wastewater, where the industrial component arises mainly from nearby textile industries. The wastewater therefore often contains dye effluent and rags. The rags regularly caused obstructions in the pump impellor chamber resulting in no-flow

periods. This problem was reduced slightly by building a triangular flow dispersing frame around the pump inlet, and surrounding the entire pump in chicken mesh (**Figure 4.2**). This reduced the incidence of pump blocking, but was unable to prevent strings and rags from entering and getting entangled in the pump.

Figure 4.2: Submersible feed pump suspended above wastewater channel at head of works, Umbilo WWTP. The pump was fitted with a triangular flow dispersing frame to reduce blockages from strings and fabric in the wastewater.

4.2 OPERATION OF THE PILOT-SCALE ABR: PHASE I

The pilot reactor was operated at Umbilo WWTP for a total of 409 d from 18 July 2000 to 31 August 2001. Data from this operating period were first presented in Dama et al. (2002). No record of the volume treated during this time was kept. For the first 228 d, the flow to the reactor was under timer control (**Section** 3.1.1.4). This resulted in a variable and unpredictable flow. The timer control was set to achieve a T-HRT of 60 h for the first 126 d. On day 127 (22 November 2000), the timer was adjusted to achieve a T-HRT time of 32 h. The timer settings were changed again on day 205 (8 February 2001) to target a 20 h T-HRT (**Figure 4.3**).

On day 228 (3 March 2001), the programmable logic controller (PLC) was brought online (**Figure** 4.3). The control algorithm aimed to control the flow rate to *not exceed* a specified flow rate. This target flow was defined as the flow required to achieve a T-HRT of 20 h. Since no measurement of the total flow was recorded, it is not possible to say what the *actual* mean A-HRT was in this period. However, it is reasonable to say that it was *greater than* 20 h since the flow rate was not allowed to exceed the target flow rate. The uncertainty regarding the actual flow rates and mean A-HRT places a limit on the amount of quantitative information that can be extracted from the data presented. In subsequent descriptions, the phrase *target hydraulic retention time* (T-HRT) is used to describe the operating principle, i.e. that the quoted retention time was targeted, if not necessarily achieved.

Figure 4.4 gives an indication of the type of flow experienced by the reactor under timer control. Rapid oscillations in flow rate were observed over relatively short time periods (<5 minutes) and the amplitude of the oscillations was large, ranging between 0.5 and 3.5 ℓ/min for a target flow rate of 1.7 ℓ/min. These flow conditions therefore represent fairly extreme conditions and the performance of the ABR under these conditions could be regarded as the limit of the ABR performance range.

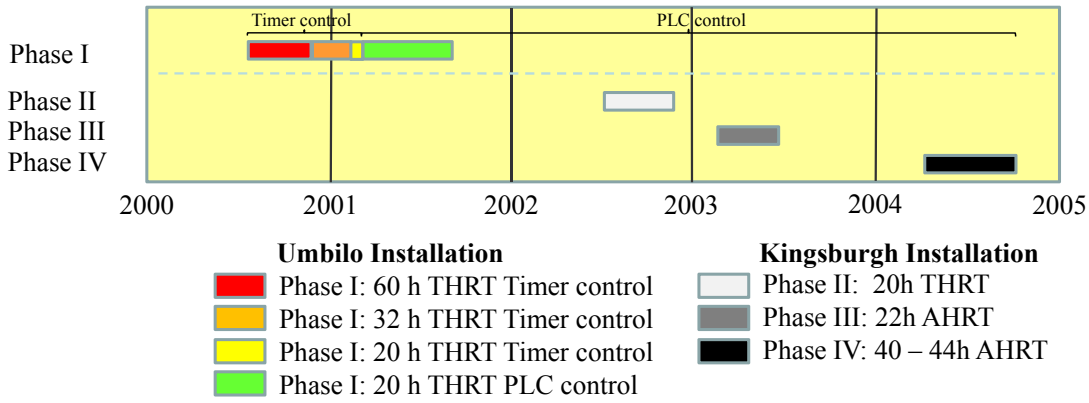


Figure 4.3: **Pilot-scale ABR project time line showing different phases of operation included in this study at Umbilo and Kingsburgh WWTP. T-HRT: Target hydraulic retention time; A-HRT: Average calculated hydraulic retention time.**

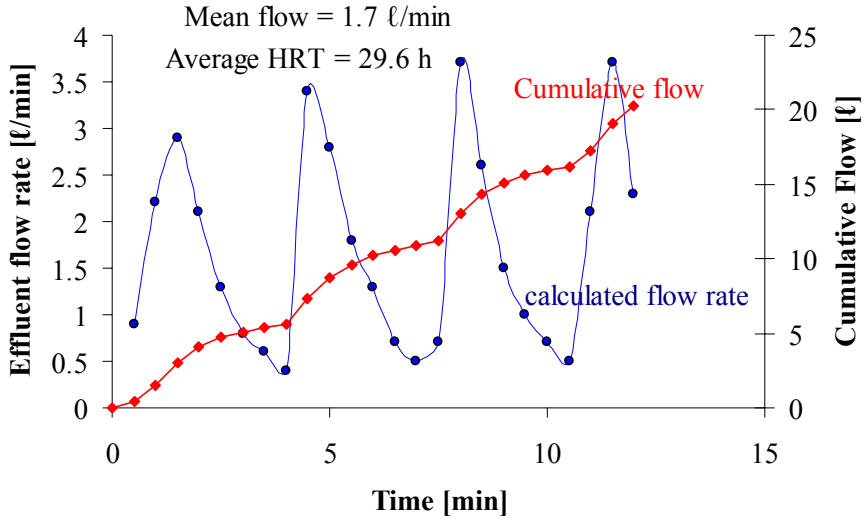


Figure 4.4: **Example of outflow flow rates measured under timer control for a target hydraulic retention time of 32 h. Red points indicate cumulative volumetric flow from the ABR outlet. Blue points show calculated average flow rates for each measurement interval (30 s)**

During operation at Umbilo WWTP, weekly grab samples of the reactor inflow and outflow and from the top and bottom of each reactor compartment were analysed for pH, COD, alkalinity, total solids (TS) and % ash. Volatile solids were calculated from the last two measurements. The inflow and outflow samples were also tested for free and saline ammonia and phosphorus. Physical measurements of the height of the sludge bed were performed on three occasions.

4.2.1 Feed characteristics: Phase I

Table 4.1 presents feed characteristics measured at the head-of-works by Umbilo WWTP staff.

COD concentration of the feed ranged from 151 to 1 845 mgCOD/ℓ during operation of the pilot-scale ABR at this site, with a mean value of 756 ± 36 mgCOD/ℓ in winter and 648 ± 47 mgCOD/ℓ in summer. This constitutes a mid-strength wastewater. Measurements of sodium, chloride, conductivity and ADMI were higher than would be expected for domestic wastewater, and may be attributed to the

significant proportion of dye effluent that is present in Umbilo WWTP wastewater. Rank probit plots of the distribution of inflow measurements for COD, total solids, free and saline ammonia and phosphorus are presented in **Appendix A3**.

No direct measurement was made of the anaerobic biodegradability of the wastewater. However, measurements of BOD_5 in Umbilo WWTP influent were made occasionally. BOD_5 is not an absolute measure of biodegradability since many limiting factors may be present in the BOD_5 test that may not exist in the biodegradation system. Nevertheless, the BOD_5 value may be regarded as a lower limit value for biodegradability. The BOD_5 was between 49 and 58% of the total COD during winter months and between 35 and 44% of total COD during summer months. This implies that at least 49% of the total influent COD is biodegradable during the winter months, while at least 35% is biodegradable during summer, but that the ultimate biodegradability is probably higher than these values. The difference between winter and summer months is most probably attributable to the increased temperature of sewerage resulting in a greater extent of biodegradation of COD during transit in the sewer before reaching the WWTP. However, water flows (dilution) and seasonal industrial activities (e.g. annual shutdown during summer) may also have contributed to this effect.

Table 4.1: Characteristics of degritted wastewater fed to ABR at Umbilo WWTP head-of-works. Data are presented as mean value ± 95 % confidence interval on the mean [min, max] (number of observations)

Determinand	Units	Winter	Summer
COD [*1,2]	mg/ℓ	756 ± 36 [287, 1 845] (157)	648 ± 47 [151, 1 255] (108)
BOD_5	mg/ℓ	298 ± 42 [260, 320] (4)	346 ± 130 [210, 500] (5)
BOD_5/COD*	%	{49, 58}[3]	{35, 44}[3]
Total solids	mg/ℓ	1 141 ± 60 [682, 694] (56)	1 077 ± 98 [486, 1 875] (41)
VS *	mg/ℓ	567 ± 41 [251, 998] (55)	494 ± 47 [166, 825] (40)
TSS *	ml/ℓ	16 ± 2.7 [1.5, 35.0] (30)	12 ± 2.1 [2, 23] (22)
Alkalinity *	mg/ℓ $CaCO_3$	234 ± 7.6 [96.0, 424.0] (158)	190 ± 8.7 [66, 389] (113)
Total Kjeldahl Nitrogen	mg/ℓ N	44 ± 4.8 [29.0, 58.] (12)	40 ± 10 [21, 68] (9)
Free and saline ammonia*	mg/ℓ N	25 ± 0.6 [13.0, 33.0] (159)	20 ± 1.0 [3.2, 40] (112)
Total phosphate	mg/ℓ P	6 ± 0.8 [2.7, 18.0] (56)	7 ± 1.0 [1.1, 14] (40)
Conductivity	mS/m	132 ± 7.5 [59.0, 254.0] (113)	134 ± 10 [20, 290] (113)
pH	(median value reported)	7.0 [6, 9.2] (156)	7.0 [6.4, 7.8] (115)
Sodium	mg/ℓ	170 ± 23 [41, 516] (58)	152 ± 23 [30, 294] (42)
Chloride	mg/ℓ	164 ± 13[43, 654] (158)	170 ± 18 [11, 493] (113)
ADMI *		200 ± 13 [84, 334] (58)	228 ± 16 [162, 390] (43)

[1] Asterisk denotes analytes for which there is a significant difference ($P<0.05$) between values recorded for Summer and Winter seasons.

[2] COD values for winter and summer were found to be significantly different using a t-test ($P<0.05$). However, the data did not have a normal distribution. A rank-sum distribution-free test was used to confirm that the two data sets were in fact significantly different.

[3] Ratio of BOD:COD is presented as a 95% confidence interval calculated using Fieller's theorem (**Appendix 4.4**)

4.2.2 Hydraulic and organic loading rates during Phase I

The hydraulic retention time (HRT) of a system provides an indication of the amount of time that fluid flowing through a treatment system resides in the system. Hydraulic loading rates (HLR) report how much volume of flow is applied to a treatment system *per unit volume* of the treatment system, while organic loading rate (OLR) calculates how much organic material is applied to the system *per unit volume* of the system.

4.2.2.1 Interpreting HRT during Phase I operation

It is impossible to calculate the hydraulic and OLRs accurately for Phase I operation due to lack of adequate flow data. Considering **Figure 4.4**, it appears that the project manager at the time calculated the HRT by first determining flow rates (blue data in **Figure 4.4**) for each measuring interval from cumulative volume of outflow data (red data in **Figure 4.4**) and then applying Eq. 4-1 to determine HRT:

$$HRT = \frac{\text{Reactor Volume}\,[\ell]}{\left(\frac{\sum \text{calculated flow rates}}{\text{number of measurements}}\left[\ell/h\right]\right)} \qquad \textbf{Eq. 4-1}$$

Eq. 4-1 returns an HRT value of 30.8 h from the sample data presented in Figure 4.4, which is close to the target value of 32 h. However, the average applied HRT for this data is more accurately calculated from the cumulative volume of outflow data according to

Eq. 4-2 and gives a HRT value of 29.6 h.

$$HRT = \text{Reactor Volume}\,[\ell] \Big/ \left(\frac{\text{Cumulative volume of effluent}\,[\ell]}{\text{time over which measurements were made}\,[h]}\right) \qquad \textbf{Eq. 4-2}$$

In addition it was observed that the average flow rate could change significantly with time depending on whether there were any objects (e.g. string) interfering with the pump inlet or not.

The error in the calculation of HRT under-predicted the average loading on the ABR. Variation due to pump performance most probably resulted in lower flow rates than intended. The magnitude of the effects is not known, therefore it is not possible to say whether the hydraulic loading was greater or less than that reported. Thus for timer control of the ABR at Umbilo WWTP (day 0 to day 228, 2000-2001) reported HRT values (and thus organic loading and solids retention) should only be regarded as a rough estimate of the true value. Similarly when the PLC was used to control the outlet flow rate to *not* exceed a fixed value (day 228 onwards) actual values are not known, but target or minimum possible HRT values, maximum possible OLR and minimum SRT values may be estimated.

4.2.2.2 Calculation of loading rates

Table 4.2 presents the approximate OLR for each of the reported HRT values during operation at Umbilo WWTP under timer control and PLC control of outlet flow rate.

Loading rates during start-up were low with a maximum loading of around 0.9 kg COD/m^3.d. These numbers are well below the recommended start-up loading rates for anaerobic digesters (**Section** 2.5.1.1)

Table 4.2: Pilot-scale ABR approximate organic loading rate [OLR] [kg COD/m^3.d] under timer control and maximum possible OLR under PLC control in winter and summer during operation at Umbilo WWTP.

	Timer Control			PLC control
T-HRT [h]	**60**	**32**	**20**	**20**
OLR Winter[kg COD/m^3.d]	0.30	-	0.91	-
OLR Summer [kg COD/m^3.d]	0.26	0.49	0.78	0.78

4.3 ABR OUTFLOW STREAM CHARACTERISTICS: PHASE I

This section presents data relating to the characteristics of the flow exiting the pilot-scale ABR during operation at Umbilo WWTP during Phase I of experimentation. Outflow solids, COD, free and saline ammonia and phosphorus data are presented since these are the indicator determinands used to assess the quality of a wastewater or effluent.

4.3.1 Influence of feeding interruptions on reported results

Samples of outflow were obtained from the outlet pipe as it discharged treated wastewater back into the WWTP. Samples and analyses were performed by staff at Umbilo WWTP. The data presented here are subject to substantial uncertainty since the project manager at the time reported that the pump was liable to trip and /or get blocked and that this would be remedied by the plant staff immediately before samples were taken. Thus the samples taken may not have been representative of normal conditions. Further, it was reported that for a few minutes after feeding was started, slugs of washed-out sludge were sometimes observed in the flow leaving the ABR. Thus it is expected that the frequently high solids concentrations measured in the outflow (occasionally above 50 g TS/ℓ) were a manifestation of transitory high sludge washout immediately after resumption of feeding. It is not possible to absolutely identify whether this was the case or not, however, the ABR outflow data for all determinands affected by solids content presented (e.g. total COD, total and volatile solids) may not be representative of the average outflow characteristics during Phase I. The data are presented nevertheless as certain trends observed therein have assisted in developing a theory of what occurs during start-up of an ABR treating domestic wastewater.

4.3.2 Total Solids

Figure 4.5 shows measurements of total solids in the inflow and outflow of the pilot-scale ABR. For the 60 h and 32 h T-HRT, the outflow concentration of solids rarely exceeded 2 g TS/ℓ.

In the 20 h T-HRT period, the outflow solids concentration showed dramatic increases, usually exceeding the corresponding inflow concentration. This was probably due to the fact that the reactor was often restarted just before sampling. However, this was also true for the lower flow operation and similar high values were not observed. The reason for this difference will become clear when compartment internal dynamics are presented, and may be attributed variously to the fact that the solids load in later compartments was substantially higher in the 20 h T-HRT time period due to the higher OLR and the fact that solids washout was facilitated by the higher flow rates employed in the 20 h T-HRT.

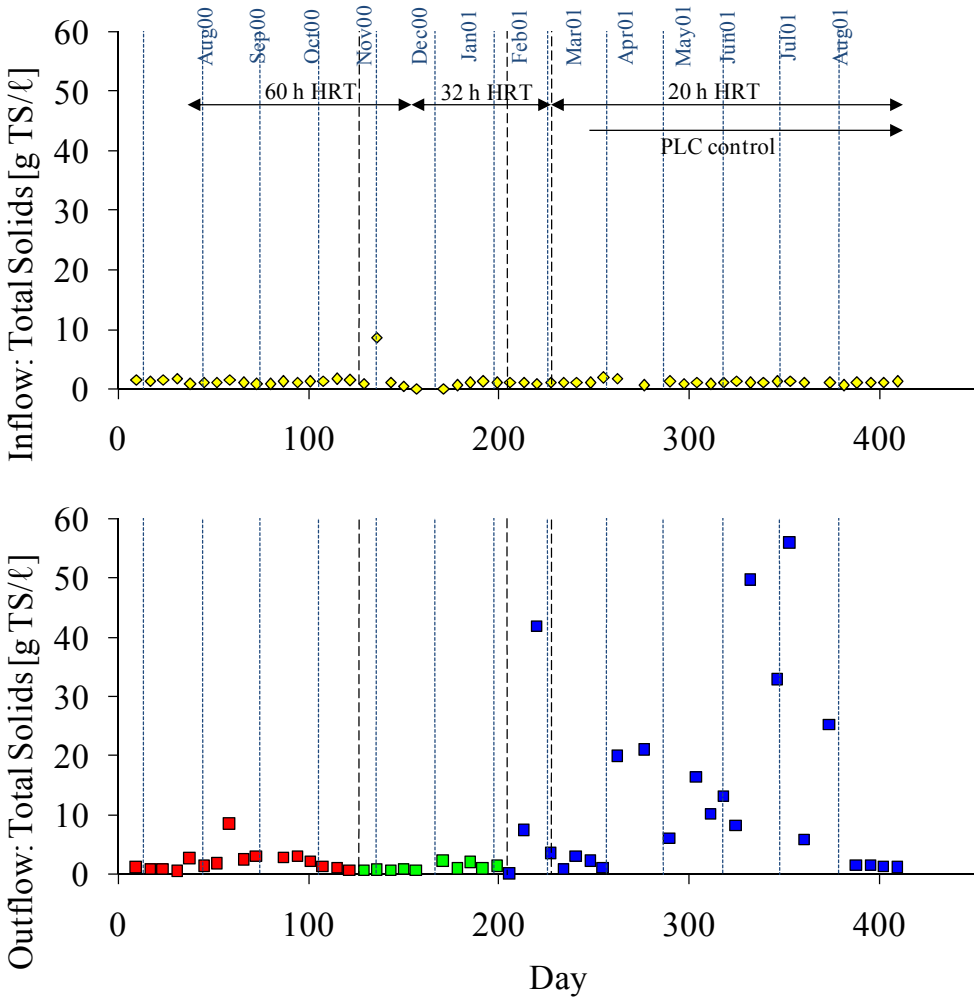


Figure 4.5: **Phase I: ABR inflow and outflow total solids (TS) concentrations. Data for inflow (yellow) and outflow at 60 h (red), 32 h (green) and 20 h (blue) T-HRT (timer and PLC control) are shown.**

It is hypothesised that the low outflow solids concentration in the 60 h and 32 h T-HRT periods was less a function of the good performance of the reactor than an indication that during start-up, solids retention in the compartments of the ABR is the predominant mechanism of treatment, particularly at the low upflow velocities experienced at low feeding rates.

4.3.3 COD

The average COD of the screened wastewater fed to the ABR was 756 mgCOD/ℓ in summer and 648 mgCOD/ℓ in winter (**Table 4.1**). **Figure 4.6** shows the measured COD in grab samples of the inflow to and outflow from the pilot-scale ABR.

Given the fairly wide scatter in measured inflow COD values (95% confidence range of 60 mgCOD/ℓ), it was expected that outflow values would show some scatter, but with a generally decreasing trend during the start-up period. Initially, outflow COD was measured to be near 600 mgCOD/ℓ, but dropped steadily to a value of 121 mgCOD/ℓ on day 129. The flow rate was increased to achieve a T-HRT of 32 h, and an immediate increase in outflow COD was observed. Few data were obtained for the 32 h T-HRT. The average outflow COD in this period was 170 ±54 mgCOD/ℓ (n = 8) (**Table 4.3**).

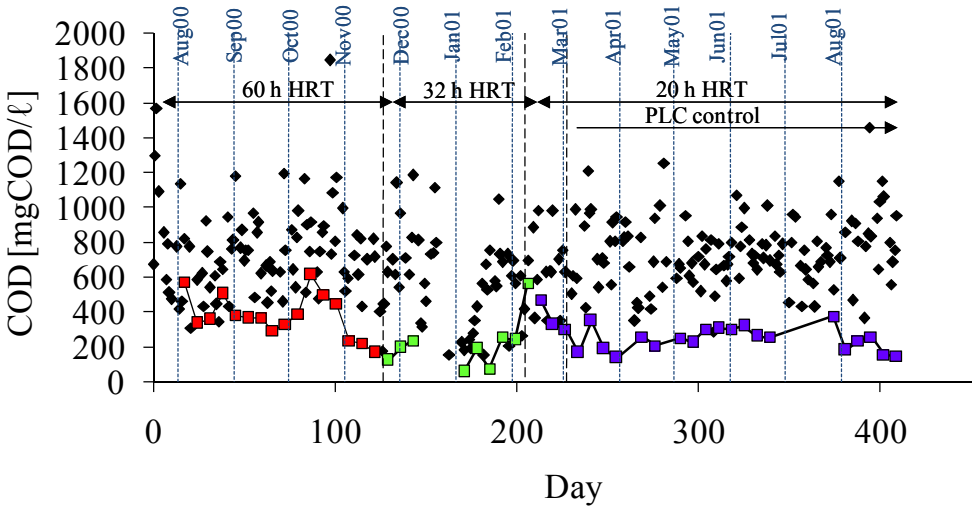


Figure 4.6: Phase I: ABR inflow (♦) and outflow (□) total COD concentrations. Data for 60 h (red), 32 h (green) and 20 h (blue) T-HRT (timer and PLC control) are shown.

The flow rate was increased again on day 205 resulting in a clear increase in outflow COD value to a maximum measured value of 564 mgCOD/ℓ on day 206. By day 234, the COD value had decreased to 165 mgCOD/ℓ. For the remainder of the operating period at a T-HRT of 20 h, the average outflow COD value was 272±40 mgCOD/ℓ (n = 20) (**Table 4.3**). This can be interpreted as an average COD reduction of 66%.

The outflow COD concentration achieved in Phase I was not substantially better than values reported for conventional septic tanks (**Table 2.3**), .i.e. between 200 and 300 mgCOD/ℓ. This result was unexpected, given the improved performance of baffled reactors compared to septic tanks reported in the literature (**Section** 2.5.1.3).

COD removal occurs as a combination of retention of particulate material and biodegradation of biodegradable material. The absolute minimum COD concentration in the ABR outflow stream if all biodegradable material was stabilised and all solid material was retained would be the concentration of the soluble non-biodegradable components in the feed, and any soluble non-degradable components generated by digestion. Additional COD in the effluent stream arises from entrained particulates and undegraded biodegradable soluble material. Thus in order to draw conclusions about the effectiveness of the system, especially in comparison to other systems, the measured outflow COD concentration should be compared to the concentration of soluble non-biodegradable components in the inflow stream. No reliable measure of this value was obtained although it is inferred from BOD_5 data that the total (particulate and soluble) non-biodegradable fraction the Umbilo WWTP wastewater COD may have been as high as 65% in summer and 51% in winter (**Section 4.2.1**). It is also not known what portion of influent particulate material was retained in the ABR by settling hence it is not known whether the *biodegradable COD* removal was better or worse than could be expected of a conventional septic tank.

4.3.4 Free and saline ammonia

Figure 4.7 presents free and saline ammonia (NH_3+NH_4^+) concentrations in the inflow and outflow of the pilot-scale ABR during operation at Umbilo WWTP.

A net increase in free and saline ammonia was observed between the values measured in the ABR inflow and outflow streams. This was attributed to liberation of organically bound nitrogen during anaerobic digestion (**Section** 2.1.1.3) (Speece, 1996).

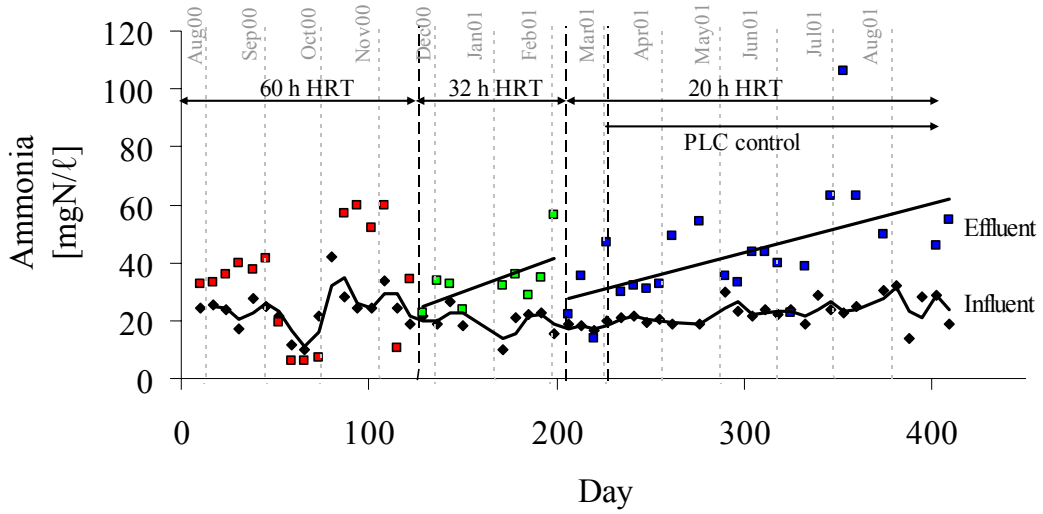


Figure 4.7: **Phase I: ABR inflow (♦) and outflow (□) free and saline ammonia (NH_3+NH_4^+) concentrations. Data for 60 h (red), 32 h (green) and 20 h (blue) T-HRT (timer and PLC control) are shown. Linear trend lines in 32 and 20 h T-HRT periods show increasing outflow free and saline ammonia concentration trend.**

The *amount* of free and saline ammonia released appeared to increase over the 32 h T-HRT period (green points in **Figure 4.7**), then decrease as a result of the change in flow rate on day 205 and subsequently increase again during the 20 h T-HRT period. Regression analysis showed that the upward trend in outflow free and saline ammonia data was significant at the 90% confidence level (P=0.06) the 32 h T-HRT period and highly significant at the 95% confidence level (P=0.006) for the 20 h T-HRT period. These data imply that a decrease in the extent of treatment (i.e. amount of biodegradable material from the feed that is degraded) occurred as a result of the change in feeding rate, but that the extent of treatment increased with time during each of the feeding rate periods.

It was observed that at end of 20 h T-HRT, the outlet free and saline ammonia concentration was often greater than the average inflow Total Kjeldahl Nitrogen (TKN)[1] value measured in the WWTP influent wastewater (a value of 42 mgTKN-N/ℓ is reported in the inflow). This value may indicate that solids that had previously accumulated were digested in the 20 h T-HRT period. This implies that micro-organism actively digested solids at a rate faster than they were supplied in this period, and would account for the apparently increasing extent of treatment that is inferred from the data presented.

[1] Total Kjeldahl Nitrogen (TKN) is a lumped measurement of free and saline ammonia and organically bound nitrogen.

4.3.5 Phosphorus

Figure 4.8 presents inflow and outflow phosphorus concentrations for the three operating periods. Total phosphorus analyses, (achieved by pre-digesting samples) were undertaken on samples from the ABR inlet, while analysis of only soluble phosphate was performed on outflow samples.

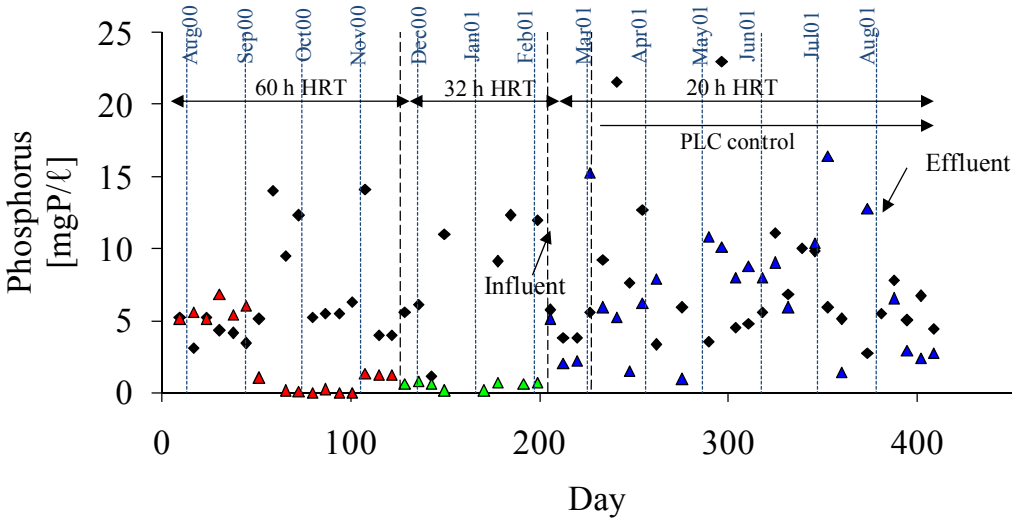


Figure 4.8: Phase I: ABR total inflow (♦) and soluble outflow (□) phosphate. Data for 60 h (red), 32 h (green) and 20 h (blue) T-HRT (timer and PLC control) are shown.

Between day 50 and day 200 outflow phosphorus concentrations were significantly lower than inflow concentrations. This should not be regarded as biological phosphorus removal since a mass balance cannot be calculated from the two dissimilar measurements (soluble and total phosphorus). No significant phosphorus reduction can be expected under ordinary circumstances from an anaerobic system (Speece, 1996). The cause of the low soluble phosphorus concentrations in the outflow from day 50 to 200 is not fully understood. Two possible explanations are:

- Low biomass concentration during the early days of operation resulted in little liberation of phosphorus from organically bound forms, thus low soluble phosphorus content would be observed in the outflow.

- If a substantial portion of the phosphorus in the inflow is associated with particulate material, then low outflow soluble phosphorus content reflects the hypothesis presented in **Section** 4.3.2 that during the early days of operation, the predominant treatment mechanism is retention of solid material as opposed to biological reduction of organic material.

4.4 COMPARTMENT DYNAMICS: PHASE I

In this section the results of analyses on measurements made within compartments are presented.

4.4.1 Height of sludge bed

In previous publications of data from this operating period, solids concentration was excluded from presented results since the samples obtained did not represent average compartment conditions (Dama et al., 2002). However, on closer examination, the data were shown to shed light on the mechanism and rate of sludge build-up in the reactor: sludge builds up in earlier compartments sooner than in later

compartments. Therefore, as start-up progresses, sludge beds will be observed first in compartment 1, then in compartment 2, etc. Thus measurements of total solids near the bottom of each compartment should show an abrupt increase at the time when the sludge bed accumulates to the level of the sampling valve, and this event will occur at a later time in each subsequent compartment.

Figure 4.9 shows values of total solids measured at the bottom sampling valve of each compartment, located 200 mm above the floor of the ABR. Total solids concentrations at the bottom sampling valve of all compartments were initially mostly less than 10 g TS/ℓ; each compartment successively showed an increase in solids concentration at a height of 200 mm represented by hand-drawn curves in **Figure 4.9**.

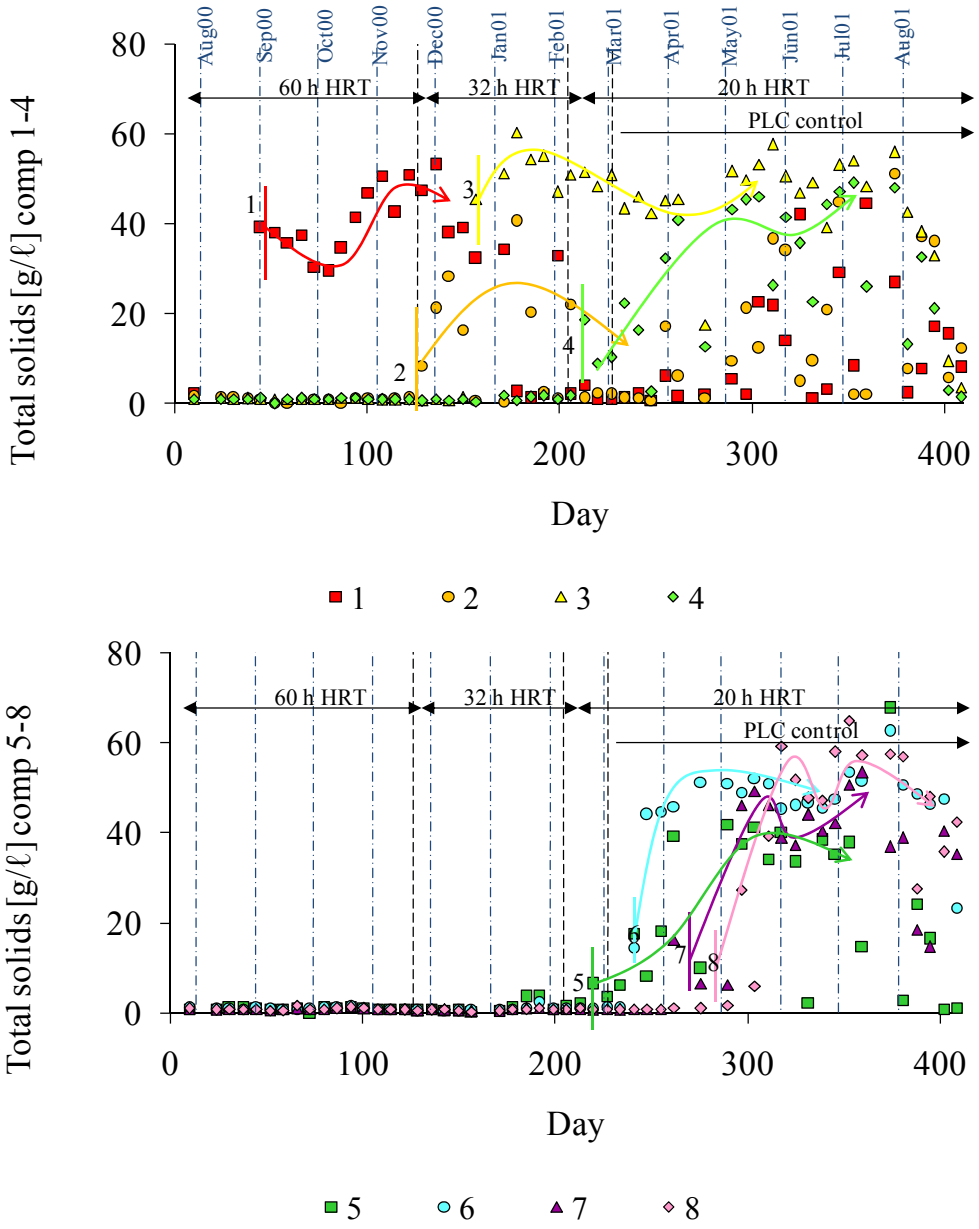


Figure 4.9: **Phase 1: Total solids concentration measured 200 mm above the bottom of each compartment, with hand drawn trends to show the appearance of the sludge level above the sampling point. The numbers 1 to 8 represent compartments 1 to 8.**

Observation of sludge in compartments showed that the upflow side of each compartment may be described as consisting of three zones: a dense, settled sludge bed at the bottom; a fluidised sludge bed; and a relatively clear zone above the sludge. This is similar to the behaviour of solids in a UASB reactor (**Section** 2.2.1) (Lin and Yang, 1991)

It is probable that total solids concentrations that are less than 10 g TS/ℓ indicate that samples have been drawn from the zone above the fluidised sludge bed, while those that are between 10 and 70 g TS/ℓ are within the fluidised bed zone.

To test whether these assumptions are reasonable, the following values were considered:

- The maximum solids concentration measured at the top of each compartment (samples withdrawn from the valve located 5 cm below the liquid level) between day 0 and day 255 was 2.1 g TS/ℓ (The full data set may be found in **Annexure 1** in the CD enclosed in the back cover of this thesis.)

- During operation at Kingsburgh (See **Section** 4.3.2) the maximum TS concentration in the outflow of the ABR observed during Phase 2, 3 and 4 was 0.8 g TS/ℓ, with an average value around 0.4 g TS/ℓ

- Mtembu (2005) calculated the maximum solids concentration of the sludge taken from the lowest port in a batch settling column to vary from 12 to 34 g TS/ℓ during batch settling tests of sludge taken from compartments during the 2002 operating period. These numbers are expected to correspond to concentrations of sludge in an uncompressed bed of sludge.

These numbers support the assumption that values below 10 g TS/ℓ are observed above the sludge bed, while those between 10 and 70 g TS/ℓ are within the sludge bed; i.e. the time at which the total solids shows a step increase to 10 g/ℓ or more indicates the time at which the sludge bed height increased above the sampling port.

The actual height of the sludge bed was not measured for most of this operating period, although a limited amount of data are available; at start-up, there was essentially no biomass in any of the compartments except compartment 1 since this compartment was seeded with a few bucketfuls (approximately 10 ℓ) of anaerobic digester sludge (**Section** 4.1). Samples drawn from the bottom sample valve of each compartment for the first month showed low (<10 g/ℓ) solids concentrations, indicating that the sludge bed had not risen above the sample valve height i.e. 200 mm above the bottom of the ABR.

From **Figure 4.9** and assuming solids concentrations exceeding 10 g TS/ℓ indicate the presence of a fluidised sludge bed, it is possible to see exactly when the sludge bed had built up above the bottom sample valve. For compartments 1 to 8 respectively, this occurred on days 45[1], 129, 157, 213, 220, 241, 262 and 290. All compartments had achieved 200 mm sludge beds by day 300.

[1] The exact date that the sludge level in compartment 1 exceeded 200 mm is not known as there is missing data for compartment 1 only in the first month of operation.

Compartments 4 to 8 only developed 200 mm high fluidised sludge beds in the 20 h T-HRT operating period. The appearances of the sludge beds above the 200 mm sampling port in successive compartments occured more rapidly during the 20 h T-HRT period than at lower loading rates due to the higher OLR. This implies that the rate of hydrolysis of solids is lower than the rate at which they are supplied.

Only one set of sludge bed height data values is available (**Figure 4.10**) from day 127 of operation (at the beginning of the 32 h T-HRT operating period). Two further data sets (in graphical form) were located in an early project report and are reproduced in **Figure 4.11 (a)** and (**b)**, although the values from which these graphs were drawn were not recovered. Of particular interest are the fluidised solids measurements (*fluidised* in **Figure 4.10** and *solids* in **Figure 4.11 a** and **b**). These data show large variations in the amount of sludge in the first compartment, but an increase with time in the amount of solids in later compartments. **Figure 4.9**indicates that sludge bed heights had exceeded 200 mm in the first 6 compartments by day 254, which agrees with the sludge bed height profile presented in **Figure 4.11 b** for this day.

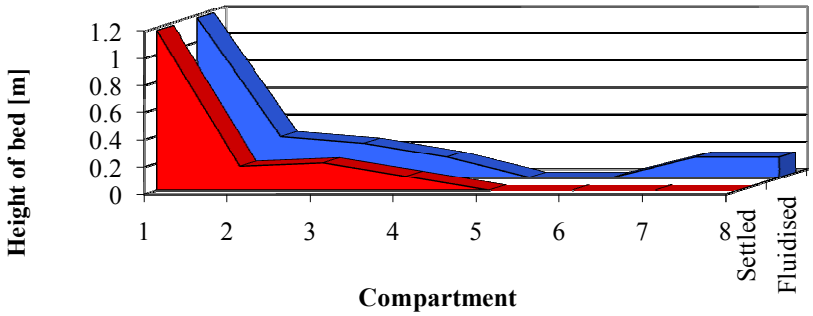
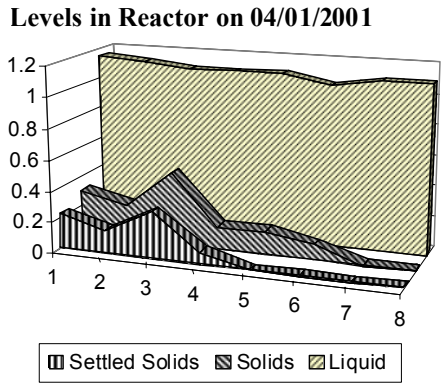
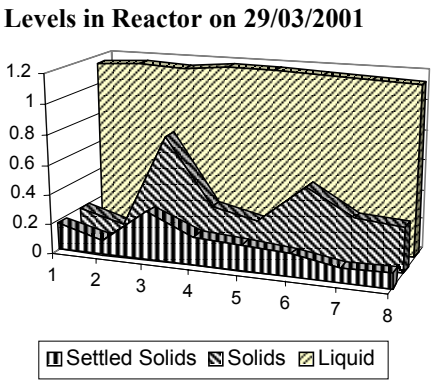


Figure 4.10: Phase I: Sludge bed height immediately after core sampling (blue: fluidised) and after settling for 5 minutes (red: settled) on day 127 during operation at Umbilo WWTP



(a) Day 170



(b) Day 254

Figure 4.11: Phase I: Sludge bed height immediately after core sampling (*solids*) and after settling for 5 minutes (settled solids) on day 170 (04/01/2001) and day 254 (29/03/2001) during operation at Umbilo WWTP (Figures recovered from WRC Project Steering Committee Report by Dama, 28 May 2001)

Another interesting observation is that the amount of solids in compartments 1 and 2 appears to be less in the recovered figures for day 170 and day 254 (**Figure 4.11**) than on day 127 (**Figure 4.10**) and that this is reflected in the lower total solids measurements at the 200 mm level in compartments 1 and 2 at these times (**Figure 4.9**). The agreement between the two independent sets of data lends credibility to the data (which otherwise might have been considered to be excessively scattered) and further indicates that either the change in flow rate, or more probably some accidental high flow incident caused some washout of compartment 1 and 2 sludge to compartments 3 and 4.

4.4.2 COD

Samples of reactor contents were withdrawn from the top sampling valve of each compartment (just below the top liquid level) and analysed for total COD. Analyses of these samples provide some idea of the COD concentration of the overflow from one compartment to the next.

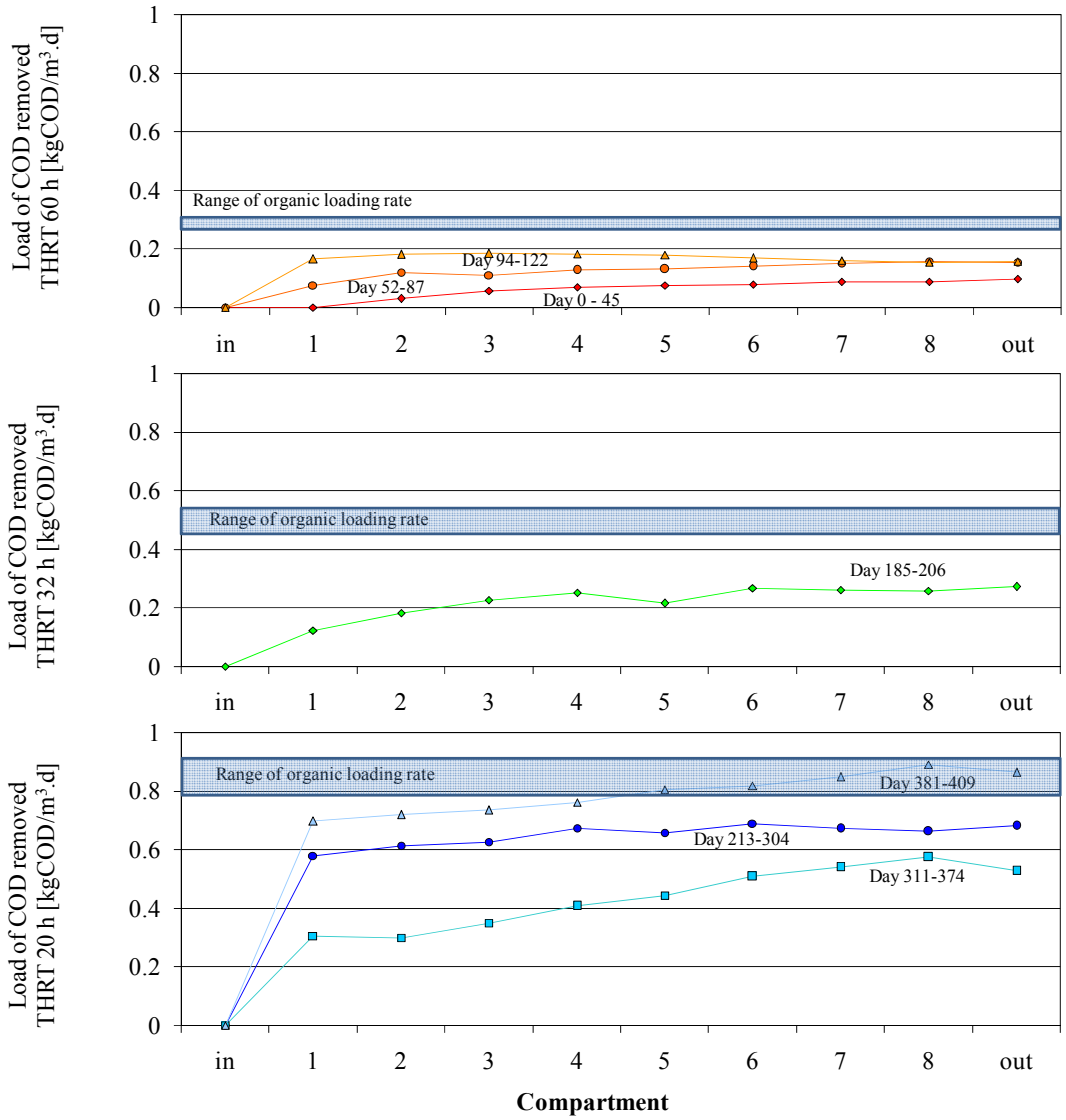


Figure 4.12: **Phase I: Cumulative load of COD removed by the end of each compartment. (Overall load of COD removed from wastewater calculated from the difference between ABR inflow and compartment overflow values) Data have been divided into the different operating periods (60 h THRT period: red/orange data; 32 h THRT period: green data; and 20 h THRT period: blue data) and into time intervals within which similar performance was observed.**

These measurements have been used to determine the relative contribution of each compartment to the total COD removal that occurs in the reactor: **Figure 4.12** presents cumulative COD removal data per compartment; i.e. the cumulative amount of COD that has been removed from the wastewater flow by the time it has reached the overflow out of each compartment, compared to the applied OLR.

No differentiation is made between COD removal by biodegradation and by solids retention. These values were calculated from the difference between the COD concentration at the top of each compartment and the COD concentration of the inflow wastewater, and are averages for the time interval reported. Data from day 129 to 178 were excluded since these days corresponded to the period from December to early January where very high and very low COD loads were observed associated with annual shutdown of the local textile mills. Data from this period showed very high COD removal rates. The full data set may be found in **Annexure 1** on the enclosed CD.

These data provide an indication of where the majority of treatment is occurring at different times during operation of the pilot reactor.

- The increasing removal rate with time is in part due to improving reactor performance, but is also strongly influenced by the OLR with higher OLRs resulting in higher COD load removal rates

- Except for the very first period reported, by far the most COD removal is experienced in the first compartment (>70% of total COD removal). For the 60 h T-HRT period, after the first 50 days, virtually no additional COD removal occurs after the first compartment.

- There are limited data from the 32 h T-HRT period; however, the data presented indicate that overall, a higher COD removal rate is achieved in this period than in the 60 h T-HRT period. It also appears that compartments 2 and 3 made an increased contribution to overall COD removal than in the 60 h T-HRT period. This is supported by sludge bed data (**Section 4.4.1**) where it was observed that the sludge bed rose above the 200 mm sampling valves in compartments 2 and 3 in this operating period.

- Data from the 20 h T-HRT operating period are confusing due to the significantly lower COD load removal in the first compartment in the middle data set from this period: the average COD load removal in the first compartment was approximately 0.3 kg COD/m^3.d from day 311 to day 374 compared to values of 0.58 and 0.70 kg COD/m^3.d for days 213 to 304 and days 381 to 409 respectively. There is no obvious explanation for this difference, but it may be due to sludge overflow from full compartments, or sludge wash-through by surge flow conditions that occurred from time to time.

- A more valuable observation for the 20 h T-HRT operating period is the *change in the slope* of the data presented in **Figure 4.12** between the time shortly after the increase in feeding rate (days 213 to 304) and those at the end of the operating period (days 381 to 409). These data show that the contribution to overall COD removal of all compartments after the first compartment increased during this operating period, although most removal still occurred in the first compartment. This corresponds well with the observations of increasing sludge bed height in these compartments (**Section 4.4.1**), which is both a cause and a result of the improved COD removal rate.

4.4.3 pH

pH measurements were performed on samples drawn from the ABR inflow, outflow, and the top sampling valve of each compartment. **Figure 4.13** presents the pH measurements of the ABR inflow and outflow in the three operating periods. No obvious trend with respect to time for the outflow pH values was observed.

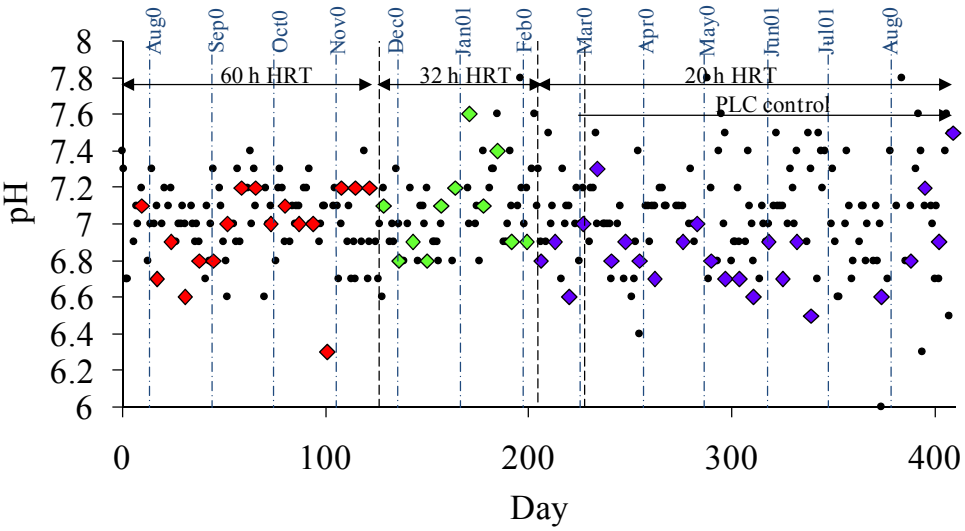


Figure 4.13: Phase I: pH measurements in inflow and outflow from the pilot-scale ABR. Points (●) indicate inflow values and diamonds (◊) are outflow values Data for 60 h (red), 32 h (green) and 20 h (blue) T-HRT (timer and PLC control) are shown.

Several anaerobic processes, but especially methanogenesis, are strongly inhibited by low pH values (below pH 6.5) (**Section** 2.1.5.2). Most pH measurements of inflow, outflow and the compartments during operation of the ABR at Umbilo WWTP have values in the range 6 to 7, and therefore, the rate of methanogenesis can be expected to be sensitive to variations in pH value. In the absence of significant H_2 gas production, methanogenesis is the step that actually removes COD from wastewater (**Section** 2.1.1.5); therefore, pH profile in the reactor can provide some clues to the overall status of anaerobic digestion. A conventional term describing extent of methanogenesis inhibition as a result of low pH values (Batstone et al., 2002) was calculated for each available pH measurement in each compartment and each day (Eq. 4-3). A value of 0 indicates complete inhibition, while 1 indicates no inhibition. These inhibition terms were averaged for each compartment, inflow and outflow for each operating period and are presented in **Figure 4.14**.

$$I = e^{-3\left(\frac{pH-pH_{LL}}{pH_{UL}-pH_{LL}}\right)^2}\Big|_{pH<pH_{UL}} \quad or \; 1\Big|_{pH\geq pH_{UL}}$$

Eq. 4-3

The minimum, maximum and median pH values for each compartment, inflow and outflow are presented for each operating period. Measurements were performed on samples that had been taken and transported to the laboratory. Exposure to air results in evolution of CO_2 due to the lower partial pressure of CO_2 in the atmosphere compared to the partial pressure in the reactor headspace. Evolution of CO_2 reduces the concentration of carbonic acid and results in an increase in pH value between

sampling and measurement (Sötemann et al., 2005). Thus pH values may have been lower in situ than reported here.

The compartment-by-compartment pH profiles indicate that acid-producing reactions dominated in the earlier compartments causing increase in acid concentration and lower pH values and that acid-removing processes recovered in later compartments. This is similar to other studies using compartmentalised reactors (see review in **Section** 2.5.1.4). Increase in pH value could be achieved by a number of routes: firstly acid produced by acidogenic reactions is consumed by methanogens, and hence an increase between earlier (acidogenesis dominated) and later compartments will be observed; Secondly, anaerobic digestion results in the production of alkalinity which increases pH; Finally, when little anaerobic digestion and therefore low gas production occurs in later compartments, gas exchange with the atmosphere through open gas vents results in low P_{CO2} values and consequently higher pH values that would be observed if the system were completely closed (Stumm and Morgan, 1996). In this case, the pH effect would not be due to phase separation, but rather, separation of head space.

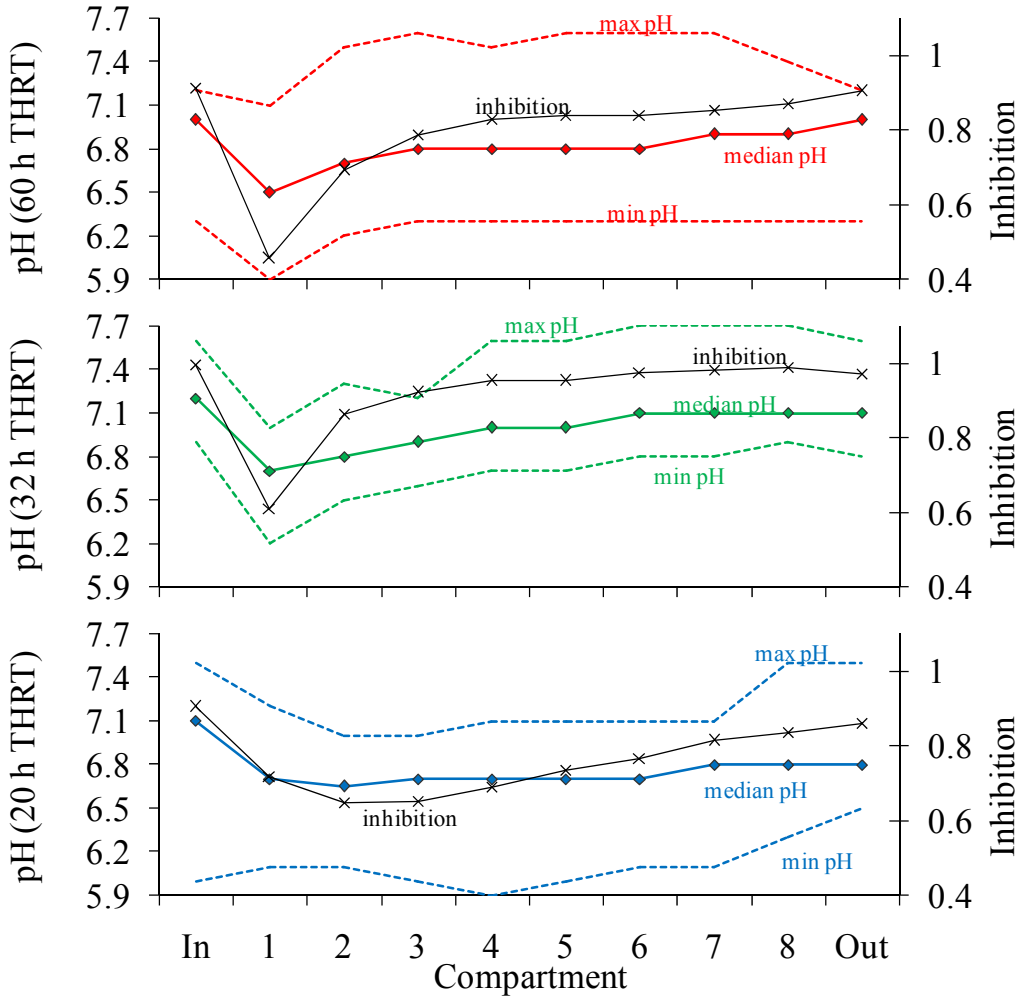


Figure 4.14: **Phase I: Minimum, maximum and median pH values measured in the inflow, outflow and at the top of each compartment of the ABR for 60 h (red), 32 h (green) and 20 h (blue) THRT periods respectively. Average inhibition factors (black) for acetoclastic methanogenesis calculated from Eq. 4-3 are presented for each operating period (1 = not inhibited, 0 = completely inhibited).**

The outflow pH value was on average slightly lower than corresponding inflow values. In dilute wastewater treatment, relatively low alkalinity generation potential and poor buffering is expected to

result in low pH values. The reasons for low reactor and outflow pH values are explored in **Section** 6.4, which investigates the range of pH values that can be expected in the outflow.

The 60 h and 32 h THRT periods show similar inhibition profiles, despite the changes in retention time. The big dip in the average inhibition term and corresponding pH value is due to most anaerobic activity occurring in compartment 1 and 2 with acid production and methanogenesis inhibition in compartment 1 and methanogenesis in compartment 2. Little digestion and therefore COD reduction of any kind occurs in the subsequent compartments due to low biomass concentration (**Section 4.4.1**). During the 60 h and 32 h T-HRT periods, there was little sludge and therefore little anaerobic activity in all but the first two compartments. Hence inhibition by acidogenesis products is small since the overall amount of acidogenesis is small.

Significant amounts of solids and therefore biomass, and associated activity only develop in compartments 4 onwards in the 20 h THRT operating period. The 20 h THRT period correspondingly shows a different trend to the two earlier periods; the methanogenesis inhibition term decreases in compartments 1 and 2 and begins to recover gradually over the subsequent compartments. This is attributed to growing amounts of sludge in all compartments resulting in acid production occurring in several compartments (not just compartment 1). Although average conditions for methanogenesis are worse than in the previous operating periods, the overall extent of treatment (fraction of biologically available COD removed) is greater since a greater amount of COD is removed.

4.4.4 Alkalinity

Bicarbonate alkalinity, measured in units of $mgCaCO_3/\ell$ was performed by acid titration using HCl on samples obtained from the inflow and outflow and from the top sampling valve of each compartment of the pilot-scale ABR.

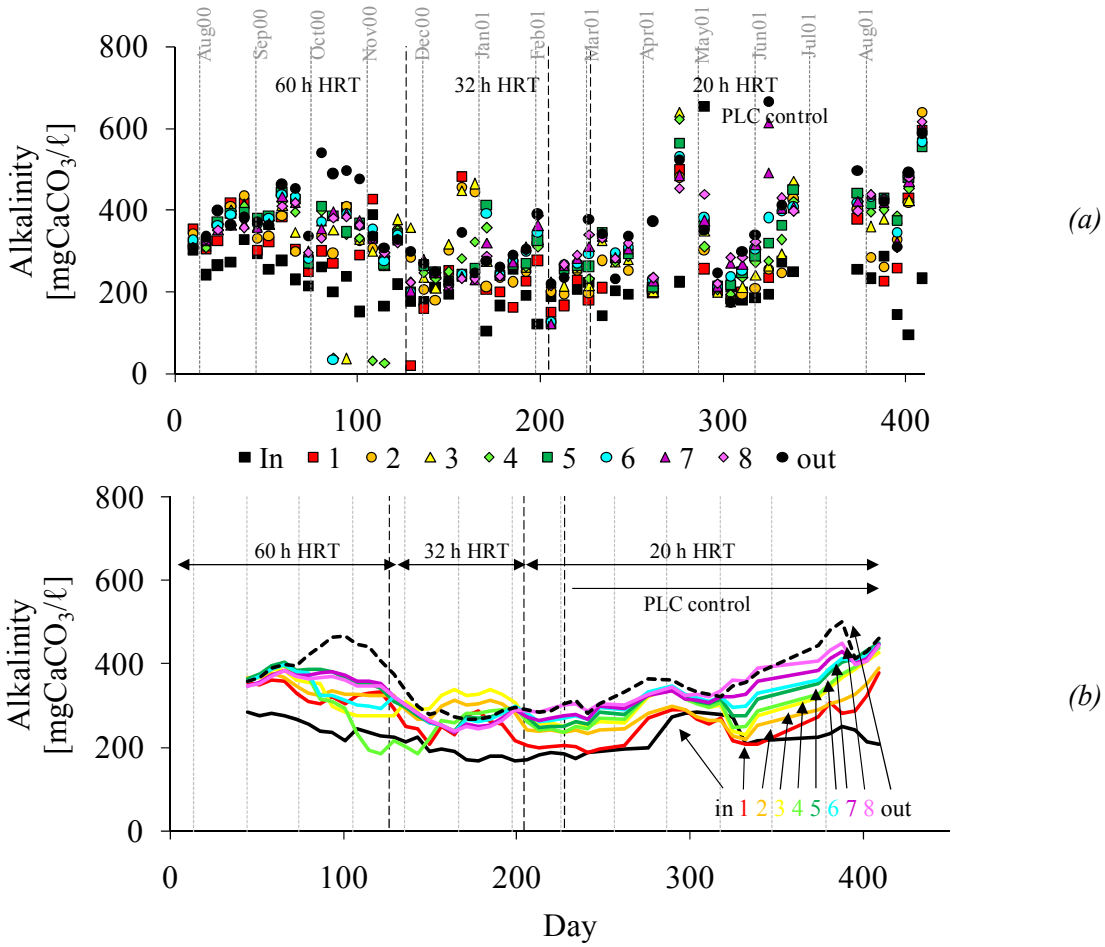


Figure 4.15: Phase I: Alkalinity measured in the inflow, outflow and at the top of each of compartments 1-8 shown as data points (a) and 6-period moving average trends (b).

Figure 4.15 (a) shows actual measured values of alkalinity in each compartment, giving an indication of the amount of scatter that was observed. **Figure 4.15 (b)** presents the same data as time-weighted averages so that overall trends may more easily be identified. From **Figure 4.15** it is possible to discern seasonal variation of inflow alkalinity with summer values (November to April) on average lower than winter values.

As expected, alkalinity concentrations in the inflow were low for a low strength anaerobic application due to the low alkalinity waters obtained in eThekwinin Municipality (**Section** 3.1.3).

In general, there is an increase in alkalinity value from one compartment to the next, with outflow values in each operating period significantly higher than inflow values.

Alkalinity shows a relatively constant increase between inflow and outflow in the first two operating periods, and for the first half of the third operating period. From day 300, the magnitude of the difference between inflow and ouflow alkalinity values increased with time, indicating that more alkalinity was being generated within the ABR. This inference is consistent with free and saline ammonia and total solids data that led to the understanding that increased sludge levels in each compartment resulted in a larger extent of treatment being obtained after day 300 of the first operating period, since alkalinity is generated during anaerobic digestion.

4.5 SUMMARY: PHASE I – OPERATION AT UMBILO WWTP 2000 -2001

The data presented in the previous sections were obtained from project records and reanalysed to assess whether any information could be gleaned on mechanisms of treatment during start-up of the ABR treating medium strength sewerage of mixed industrial/domestic origin. There was some uncertainty as to the validity of some of the data, particularly the outflow solids analysis due to reports that samples were taken immediately after the reactor was reset, and at which time washout of solids was regularly to be observed.

4.5.1 Summary of reactor performance: Phase I –Umbilo WWTP 2000 - 2001

Table 4.3 presents a summary of all inflow and outflow measurements averaged for operation of the pilot-scale ABR at Umbilo WWTP. During the three operating periods (60 h, 32 h and 20 h THRT), COD reduction of between 330 and 580 mgCOD/ℓ (48 to 81%) was achieved (calculated from the average inflow COD and high and low outflow COD values). Outflow COD concentrations were not substantially better than could be expected in the outflow of septic tanks.

Operating pH values below 7 indicate that the digestion in the pilot-scale ABR was relatively poorly buffered and that inhibition of methanogenesis occurred to a significant extent in early compartments. No measurements of VFA were made during this phase of experimentation. No analyses of the condition of the anaerobic sludge (i.e. in terms of microbial population analysis, or extent of granulation) were made.

4.5.2 Sludge bed accumulation during start-up

The results presented in **Section** 4.4 provide some information that may be used to understand the sequence of events during reactor start-up.

- The first compartment has an important role in retaining solids (**Figure 4.11**). There were no incidences of blockages of compartment 1 by accumulated solids; the solids were not found to have any negative impact on operation of the pilot-scale ABR.

- Only compartment 1 appears to build up any significant amount of sludge in the 60 h THRT operating period. Compartments 2 and 3 develop sludge beds that rise above the 200 mm sampling port in the 32 h THRT period, while all 5 remaining compartments achieve this only in the 20 h THRT period (**Figure 4.9**). These results suggest that solids do not easily move out of the first compartment under low flow conditions. Sludge beds in later compartments accumulated in quick succession at the higher flow rate and OLR.

- Build-up of solids after day 300 (2001) caused the change in free and saline ammonia and alkalinity measurements between compartments to increase (**Figure 4.7** and **Figure 4.15**). Free and saline ammonia and alkalinity are by-products of anaerobic digestion; therefore it was inferred that the extent of wastewater treatment achieved at this time was increasing. This corresponds to increasing sludge loads and implies an increasing extent of treatment.

- The average value for free and saline ammonia in the outflow stream is at times greater than the reported TKN value in the inflow for the 20 h HRT period. This implies that particulate solids that had previously accumulated were being digested and releasing free and saline ammonia to the liquid phase.

- Investigation into COD removal in each compartment shows that the majority of COD was removed in the first compartment (**Figure 4.12**). Much of the removed COD was retained particulate material that could not move beyond compartment 1 due to the baffle construction. The sludge bed created also acted as a filter that ensured good contact between organic material (particulate or soluble) and anaerobic micro-organisms. For very low loading rates, much of the reactor volume was not used for biological activity. This was as much a function of the lack of sludge present in these compartments as of the apparent efficiency of the first compartment. When loading rates increased, it was seen that the amount of treatment that occurred in later compartments increased, as did the sludge load. It was also seen that, although the contribution to overall COD removal of later compartments was low, they fulfilled an important buffering function to ensure that later compartments were not subject to low pH values caused by acidogenesis in the early compartments (**Section** 4.5.3 below).

As may be expected, the rate of accumulation of sludge in the ABR was directly related to the OLR. The reactor was tested up to hydraulic and organic loading rates approaching 20 h and 0.9 kg COD/m^3.d respectively, although precise figures cannot be calculated due to uncertainty in flow rate data. The loading rate was substantially less than the value of 1.5 kg $COD_{biodegradable}$/m^3/d proposed by Lettinga (2001) who indicated that 80 % COD removal could be achieved at a HRT as low as 4 h for anaerobic systems for the pre-treatment of domestic wastewater in tropical climates.

4.5.3 Phase-separation between compartments

A number of authors observed a phase-separation phenomenon during the operation of compartmentalised anaerobic digesters (**Sections** 2.5.1.4 and 2.2.3). The following observations were made in this study:

- In situ pH values may have been lower than those measured because of the lag between sampling and analysis (Section 4.4.3)

- pH measurements made on samples taken from within compartments show that significant acidification occurred in the first compartment at longer retention times. It was inferred that methanogenesis did not proceed at the same rate as acidogenesis, resulting in the accumulation of VFA and lowering of the pH value. However subsequent analysis indicated that low pH values may not necessarily have been due to methanogenesis inhibition (**Section** 6.4)

- Conventional inhibition terms were used to estimate the extent of inhibition of methanogenesis that may have occurred as a result of low pH values. Significant inhibition of methanogenesis is expected in compartment 1, although the extent of inhibition decreases in subsequent compartments.

- pH profiles for the 60 h and 32 h THRT periods were similar, with pH depression observed in the first compartment. However, in the 20 h T-HRT, low pH values were observed in the first three compartments. The difference was attributed to different sludge load profiles in the different operating periods. The pH value at the reactor outflow during the 20 h THRT period was similar to those observed in the earlier periods.

These results point to a very useful attribute of the ABR in the treatment of domestic wastewater. The compartmentalisation of the ABR provides a degree of phase separation such that hydrolysis and acidification of organic material in the early compartments can proceed at pH values that are lower

than the optimal range for methanogenesis, but that pH values remain near-neutral in later compartments ensuring that methanogenesis can occur. Having more than two compartments means that it is not necessary to optimise the volumes allocated to acidification and methanogenesis since predominance of acidogenesis will extend to subsequent compartments if the OLR is sufficiently high, while compartments are still available further down the reactor to allow methanogenesis to occur uninhibited.

It is understood that digestion of sewage in the ABR is poorly buffered due to the low alkalinity producing potential of the dilute wastewater feed, and the fact that no buffering agents were added during the process. However, the compartmentalised design ensured that digestion did occur with good overall process stability despite this fact.

Table 4.3: Phase I: Summary of inflow and outflow stream characteristics. Data are presented as mean value ± 95 % confidence interval on the mean [min, max] (number of observations)

		In[1]	Out (60 h T-HRT)	Out (32 h T-HRT)	Out (20 h T-HRT)
COD	mgCOD/ℓ	712 ± 29 [151, 1 845] (265)	379 ± 61 [166, 612] (16)	170 ± 54 [55, 255] (8)	272 ± 40 [137-564] (24)
Alkalinity	mg$CaCO_3$/ℓ	215 ± 2 [66, 424] (271)	396 ± 35 [303, 540] (17)	286 ± 28 [225, 387] (11)	371 ± 57 [172, 666] (20)
Free and saline ammonnia	mgN/ℓ	23 ± 0.6 [3, 40] (271)	33 ± 9 [6, 60] (16)	33 ± 6 [22, 56] (11)	44 ± 8 [14, 106] (21)
Total phosphate	mgP/ℓ	6.3 ± 0.6 [1.1, 18.0] (96)	n.d.	n.d.	n.d.
Soluble phosphate	mgP/ℓ	n.d.	2.4 ± 1.3 [0, 6.8] (16)	1.1 ± 0.9 [0.1, 5.1] (10)	7.0 ± 1.8 [0.9, 16.4] (23)
Total solids	mgTS/ℓ	1 256 ± 295 [505, 8 645] (52)	2 177 ± 927 [405, 8 453] (16)	1 080 ± 359 [387, 2 191] (10)	13 782 ± 6 529 [22, 55 874] (24)
Volatile solids	mgVS/ℓ	655 ± 249 [145, 6 657] (49)	1 210 ± 628 [231, 5 579] (16)	371 ± 158 [90, 789] (10)	10 297 ± 1 579 [10, 37 436] (19)
pH	(median value reported)	7.0 [6.0, 9.2] (272)	7.0 [6.3, 7.2] (17)	7.1 [6.8, 7.6] (11)	6.8 [6.5, 7.5] (24)
TKN	mgN/ℓ	42 ± 4 [21, 68] (21)	n.d.	n.d.	n.d.

[1] Overall average conditions are reported for the inlet stream (i.e. all data from Winter and Summer are considered in the reported descriptive statistics.

5 EXPERIMENTAL PHASES II-IV: KINGSBURGH WWTP 2002-2004

The pilot-scale ABR was moved to Kingsburgh WWTP in January 2002. Operation at Kingsburgh WWTP formed the basis of an MScEng Thesis (Mtembu, 2005). Kingsburgh WWTP is situated approximately 30 km south of Durban and 30 minutes drive from UKZN. This works treats a wastewater that has no formal industrial effluent component. It serves a community of about 350 000 population equivalents from middle-income suburbs. The pilot-scale ABR was moved from Umbilo WWTP to Kingsburgh as it was believed that a better understanding of the functioning of the ABR in sanitation would be obtained without complications from trade effluent.

A disadvantage of Kingsburgh WWTP compared to Umbilo WWTP was that there was no analytical laboratory on site. Plant workers could be persuaded to record flow rate data and to report if the reactor did not seem to be working, but were otherwise not involved in the operation of the plant. Composite samples of inflow wastewater were withdrawn from the head of works by WWTP staff, and the results of routine analyses on these samples were made available to the project team by the municipality. Sampling was performed by project team members and samples had to be transported back to the university laboratory or to the central municipal laboratories by project team members with some unavoidable sample deterioration. While improved control and documentation of plant operation was achieved at Kingsburgh WWTP, there was an inevitable decrease in the abundance and (in some analytes), quality of data that were obtained during this time.

5.1 KINGSBURGH WWTP INSTALLATION

Prior to removal from Umbilo WWTP, the ABR was allowed to stand for a week to allow sludge to settle. The liquid fraction in each compartment above the bottom sample/drain valve was drained away, leaving 200 mm of sludge in the bottom of each compartment. This sludge was available as seed sludge for start-up at Kingsburgh WWTP.

The pilot-scale ABR was installed near the head of works at Kingsburgh WWTP. The feed pump was lowered into a sump (**Figure 5.1** (right)) that sent screened and degritted wastewater to the activated sludge plant. Incidences of rags blocking the pump were substantially less during operation at Kingsburgh WWTP compared to those experienced at Umbilo WWTP since the Kingsburgh wastewater contained no textile industry effluent. However, string, rubber and hair regularly found their way into the pump, causing interruptions to pumping and damage to the pump motor. Several means of eliminating these from the pump were attempted, but by far the most successful measure was installing the pump in a laundry basket (**Figure 5.1**). Instances of pump blockages were reduced by more than half as a result of the laundry basket system (Mtembu, 2005). During the time the ABR was operated at Kingsburgh, at least 5 laundry baskets were used in this way!

In the first year of operation at Kingsburgh WWTP (Phase II), there were a number of occasions when the outlet pipe from the ABR blocked. Under normal operation, a siphon breaker on the outflow line controlled the height of liquid in the reactor. However, when the outlet pipe became blocked, the reactor would fill up to as much as 100 mm above the normal operating level, causing the accumulation of up to 300 ℓ of excess liquid in the reactor.

Figure 5.1 **Installation of the ABR at Kingsburgh WWTP. The outlet end of the pilot-scale ABR (left); and a laundry basket housing the submersible pump in a wastewater sump near the feed end of the ABR (right)**

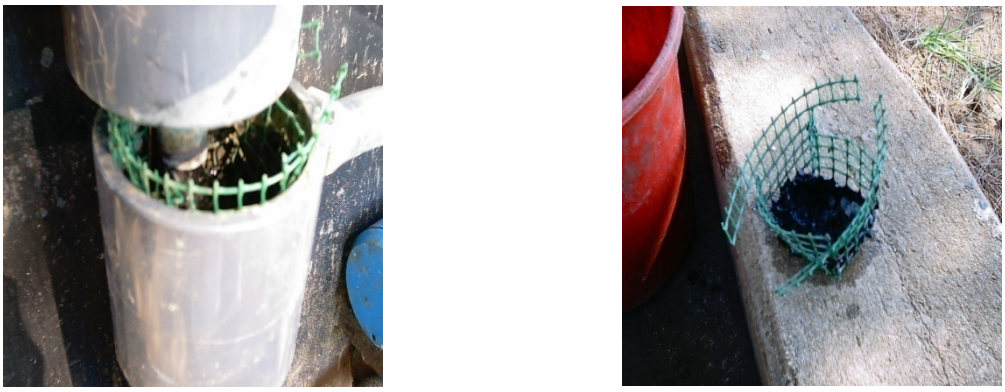


Figure 5.2: **Modified outlet of the ABR showing mesh for preventing coarse solids entering the flow meter.**

When the blockages were removed, large amounts of backed-up liquid would flow out achieving flow rates (and therefore internal upflow velocities) up to 10 times greater than during normal operation, and resulting in large amounts of sludge being washed out. In this way, considerable amounts of sludge were lost during operation in Phase II. The cause of the blockages was found to be tiny cones (about 15 mm diameter) from conifers growing next to the installation (just visible in the top right of **Figure 5.1** (left)). These would block the outlet pipe just before the flow-meter. Three changes were instituted part-way into Phase III to eliminate this problem:

- A lid was built for the feed splitter box to prevent ingress of the cones
- The outlet pipe was split on a vertical (downflow) section, upstream of the flow meter, and a large mesh screen was installed in the split to intercept large objects with diameter >7 mm (**Figure 5.2**).
- The PLC programme was adjusted to switch off power to the pump if no flow was obtained at the flow-meter.

These steps both reduced incidents of blockages and prevented overfilling of the reactor when any malfunction caused blockages within the reactor or at the outflow. One other problem of this nature was observed: gas vent valves were accidentally left closed, or became blocked with accumulated

sludge on a number of occasions. When this occurred, gas production would result in the accumulation of significant gas pressure in the headspace of some of the early compartments (usually compartment 2). Eventually the accumulated gas would depress the level in the upflow side of the compartment to the extent that overflow to the downflow side could not occur, despite the liquid head exerted from the previous, overfilled and pressurised compartment. Thus flow through the reactor stopped. This caused considerable confusion since no blockages could be found in the path of the liquid flow. However, the problems were usually resolved by clearing the gas vent valves with steel wire kept handy for the purpose.

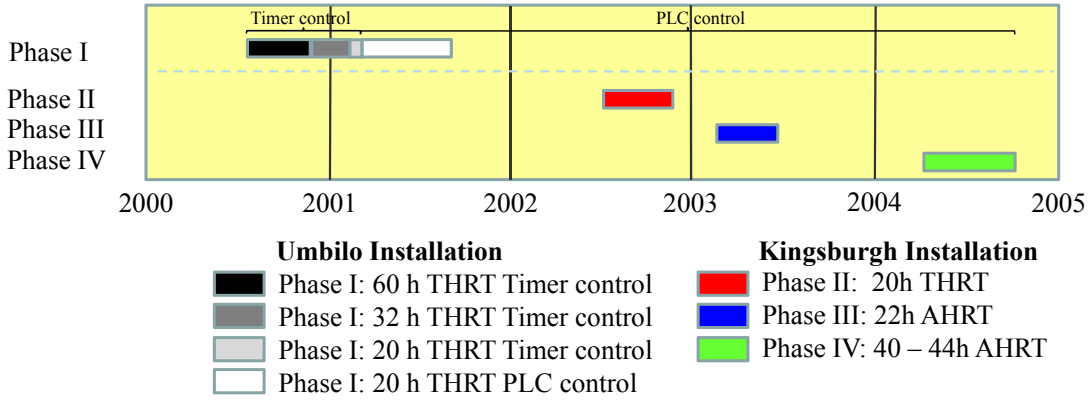


Figure 5.3: Pilot-scale ABR project time line showing different phases of operation included in this study at Umbilo and Kingsburgh WWTP. THRT: Target hydraulic retention time; AHRT: Average calculated hydraulic retention time.

5.2 OPERATION OF THE PILOT-SCALE ABR: KINGSBURGH WWTP INSTALLATIONS

Three periods of operation at Kingsburgh WWTP yielded data that are presented in this thesis.

5.2.1 Operation of the pilot-scale ABR: Phase II

The pilot-scale ABR was moved from Umbilo WWTP to Kingsburgh WWTP at the beginning of 2002. Phase II was from 2 July 2002 to 20 November 2002.

During Phase II, the A-HRT was set to 20 h, using a proportional-integral (PI) control algorithm implemented in the PLC. Figure 5.5 provides an indication of the nature of the flow rate under PI control. It can be seen from the upper figure that the flow rate is relatively constant, and from the lower figure, that the error between recorded flow rate and set-point $\left(error = \frac{flow\ rate}{setpoint}\%\right)$ was usually less than 10% of the specified flow This may be compared to flow rate characteristics under bang-bang control (Phase I and II of operation, Figure 4.4); a significant reduction in the amplitude of the oscillations in flow rate may be seen from >100% error under bang-bang control within an oscillation to less than 10% error under PI control.

The reduced amplitude of oscillation is particularly significant since the variation in effluent flow rate is directly related to variations in upflow velocity. The flow rate data presented in Figure 5.5 imply that, as no large oscillations in effluent flow rate were observed, there were no large oscillations in up-flow velocity in compartments due to the time-slicing algorithm on the feed flow rate controller.

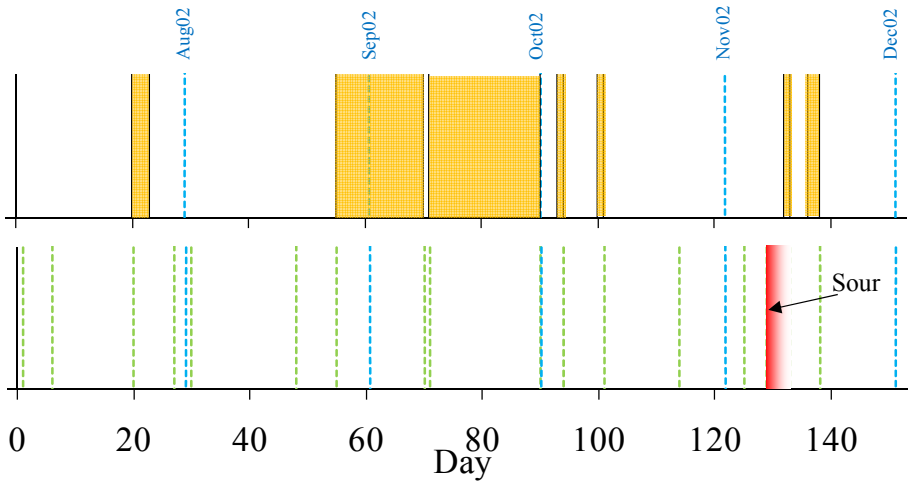


Figure 5.4: **Phase II: Incidents and down time. Orange shading indicates reactor down time. Green lines indicate potentially performance affecting incidents such as sludge washout. A "souring" incident on day 129 is indicated on the lower part of the figure.**

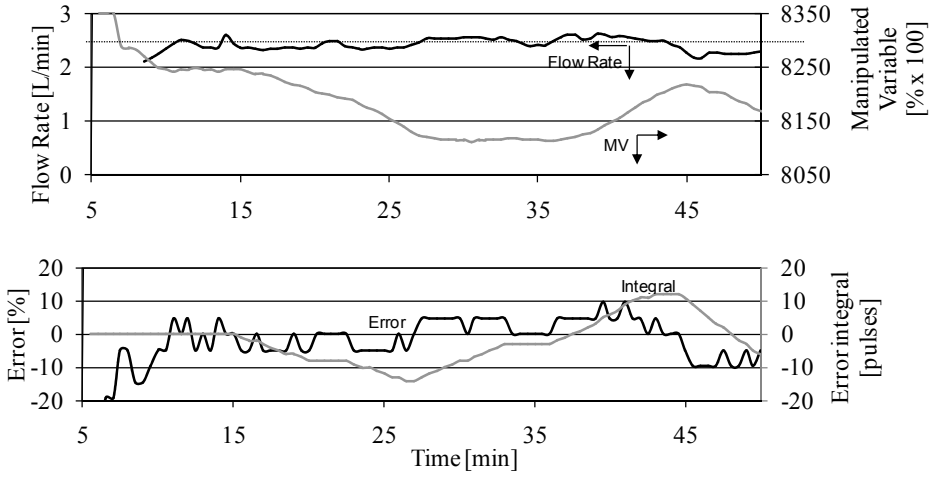


Figure 5.5: **Phase II: Flow characteristics under PI control. The manipulated variable (MV) is the fraction of each 90s cycle that the bypass valve is open and is shown with the resulting flow rate in the upper figure. The difference between the measured flow rate and flow setpoint, and the error integral are shown in the lower figure.**

A total of 350 kℓ of wastewater was treated at an average of 2 800 ℓ/d in 139 days. Incidents that caused high flow with sludge loss or down time were due to electrical and mechanical problems with the pump, compressor and pneumatic valve. A graph of down-time and performance-affecting incidents in Phase II is presented in **Figure 5.4**. A souring incident occurred on day 129. The souring incident is discussed in detail in **Section** 5.6.3.

5.2.2 Operation of the pilot-scale ABR: Phase III

In the second operating period at Kingsburgh WWTP, Phase III, the pilot-scale ABR was operated with a T-HRT of between 20 and 24 h, with an average of 22 h being achieved. A total flow of 353 kℓ of Kingsburgh WWTP wastewater was treated in 126 days. Considerably less down time, or performance-affecting incidents occurred during this operating period as a result of improvements to the control algorithm. Incidents, down time and volume of wastewater treated in Phase III are presented in **Figure 5.6**.

On days 99 and 100 of Phase III, a 24 h sampling campaign was undertaken in which inflow and outflow were sampled hourly for 24 h (outflow samples were taken approximately one A-HRT i.e. 20 h after inflow samples) to assess the effect of diurnal variations in wastewater strength and composition.

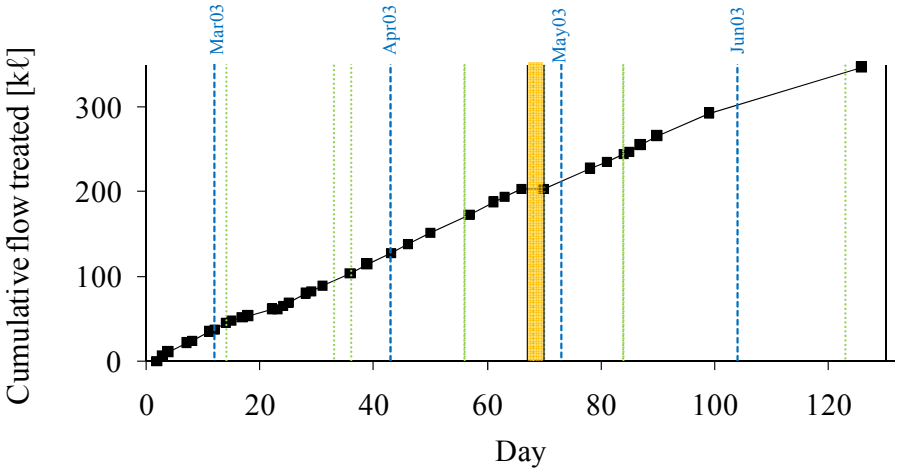


Figure 5.6: Phase III: Cumulative flow treated, incidents and down time. Orange shading indicates reactor down time. Green lines indicate potentially performance affecting incidents such as sludge washout.

5.2.3 Operation of the pilot-scale ABR: Phase IV

A third period of operation at Kingsburgh, Phase IV, took place between 7 April 2004 and 8 October 2004, with an A-HRT of between 40 and 44 h. In this time, 293 kℓ of Kingsburgh WWTP wastewater was treated (**Figure 5.7**).

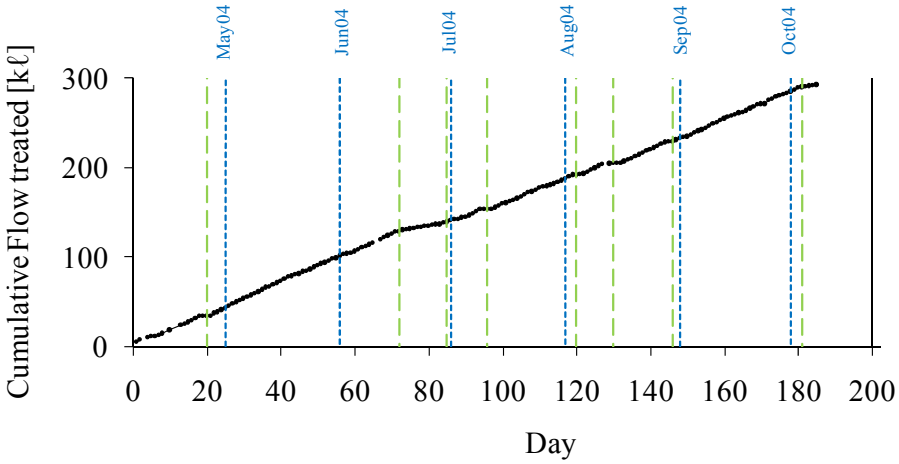


Figure 5.7: Phase IV: Cumulative flow treated and incidents. Green lines indicate potentially performance affecting incidents such as low feeding rate.

There were no significant periods of down time during Phase IV. **Figure 5.7** shows a number of *performance-affecting* incidents. However, these were all low-flow incidents related to problems with the feed system. Fine tuning of the feeding and control systems as well as improvements in the general maintenance plan meant that there were no incidents in which significant amounts of sludge washout or wash through of the reactor occurred at any time during this phase.

5.3 FEED CHARACTERISTICS AND LOADING RATES: PHASE II-IV

The wastewater that enters Kingsburgh WWTP is drawn from a mixed-income, (but predominantly middle-income) community with no heavy industrial component. Although some small businesses such as laundrettes and restaurants discharge more highly loaded wastewaters than households into the sewers that feed Kingsburgh WWTP, the sewerage is considered to be fairly representative of a purely domestic wastewater.

5.3.1 Characterisation of Kingsburgh WWTP wastewater

The feed characteristics of wastewater entering Kingsburgh WWTP and the pilot-scale ABR are presented in **Table 5.1**. These data are a combination of data generated by the municipality for head-of-works wastewater composition, and data from samples taken from the feed box to the pilot-scale ABR drawn and analysed by the project team. Where appropriate, statistical analyses were performed to determine whether there was any significant difference between municipal data and project team data, and these showed no significant difference (95% confidence level).

Table 5.1 Characteristics of degritted wastewater fed to ABR at Kingsburgh WWTP head-of-works.

	Unit	Mean (± 95% conf. interval)	Min.	Max.	Number of observations (n)
COD	mgCOD/ℓ	680[1] (± 25)	246	1 749	258
Soluble COD	mgCOD/ℓ	154 (± 40)	69	395	20
Biodegradability	mgCOD / mgCOD %	> 47[2]			
Alkalinity	$mgCaCO_3/ℓ$	243[1] (± 6)	43	369	269
pH	-	7.1[3]	4.0	7.94	289
Free and saline ammonia	mgN/ℓ	39 (± 1.8)	3.8	204	273
PO4	mgP/ℓ	15 (± 0.8)	1	64	245
Total solids	mgTS/ℓ	673 (± 66)	253	1 076	43
Volatile solids	mgVS/ℓ	417 (± 66)	125	705	25
Suspended solids	mgSS/ℓ	331 (± 13)	94	771	224

[1] Data from Phase II was not used for calculating average and range of COD and alkalinity data, since Phase II data was shown to be significantly different from Phase III and Phase IV data for these determinands.

[2] A measurement of inflow stream biodegradability was determined using a serum bottle test by Mtembu (2005). However, the test was not allowed to run to completion, thus the reported biodegradability of 47% should be regarded as a lower limit, with the expected value being somewhat higher.

[3] The median recorded value for pH is reported.

5.3.2 Analysis of variations in feed wastewater between phases of operation

The feed characterisation data from Phases II, III and IV were analysed to determine whether there were significant variations in feed composition between the different operating periods:

- It was found that the mean inflow COD concentration was significantly *higher* in Phase II than in Phase III (Students T-test, unequal variances, $P = 1.4 \times 10^{-4}$) and Phase IV ($P = 2.0 \times 10^{-4}$), and that there was no significant difference between mean COD concentration measured in Phase III and IV. These differences were observed independently in both project team measured data and municipal data. It would therefore appear that there was some change at a community level that resulted in significant reduction in inflow COD between Phase II and Phase III. There was no significant difference in wastewater strength in terms of COD between wastewater from Umbilo WWTP and the average Kingsburgh inflow ($P = 0.78$).

- There were no significant differences in free and saline ammonia concentration between operating phases at Kingsburgh WWTP. However, the concentration of free and saline ammonia in Kingsburgh WWTP wastewater was significantly higher than in Umbilo WWTP wastewater (Student's t-test, $P = 3 \times 10^{-32}$)

- Alkalinity concentrations measured in the Kingsburgh WWTP wastewater during Phase II were significantly lower than for any of the other phases of operation at Kingsburgh or Umbilo WWTP. (Students T-test, unequal variances, $P << 0.05$ for all combinations). There were significant increases in inflow alkalinity between all of the different phases of operation at Kingsburgh WWTP; i.e. the load of alkalinity in the incoming wastewater showed significant increases with time. The reason for these changes is not known; they could be related to changes in the population served by the wastewater treatment plant. It was noted that there was no change in the median pH value measured in the different operating periods. Higher values of inflow alkalinity were measured at Kingsburgh WWTP than at Umbilo WWTP (Students T-test, unequal variances, $P = 6 \times 10^{-4}$).

- Lower values of total and volatile solids were measured in Kingsburgh WWTP inflow compared to Umbilo WWTP inflow, although the difference in volatile solids measurement was not highly significant (Students T-test, unequal variances, $P = 4 \times 10^{-4}$ (TS) and $P = 0.07$ (VS)).

- Suspended solids measurements were performed only by municipal staff on samples from head of works. There were highly significant differences between measurements from Umbilo WWTP and Kingsburgh WWTP ($P < 1 \times 10^{-100}$) with values from Kingsburgh WWTP wastewater being more than an order of magnitude higher than those from Umbilo WWTP. Although the Kingsburgh wastewater was expected to have higher suspended solids content due to the greater fraction of domestic wastewater in the sewerage, the magnitude of the difference suggests that different sampling or analyses techniques may have been used that would have accounted for the difference:

  Although suspended solids data from Umbilo and Kingsburgh WWTP were both sampled and analysed by Municipal staff, the people and the laboratories were different. Unlike COD and total solids measurements, which are understood to apply to the whole sample and are measured using standard techniques, settleable or suspended solids required a separation step

that may be undertaken in a number of different ways (e.g. through flocculation and settling, free settling, filtration or centrifugation) resulting in widely varying results.

There were also significant differences between suspended solids concentrations in the different phases of operation at Kingsburgh ($P < 0.02$ for all combinations) with the highest suspended solids values associated with Phase II. These results support the observation that higher COD values were obtained in Phase II and indicate that the wastewater was generally of a higher strength during this period.

All of the differences between Umbilo and Kingsburgh wastewaters can be explained by the different sources of the wastewater: it is expected that the textile effluent component of the Umbilo wastewater would have low alkalinity and low free and saline ammonia / organically bound nitrogen compared to domestic wastewater, but could have high total solids due to the use of salts for dye fixing and the presence of textile fibres in the effluent.

These data indicate that the feed characteristics may have been fundamentally different between Phase I (Umbilo WWTP) and Phases II-IV (Kingsburgh WWTP), and that this may have affected the sludge dynamics and outflow stream characteristics of the reactor at the different locations.

5.3.3 Hydraulic and organic loading rates: Phase II – IV

During Phase II and Phase III, the intention was to reproduce the hydraulic loading conditions that were obtained in Phase I. Thus the A-HRT was between 20 and 24 h.

In Phase II, flow rate data were not recorded regularly. However, the average flow rate was inferred from the total flow treated (350 kℓ) and the number of days of operation (excluding down-time: 99 days):

- Towards the end of operation of the pilot-scale ABR at Umbilo WWTP, the previous project manager had caused the siphon breaker on the outlet line to be raised to the height of the first baffle. This increased the internal volume of the ABR to (1 m × 3 m × 1.14 m =) 3.420 m^3
- The average flow rate through the reactor was 350 000 ℓ / 99 d = 3 500 ℓ/d
- The average A-HRT was thus 3 420 ℓ / 3 500 ℓ/d × 24 h/d = 23 h

In Phase III the pilot-scale ABR was operated with a T-HRT of between 20 and 24 h, with an average of 22 h being achieved.

In Phase IV the loading rate was reduced to counteract some perceived instabilities in the system. During this phase, three continuous operating periods were achieved with mean A-HRT of 41 h, 44 h and 42 h.

The average OLRs exerted in each of the phases were calculated using average A-HRT and inflow COD data and are presented in **Table 5.2**. The values of OLR presented in **Table 5.2** are generally lower than loading rates used in the treatment of domestic wastewater at ambient temperature reported in literature (**Table 2.5**).

Table 5.2: Pilot-scale ABR average organic loading rate [OLR] [kg COD/m^3.d] and average upflow velocity in upflow compartments during stable operation at Kingsburgh WWTP (i.e. excluding high and low flow incidents)

	Ave. HRT	Ave. inflow COD	Ave. OLR	Ave. Upflow velocity
	h	mgCOD/ℓ	kg COD/m^3.d	m/h
II	23	918	0.95	0.52
III	22	676	0.74	0.55
IV	41	753	0.44	0.30
IV	44	664	0.36	0.27
IV	42	689	0.39	0.28

5.4 ABR OUTFLOW STREAM CHARACTERISTICS: PHASE II - IV

The overall performance of the pilot-scale ABR under the conditions tested was assessed by examination of the outflow characteristics of the reactor in terms of COD, nutrients and solids concentration.

5.4.1 COD

Figure 5.8 shows the inflow and outflow COD concentrations measured during operation of the pilot-scale ABR at Kingsburgh WWTP during the Phase II, II and IV. Red squares indicate weekly measurements of inflow COD made by the project team at the inlet to the pilot-scale ABR, while green circles (Phase III and IV) are similar measurements performed on a daily basis by the municipal staff at the head of works. ABR outflow samples were in some cases filtered through 0.45 µm acetate filters and analysed for COD as an indication of the amount of soluble COD present in the outflow.

5.4.1.1 Phase II

In Phase II, outflow COD values were around 300 mg/ℓ, with 20% to 50% of the measured COD being associated with suspended solids. The ABR consistently removed 500 to 600 mgCOD/ℓ except during the souring incident. The lowest outflow COD measurement was 125 mgCOD/ℓ. (A value of 64 mgCOD/ℓ measured on day 90 was ignored since this value was recorded on the first day after a lengthy period in which the reactor did not function at all, following electrical damage to the feeding system.)

The souring incident on day 129 (**Figure 5.8**, Phase II) resulted in a large spike in the outlet COD concentration to 2 630 mgCOD/ℓ. Souring is a result of acidogenesis rate exceeding the rate of methanogenesis; volatile fatty acids accumulate, thereby lowering the pH and inhibiting methanogenesis. If no methanogenesis occurs, no COD removal will occur, and outflow COD values will be high. It was observed that when sour, anaerobic sludge showed poor settling characteristics relative to normal operation, and consequently increased sludge washout was inferred from the black and turbid appearance of the outflow in this incident, further increasing the outflow COD value. Thus high outflow COD values were due to a combination of high VFA concentration and washed-out anaerobic sludge.

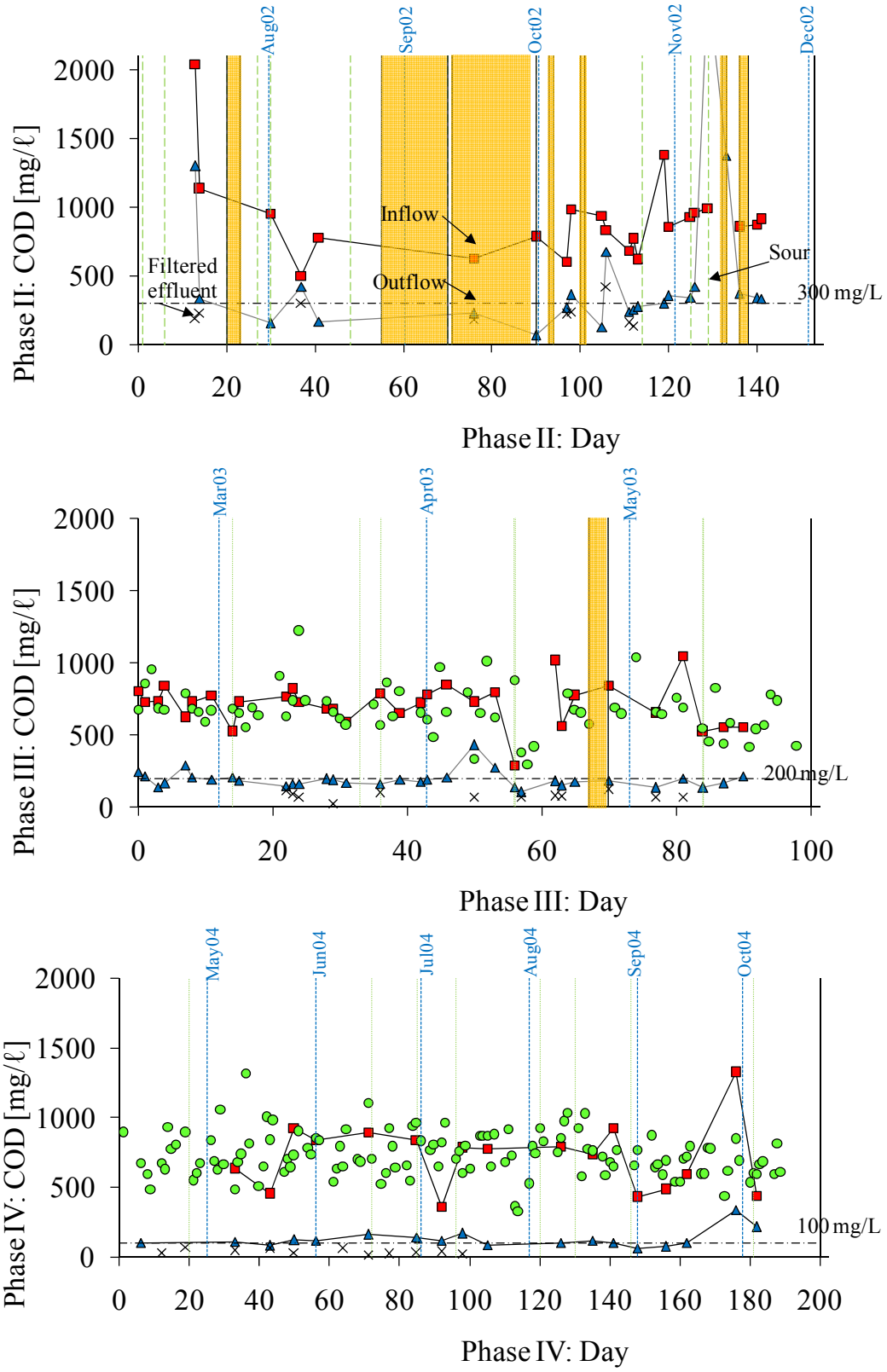


Figure 5.8: **COD concentrations in inflow and outflow samples (Phase II, III and IV). Project team-measured inflow (red), municipality-measured inflow (green), outflow (blue) and 0.45 µm filtered outflow (×) measurements are shown. The black dash-dot (-·-·) line indicates a COD value of 300 mg/ℓ in Phase II, 200 mg/ℓ in Phase III and 100 mg/ℓ in Phase IV.**

5.4.1.2 Phase III

In Phase III, (**Figure 5.8**), although the operating A-HRT were similar during this period to the Phase II, the COD removing performance of the reactor was significantly better. This was attributed to more stable flow conditions, and fewer sludge carry-over incidents than were observed in Phase II. A more concentrated biomass, better suited to compartment conditions was able to develop. This is corroborated by the higher sludge levels seen in Phase III as compared to Phase II (**Section 5.6.1**).

5.4.1.3 Phase IV

In Phase IV, even lower outflow COD values were obtained, with a mean outflow COD value of 130 mg/ℓ, and with values regularly dipping below 100 mg/ℓ (**Figure 5.8**, Phase IV). It was hypothesised that the longer retention time of Phase IV (42 h) as compared to the Phase III (22 h) resulted in a greater *extent of removal* (fraction of biodegradable COD converted to CH_4). Mass balance calculations were undertaken to test this hypothesis (**Section** 6.2). However, the amount or load of COD converted was less than in previous periods, because of the lower OLR.

5.4.2 Total and volatile solids

Figure 5.9 shows total and volatile solids measurements from inflow and outflow streams for Phase II and Phase III, and total solids data only for Phase IV. Suspended solids data (solids retained after settling, filtration, or centrifugation) measured by municipal staff are also presented. Municipal data are always lower than the equivalent values of total solids measured by the project team as the latter measurement includes dissolved solids. In Phase II, average total solids and volatile solids removals of 44% and 52% respectively were obtained. These increased to 53% and 59% respectively in Phase III.

The general improvement in removal rates between Phase II and Phase III was attributed to better flow control resulting in fewer washout incidents and development of a more stable and concentrated sludge bed.

No volatile solids measurements were made in Phase IV. An overall total solids removal rate of 48% was calculated for Phase IV. The average concentration of total solids in the outflow was also significantly higher in Phase IV than in Phase III (Student's t-test, $P = 0.0002$), despite the lower hydraulic loading rate. The cause of this difference is not known.

5.4.3 Alkalinity

Figure 5.10 shows inflow and outflow alkalinity concentrations for Phases II and III. Only a few data points are available for Phase IV; averages for Phase IV are reported in **Table 5.6**.

In both Phase II and Phase III outflow alkalinity values were above inflow alkalinity values indicating that alkalinity is generated by the partial conversion of COD to bicarbonate and concomitant release of cations (**Section** 2.1.6.3).

Speece (1996) recommends that alkalinity concentration in the operation of anaerobic digesters is maintained at sufficiently high concentrations to provide a *reserve alkalinity* that is available to neutralise additional acids produced by fermentation. The reserve alkalinity ensures that the operating pH value does not drop below a range of between 6.2 and 6.5 since metabolic rates may be adversely affected below these values. It is shown in **Section** 5.6.2 that the pilot-scale ABR treating a relatively dilute (700 mgCOD/ℓ) wastewater was operating with very little reserve alkalinity since pH values are lower than these recommended targets. It is therefore probable that poor buffering as a result of low

alkalinity, and consequently low pH values, caused non-optimal conditions for microbial activity, resulting in relatively low treatment rates being achieved.

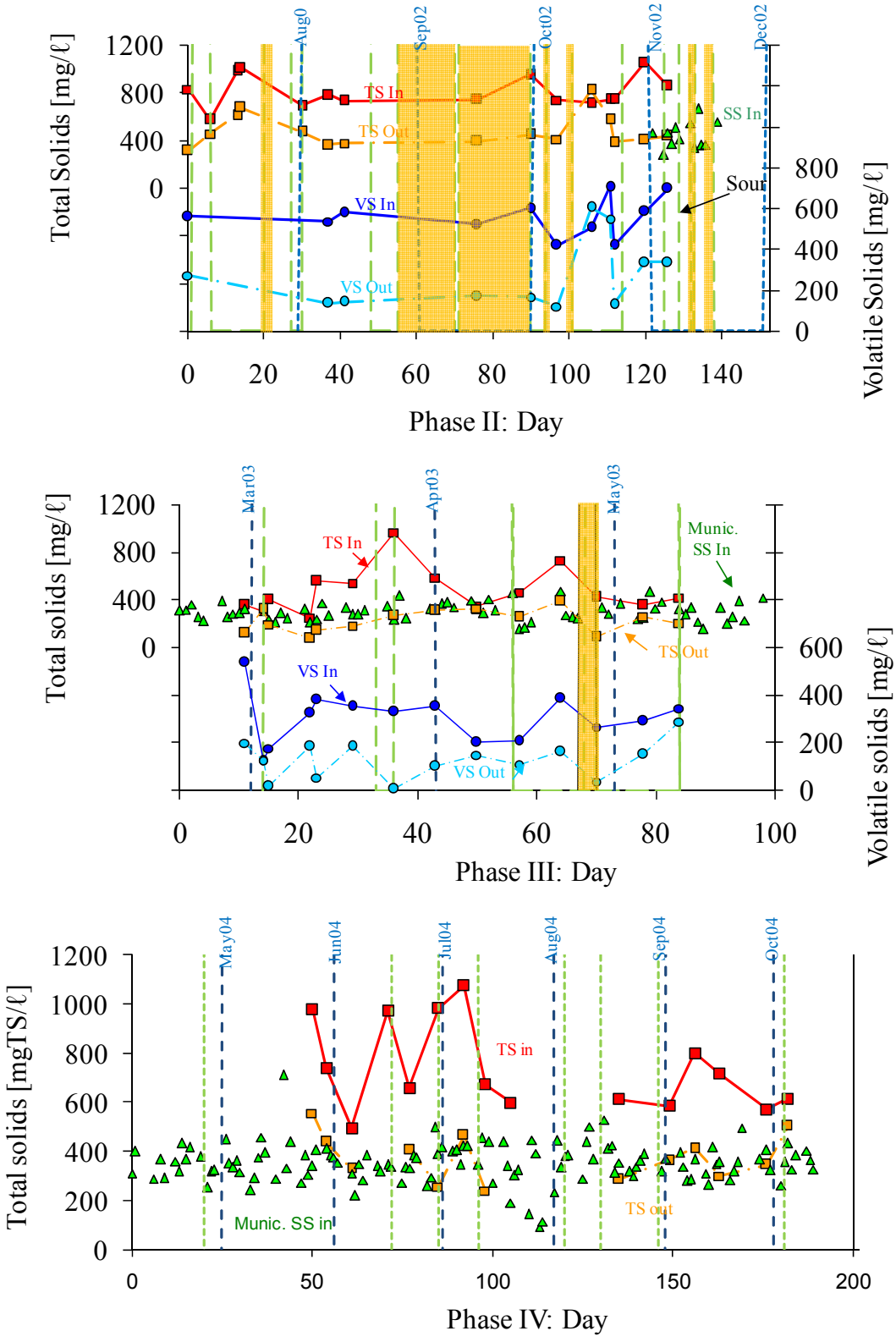


Figure 5.9: **Total solids (TS) and volatile solids (VS) concentrations in inflow and outflow samples (Phase II, III and IV). Project team-measured TS inflow (red) and TS outflow (orange), VS inflow (blue) VS outflow (turquoise) and municipality-measured inflow suspended solids (green) are shown.**

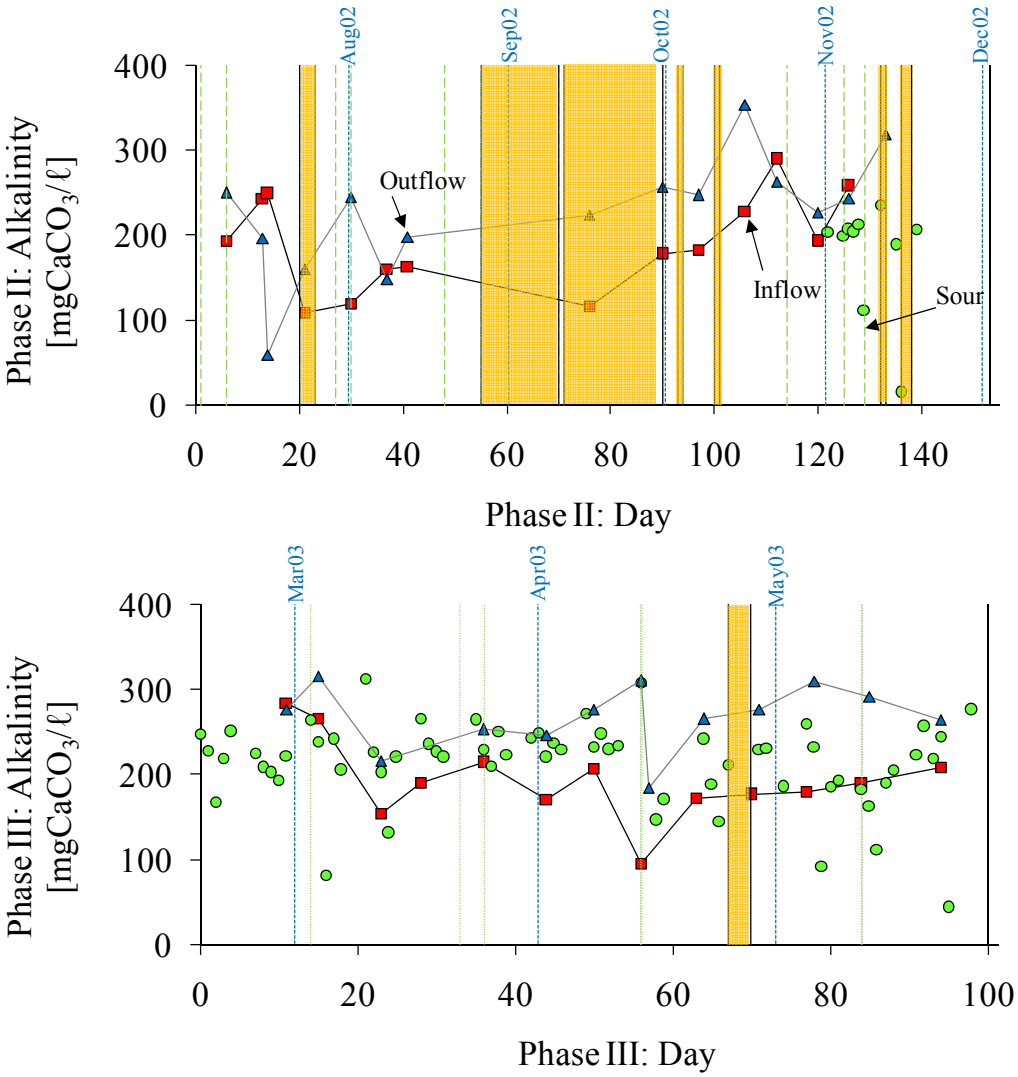


Figure 5.10: Alkalinity concentrations in inflow and outflow samples (Phase II, III) measured by titration to pH 4.5 endpoint. Project team-measured inflow (red), municipality-measured inflow (green) and outflow (blue) measurements are shown.

5.4.4 Free and saline ammonia and Total Kjeldahl Nitrogen

Anaerobic digestion liberates organically bound nitrogen in the feed material as free and saline ammonia, resulting in a net increase in free and saline ammonia concentration. No *retention* of free and saline ammonia is expected, as would be the case for a determinand with a particulate component such as TKN, since these species are soluble.

Free and saline ammonia, and total Kjeldahl nitrogen data are presented in Table 5.6. In Phase III, as part of the 48 h sampling campaign, 6 *ad hoc* samples were obtained from inlet and outlet of the pilot-scale ABR. ABR outflow samples were taken approximately one A-HRT after the equivalent inflow sample was taken, and both were analysed for free and saline ammonia. As expected, a statistically significant increase in mean free and saline ammonia concentration was observed (Paired two-sample Student's t-test, P = 0.018).

In Phase IV, 10 *ad hoc* samples were analysed for free and saline ammonia. For these samples, inflow and outflow samples were taken simultaneously. No significant increase was observed for these

samples (Paired two-sample Student's t-test, P = 2.2). The measurement obtained at the outlet was related to wastewater that entered the reactor approximately one retention time before. Thus it is not possible to determine the amount of free and saline ammonia produced by anaerobic digestion of the feed wastewater by considering an inflow and outflow sample drawn simultaneously The coefficient of variation[1] for inflow free and saline ammonia measurements was high (38 %); thus it is possible that measurements from inflow samples exceeded those measured in the outflow at a particular point in time, despite the fact that the outflow sample had increased free and saline ammonia concentration relative to that which entered the reactor one retention time before. This problem could have been overcome by analysing many more samples; the mean values calculated from the inflow and the outflow free and saline ammonia concentrations would have been better estimates of the population means, and a significant increase between inflow and outflow concentrations may have been observed.

Total Kjeldahl Nitrogen (TKN) measures the sum of organically bound nitrogen and free and saline ammonia. In a steady-state anaerobic digestion system, (i.e. no net accumulation) the TKN concentration should not change between the inlet and outlet since no TKN exits in the gas stream. During Phase IV, 8 *ad hoc* measurements of TKN were made of inflow and outflow samples, yielding average TKN concentrations of 44.6 and 37.1 mgN/ℓ respectively. This indicates a statistically significant reduction of mean TKN between inflow and outflow (Paired two-sample Student's t-test, P = 0.02). A considerable proportion of TKN is understood to be associated with organically bound nitrogen in particulate form (Raunkjear et al., 1994). Thus retention of solids in compartments of the ABR would account for the reduction in TKN observed between the inflow and outflow. However, given the large variation seen in most determinands measured, the TKN measurements were probably too few to provide reliable information about the fate of TKN in the ABR.

5.4.5 Phosphorus

Few phosphate measurements were obtained in Phase II and III. The number of measurements was too small to be able to infer any kind of statistically verifiable pattern in phosphate consumption or production. Spectrometric dissolved phosphate measurements were obtained on inflow and outflow samples in Phase IV. Mean values and 95% confidence intervals are presented in **Table 5.6**. A small but significant decrease in phosphate was observed but, since anaerobic digestion has no mechanism for the removal of significant amounts of phosphate, the apparent removal is expected to be a sampling phenomenon related to the small number of samples analysed (n=7).

5.4.6 Sulphate

No sulphate measurements were performed in Phase II and III. Spectrometric sulphate measurements were obtained on inflow and outflow samples during Phase IV. A statistically significant decrease (Student's t-test, P = 0.001) from 4.5 $mgSO_4^{2-}/\ell$ in the inflow to 0.4 $mgSO_4^{2-}/\ell$ in the outflow was observed. Mean values and 95% confidence intervals are presented in **Table 5.6**. It is probable that active sulphur-reducing micro-organisms were present and active in the ABR converting sulphate to sulphide according to Eq. 5-1 (Speece, 1996).

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[1] $\text{C of V} = \frac{\text{standard deviation}}{\text{arithmetic mean}} \cdot 100\%$

$$CH_3COO^- + SO_4^{2-} \rightarrow HS^- + 2HCO_3^-$$ **Eq. 5-1**

However, at a COD/SO_4^{2-} ratio of greater than 100:1, an average removal of 4 mgSO_4^{2-}/ℓ will not support a large sulphate reducing micro-organism population and the competition for COD substrate between sulphate reducing micro-organisms and methanogens would be negligible. The smell of H_2S was never noticed around the pilot-scale ABR during any of the three phases of operation at Kingsburgh WWTP.

5.5 24 H SAMPLING CAMPAIGN – PHASE III, DAY 99-101

Wastewater entering a treatment plant undergoes diurnal variations in strength and composition. There were concerns that the practice of regular sampling at roughly the same time of day could provide a biased indication of the actual loads that the pilot-scale ABR was receiving. The 24 h sampling campaign was undertaken to obtain an indication of the extent of the diurnal variations.

The pilot-scale ABR was monitored over 44 hours by 4 project team members in overlapping 6 hour shifts.

Samples of ABR inflow and outflow were taken at hourly intervals and analysed for COD, alkalinity and pH. Every second outflow sample (i.e. at 2 hourly intervals) was analysed for soluble COD by pre-filtering the sample through 0.45 µm acetate filter cartridges. pH and alkalinity measurements were undertaken immediately on site. Samples for soluble COD analysis were filtered immediately and the filtrate stored on ice. Samples for COD analysis were decanted into 30 mℓ screw top vials and stored on ice. Five samples each of inflow and outflow were withdrawn for free and saline ammonia and phosphate analysis. These samples were filtered through 0.45 µm acetate filter cartridges on site and stored in screw-top bottles for analysis.

Alkalinity, pH and 0.45 µm filtered COD and free and saline ammonia analyses were performed for samples from each compartment once or twice during the campaign.

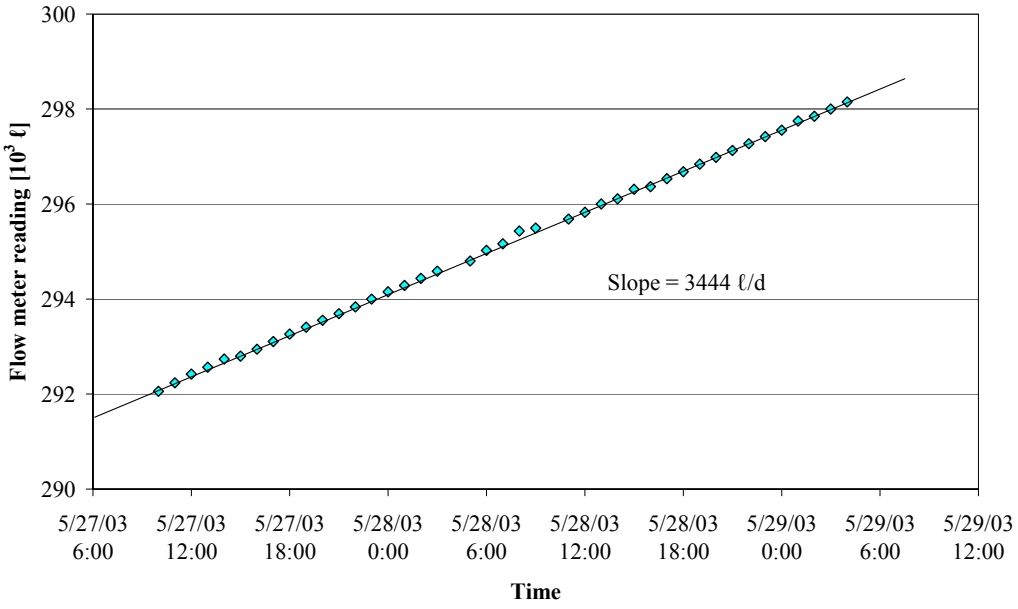


Figure 5.11: Phase III: 24 h sampling campaign flow meter readings taken hourly for 44 h

Due to the large number of samples for COD analysis (72 samples in duplicate) COD analyses were performed using small samples and a closed reflux / spectrophotometric method. For samples known to contain particulate matter, this method is less accurate than an open reflux method with larger sample size and back-titration. The School of Chemical Engineering Biochemical Engineering Laboratory could only accommodate 10 samples for open reflux COD analysis at a time. Therefore it was felt that the loss of accuracy using the closed reflux method would be less than the accuracy lost by storing the samples for the week that would be required to complete the analyses using the open reflux method. For similar reasons, samples were only analysed in duplicate rather than in triplicate as was usually favoured for these types of measurements.

Figure 5.11 presents flow rate data for the duration of the campaign. An average flow rate of 3440 ℓ/d was obtained during the campaign. This corresponded to an A-HRT of 21.8 h.

5.5.1 Diurnal variation of inflow characteristics

Figure 5.12 and **Figure 5.13** present data from COD, alkalinity and pH analyses from the ABR inflow and outflow. Outflow curves have been back-transposed 16 h in time to show possible correspondence between inflow and outflow data. (The rationale for the selection of 16 h is presented in **Appendix** A4 and **Section** 5.5.2.3).

5.5.1.1 Inflow COD concentration

The inflow COD profile (**Figure 5.12**) showed a diurnal variation in load, with highest COD concentrations observed between 10h00 and 14h00, and lowest COD loads between 01h00 and 05h00. There were no corresponding peaks or troughs on the outflow COD profile. The outflow COD concentration remained around 200 mg/ℓ, while filtered COD values were fairly constant at 64 ± 6 mgCOD/ℓ.

The purpose of the campaign was to determine how representative the regular measurements made by the project team were given the diurnal variation of wastewater characteristics.

- Regular sampling usually took place between 09h00 and 15h00 and yielded an average inflow COD concentration of 715 mgCOD/ℓ (project-team COD concentration data for Phase III). The average COD concentration for the period 09h00 to 15h00 obtained from the 24 h sampling campaign was 751 mgCOD/ℓ.

- The average inflow COD value calculated from all available Phase III data (project team and municipality data) was 676 mgCOD/ℓ. The overall average inflow COD concentration value for the 24 h campaign was 564 mgCOD/ℓ Thus it appears that routine sampling was done at a time when wastewater strength was greater than the average wastewater concentration, resulting in calculation of average wastewater strength and OLR that was higher than was actually experienced by the ABR.

Generally, both the volumetric flow rate of wastewater received by a wastewater treatment plant, and the wastewater strength will follow a diurnal pattern with a significant decrease due to reduced water use at night time. A flaw existed in the design of the flow control programme in that a constant volumetric flow was delivered to the pilot-scale ABR irrespective of the strength of the wastewater. Thus the pilot-scale ABR received on average a lower strength of wastewater than the rest of the plant would have received. The flow-weighted average value for COD concentration in Kingsburgh WWTP

wastewater was probably higher than that experienced by the pilot-scale ABR, as shown in the averages presented above, where a difference of more than 100 mgCOD/ℓ was observed.

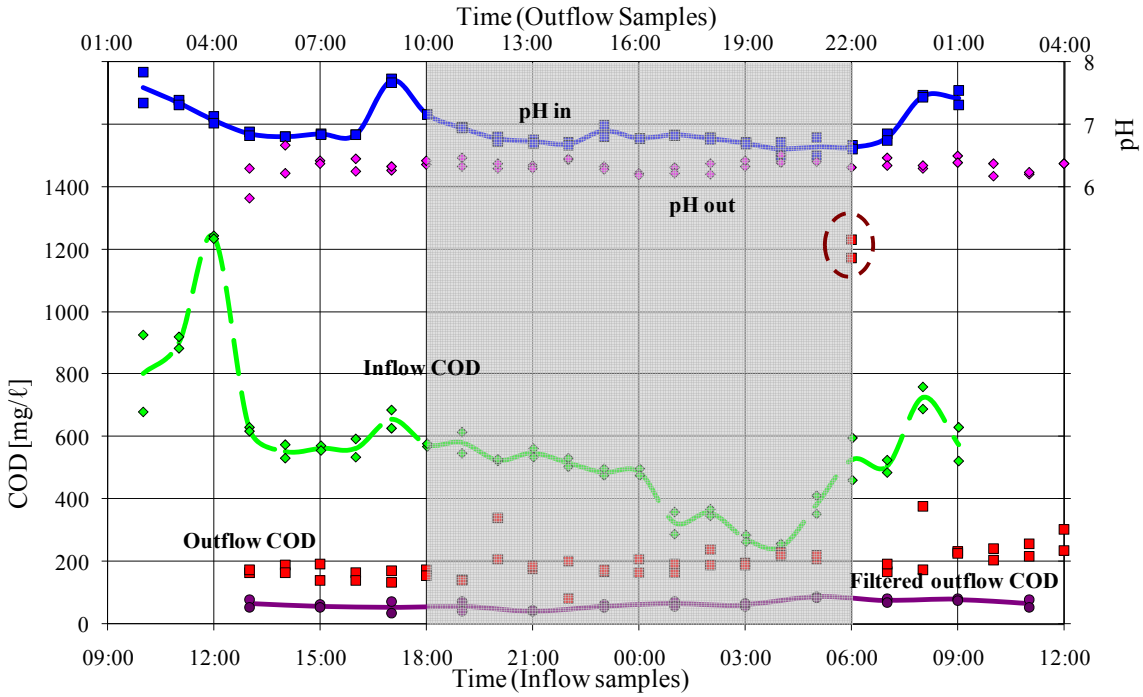


Figure 5.12: Phase III: 24 h sampling campaign COD concentrations and pH values showing diurnal variation in inflow and outflow streams from the pilot-scale ABR. Shading indicates night (for inflow sampling).

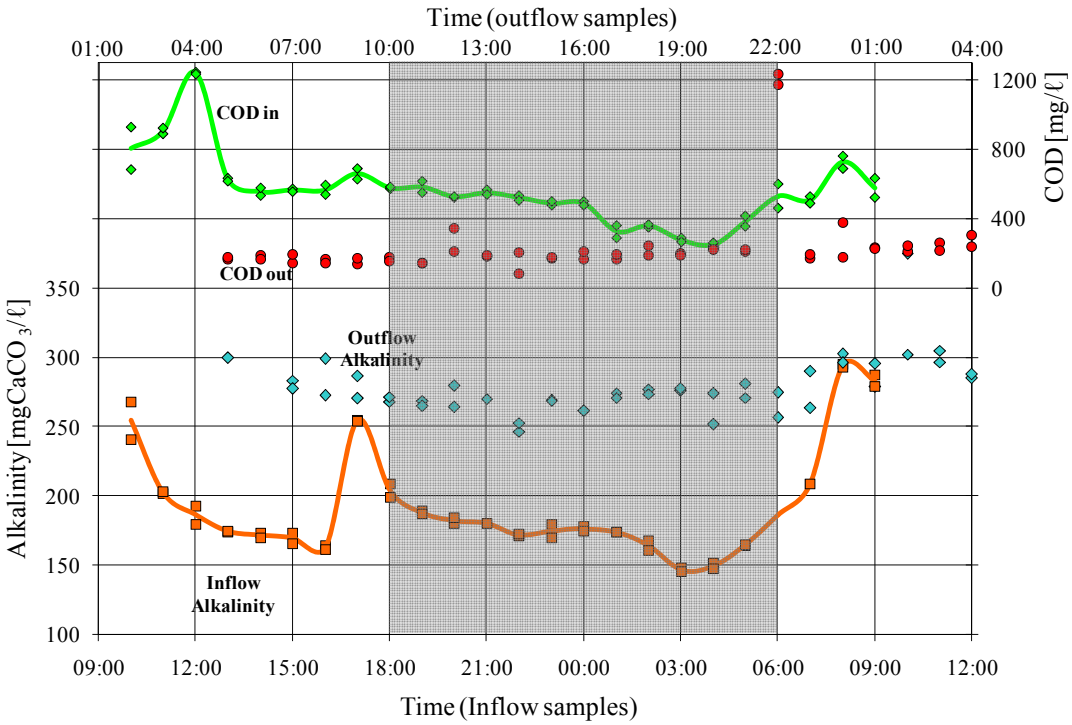


Figure 5.13: Phase III: 24 h sampling campaign COD and alkalinity concentrations showing diurnal variation in inflow and outflow streams from the pilot-scale ABR. Shading indicates night (for inflow sampling).

Thus pilot-scale ABR inflow COD concentration data and related calculations presented from any of the operating periods may be considered to be overestimated by as much as 100 mgCOD/ℓ. More data of this type (hourly samples over 24 h) are required to support this conclusion.

5.5.1.2 Inflow alkalinity concentration

Figure 5.13 presents values for alkalinity in inflow and outflow samples. In general, the outflow alkalinity is higher than the inflow alkalinity concentration, as expected. There appears to be a systematic diurnal oscillation of inflow alkalinity values, although several days worth of data would be required to confirm this. Highest alkalinity values were observed between 07h00 and 11h00. The average value was calculated to be 180 mg$CaCO_3$/ℓ and the average value for the normal sampling hours (10h00 to 15h00) was approximately 193 mg$CaCO_3$/ℓ. These values are not significantly different to the average value reported from all available data during regular sampling (an average value of 193 mg$CaCO_3$/ℓ was obtained during Phase III).

5.5.1.3 Inflow pH value

Figure 5.12 shows inflow and outflow pH values measured in duplicate on an hourly basis over a 24 h period. Inflow pH values varied between 6.5 and 7.5. Outflow values showed smaller variations, and were consistently lower than inflow values.

As with COD and alkalinity concentrations, an increase in inflow pH value was observed between 08h00 and 13h00. The median pH value measured for the regular sampling times of day (10h00 – 15h00) was 7.0. This is the same as the median value of various combinations of regular data (Municipal data only, project team data only and both combined). However the overall median pH value for the sampling campaign was 6.8. Thus the pH values presented for inflow wastewater for regular sampling may be slightly high when considering the overall effect of pH of the inflow wastewater on ABR performance (i.e. for modelling purposes, it may be appropriate to use an inflow wastewater pH value of 6.8, rather than the value of 7.0 that arises from data from regular sampling). However, more than 1 data set is required to confirm this observation.

5.5.2 Diurnal variation of outflow characteristics

Analysis of outflow characteristics led to a number of observations about the action of the ABR on the wastewater as it passed through the reactor.

5.5.2.1 Burping

At 22h00, the outflow sample COD concentration exhibited a sudden spike to 1 230 mgCOD/ℓ (circled in **Figure 5.12**), caused by a sudden and brief expulsion of sludge. Visually, the samples appeared substantially more turbid than usual with more solids than usual settling out at the bottle of the sample vial during storage. There was no significant change in outflow COD or filtered outflow COD values for the sampling times immediately before and after this measurement, suggesting that there was no biological upset that caused the high outflow COD observed. The incident was ascribed to a *burping* phenomenon, whereby gas production or other fluid effects caused mixing of the sludge in the last compartment, with a short term overflow of the sludge to the outlet. Biochemical upsets are most commonly associated with inhibited methanogenesis, and corresponding increase in acid and soluble COD concentration, which was clearly not the case in this event. *Burping* was (visually) observed from time to time in all operating periods.

5.5.2.2 Damping

Figure 5.12 and **Figure 5.13** show that treatment in the pilot-scale ABR has a damping effect on the pH values and COD and alkalinity concentrations observed in the outflow. This can be shown using measurements of variance:

- The coefficient of variation (defined on page 98) for the inflow COD concentration is 37%, while that for the outflow is 26%. (This value is calculated ignoring the 22h00 "burp" since it does not appear to be a function of the overall reactor biological or hydraulic dynamics, but rather is caused by specific solids and gas dynamics in the last compartment)

- The pH range measured in inflow samples was 1.23 pH units (minimum = 6.60, maximum = 7.83), while the range of outflow pH values was 0.85 pH units (minimum = 5.83, maximum = 6.67)

- Alkalinity concentrations showed higher co-efficient of variation in the inflow (21%) compared to the outflow (7%).

The cause of this damping effect was not investigated, however, it is expected that the damping of oscillations in concentration was due to the combined effect of axial mixing (non-plug flow conditions in and between compartments) and non-linear dependence of biological reaction rates on substrate and product concentrations.

5.5.2.3 Applied vs. Apparent hydraulic retention time

The hydraulic retention times (A-HRT) quoted throughout this thesis were calculated from the applied flow rate and the empty reactor volume i.e.

$$\mathrm{HRT}[\mathrm{h}] = \frac{\text{reactor volume}\,[\mathrm{m}^3]}{\text{Flow rate}\,[\mathrm{m}^3/\mathrm{h}]} \qquad \textbf{Eq. 5-2}$$

This implies that fluid spends the same amount of time in the ABR as it would if passed through a plug flow reactor in which the entire working volume was available for through flow. However, retention of solids in each compartment would have resulted in a reduction of the volume of the compartment available for flow of the liquid phase. This concept is depicted graphically in **Figure 5.14** where reactor volume available for fluid flow is equal to the total compartment volume less the volume filled with sludge.

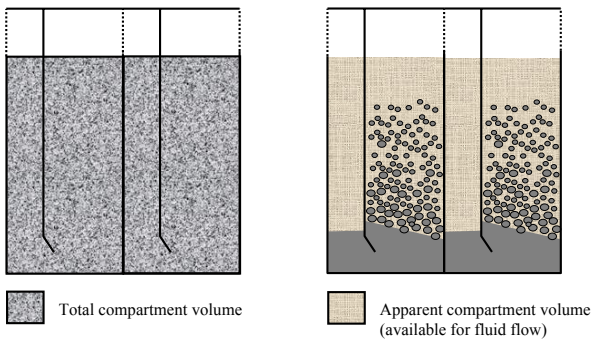


Figure 5.14: Apparent compartment volume is related to the total compartment volume and the volume occupied by solids that are retained in the compartment.

Thus for a particular flow rate, the average amount of time spent by a package of fluid in any compartment should have been less than that predicted by the clean water CFD modelling undertaken by Dama et al. (2001).

No tracer tests were undertaken on the pilot-scale ABR. However, analysis of the variations in the inflow and outflow alkalinity data led to the tentative estimation that the bulk of the fluid spent approximately 16 h (the apparent HRT) in the ABR during the 24 h campaign (i.e. at a flow rate of 3444 ℓ/d), indicating that a fraction equal to $\left(1 - {}^{16h}/_{22h}\right)$ or 27 % of the reactor volume was taken up by material through which soluble and liquid components did not pass, or through which only slow diffusion was possible. The rationale and analysis for this is presented in **Appendix** A4.

5.5.2.4 Comparison between 24 h campaign outflow data and regular sampling outflow data for phase III

Comparison of the outflow COD concentration values from the sampling campaign with regular measurements obtained during Phase III showed that there was no significant difference between the two data sets (Student's T-test, unequal variances, P = 0.15). The probability that the two data sets are not significantly different increases to P = 0.75 if the peak outflow COD values observed at 22h00 are ignored.

Comparison of campaign outflow alkalinity measurements with regular measurements showed no significant difference (P= 0.36) between the data sets.

5.5.3 Conclusions – 24 h sampling campaign

By monitoring inflow and outflow characteristics of the pilot-scale ABR over 44 h (24 hourly samples each of the inflow and outflow), the following conclusions were drawn:

5.5.3.1 COD concentration dynamics

- Significant diurnal variation of inflow COD concentration was observed. Regular sampling time for normal operation (i.e. not during the campaign) coincided with higher inflow COD concentrations than the daily average. Thus inflow sample measurements for regular operation may have overestimated the average inflow COD concentration to the pilot-scale ABR by as much as 100 mgCOD/ℓ. However, this cannot be stated with certainty on the basis of only one day's worth of analyses. Therefore in the analysis that follows, the measured mean COD value of 680 ± 25 mgCOD/ℓ is used, while keeping in mind that it may be higher than the true mean.

- The outflow COD and alkalinity concentration profiles showed less variation around the mean value than the inflow profile. Thus the pilot-scale ABR played some role in damping out extreme conditions due to diurnal oscillations of feed strength.

- A spike in outflow COD concentration was observed for one sample along with observations of increased turbidity and larger amounts of sludge in the samples. However, these observations were not matched by any disruptions to the outflow filtered COD, pH or alkalinity profiles. The event was described as a 'burping" phenomenon in which gas production and hydraulic activity in the last compartment caused an overflow of solids into the outflow above the usual amount.

- The outflow COD values were not significantly different to those observed during regular operation. Thus it is concluded that diurnal variations will not affect the validity of the mean values of outflow COD calculated for regular operation.

5.5.3.2 pH value dynamics

Outflow pH values did not appear to be strongly dependent on the corresponding inflow pH values. Outflow pH values did not show large variations during 24 h, but were consistently lower than corresponding inflow values. It is concluded that the pilot-scale ABR was effective in buffering the pH value, resulting in stable digestion in the later compartments despite variations in inflow pH values.

5.5.3.3 Alkalinity dynamics

- Oscillations were seen in both inflow and outflow alkalinity profiles.

- Throughout the 24 h period, the outflow alkalinity was consistently greater than the corresponding inflow value.

- Both inflow and outflow alkalinity values were similar to those measured during regular sampling

- Comparison of inflow and outflow alkalinity values led to the tentative estimation that the apparent HRT i.e. the amount of time that the bulk of the liquid flow spent in the pilot-scale ABR, was around 16 h. The corollary is that approximately 27% of the reactor volume is not available for liquid flow due to the presence of sludge.

5.6 COMPARTMENT DYNAMICS: PHASE II - IV

In this section the dynamics of various determinands measured within the compartments of the pilot-scale ABR are presented.

5.6.1 Solids levels and concentration

Figure 5.15(a), **Figure 5.16(a)** and **Figure 5.17(a)** show the settled sludge bed height in each compartment as a fraction of the total compartment height for Phase II, III and IV respectively.

Sludge bed height data were obtained using a core sampler (**Appendix** A2**.**1) and recording the height of the sludge bed after 5 min settling time. Settled sludge bed height is not an absolute indication of the amount of sludge in a compartment since the bulk density of sludge (mass of sludge granules per unit volume of sludge bed) can change significantly according to extent of granulation/dispersion, pH, redox potential, operating conditions and inert content. However, it provides a good visual indication of how the amount of sludge in each compartment varies with time, and how the sludge load varies from one compartment to the next.

Figure 5.15(b), **Figure 5.16(b)** and **Figure 5.17(b)** show the calculated volume of settled sludge bed present in the reactor as it changes with time.

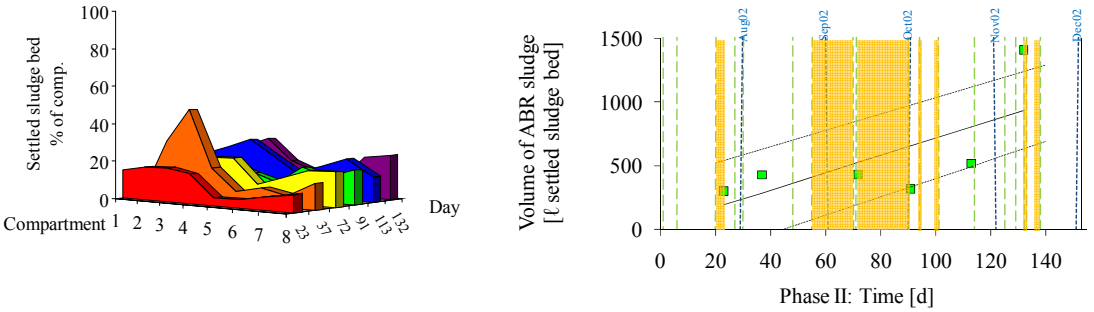


Figure 5.15: Phase II: (a) Settled sludge bed height in ABR compartments (% of compartment height, n=6). (b) Volume of sludge in the upflow compartments of the pilot-scale ABR showing linear regression line with 95% confidence interval on the regression (------).

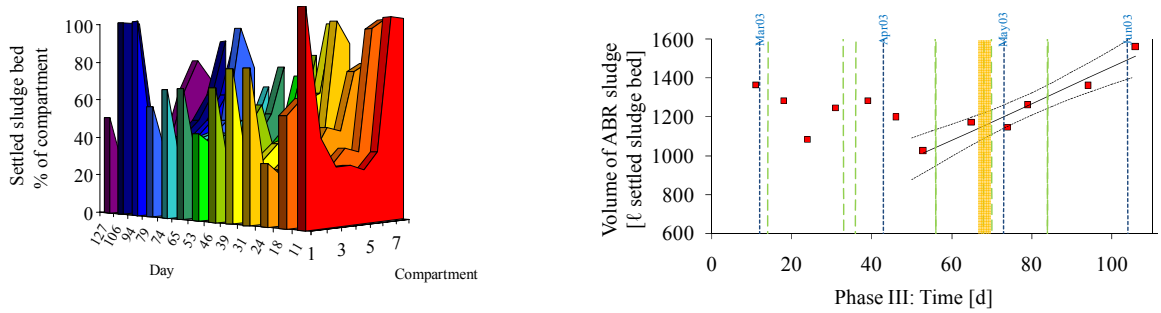


Figure 5.16: Phase III: (a) Settled sludge bed height in ABR compartments (% of compartment height, n=13). (b) Volume of sludge in the upflow compartments of the pilot-scale ABR showing linear regression line with 95% confidence interval on the regression (------) for region of operation with no major sludge loss incidents.

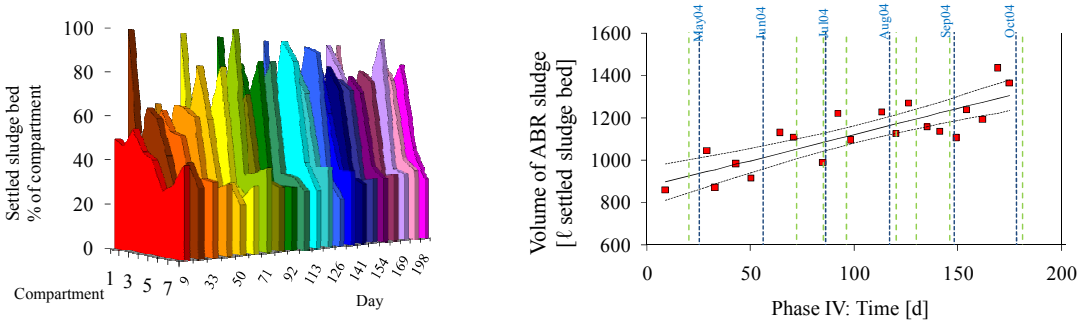


Figure 5.17: Phase IV: (a) Settled sludge bed height in ABR compartments (% of compartment height, n =21). (b) Volume of sludge in the upflow compartments of the pilot-scale ABR showing linear regression line with 95% confidence interval on the regression (-----).

Settled sludge volume is calculated from the settled bed height in each upflow compartment and thus excludes the sludge volume in the downflow side of each compartment. There are two reasons for this representation:

- The sampling ports on the downflow side of each compartment had rusted closed during Phase I. A decision was made to leave them in an un-openable state to preserve the gas seal on the reactor. Therefore downflow compartment data were not available.

- Information about the volume of accumulated sludge is important for planning of desludging operations. With the current reactor design, desludging would only take place from the upflow side of the reactor; hence volume of sludge calculated only from the upflow compartments provides sufficient indication of the desludging requirements for the system.

Table 5.3: Results of regression analysis on solids accumulation in the pilot-scale ABR during operation at Kingsburgh WWTP

Phase	Calculated quantity	Units	Average slope of regression [Confidence interval on slope]	Significance of regression (P)	No. of observations used in regression
II	Sludge volume	m^3 settled sludge/year	2.49 [-1.13, 6.12]	0.128 (not significant)	5
III	Sludge volume	m^3 settled sludge/ year	3.46 [2.28, 4.64]	0.00125 (significant)	6
IV	Sludge volume	m^3 settled sludge/ year	0.901 [0.602, 1.20]	5.87×10^{-6} (highly significant)	20
IV	Mass of total solids (dry)	kg dry solids/year	60.7 [33.5,87.8]	2.54×10^{-4} (highly significant)	17
III/IV	Ratio of sludge volume slopes: Phase III/Phase IV	-	3.84 [2.55, 6.06]	-	-

The volume of sludge was calculated as follows:

$$V_{t,\ settled\ \ sludge} = \left(\sum_i y_i\right) \cdot (\text{c/s area of up - flow compartment}) = \left(\sum_i y_i\right) \cdot (w \cdot l) \qquad \textbf{Eq. 5-3}$$

where y_i is the height of settled sludge in compartment i and w and l are the width and length respectively of the upflow side of each compartment.

There is a clear difference in the height of the beds between the different operating periods, with Phase II (**Figure 5.15**) characterised by significantly lower sludge beds than in Phase III and Phase IV.

5.6.1.1 Phase II: Solids levels

There were many sludge washout incidents during Phase II. In these instances, high flow through the reactor (due to poor control of feed rate, or sudden emptying following blockage of the outlet) caused high internal flows with high upflow velocities, and carryover of sludge. Greatest settled sludge bed height was usually found in compartment 2, although the difference between compartments was generally not great. **Figure 5.15(b)** seemed to show a slight increase in volume of sludge with time during Phase II. However, a regression analysis showed that there was no significant increase in accumulated sludge volume with time, due to the small number of observations and the significant scatter in the data (**Table 5.3**). Furthermore, such a correlation would not provide any indication of sludge volume accumulation rates under stable operation due to the many sludge washout events in this phase.

5.6.1.2 Phase III: Solids levels

A gradual increase in the overall sludge bed height with time is observed in Phase III. **Figure 5.16(b)** shows a consistent increase in sludge bed volume from day 53 onwards. (In the period

before this, there were two incidents that may have caused washout of sludge, on day 14 and day 36. these incidents resulted in reduction of the sludge bed volume and thus data before day 36 were not included in the regression.) The regression considers 53 days of stable operation from day 53 to day 106. The regression was found to be significant (**Table 5.3**) and the sludge volume increased at a rate in the 95% confidence interval from 6.2 to 12.7 ℓ/d.

In Phase III, there were few sludge washout incidents and there was a much less even distribution of sludge between compartments (**Figure 5.16 a**). The average sludge bed height was more than double that observed in Phase II, and the highest sludge bed was usually observed in compartment 1.

5.6.1.3 Phase IV: Solids levels

In Phase IV, the flow rate and applied OLR were reduced. Operation was stable as a result of improvements in the control algorithm, and few incidents of any kind were observed. There were no significant sludge washout incidents. The average height of the sludge beds (**Figure 5.17 b**) increased at a slower rate than in Phase III, and the shape of the profiles also changed, with higher settled sludge beds in the earlier compartments (**Figure 5.17 a**). **Figure 5.17 (b)** presents the calculated sludge volume in the upflow compartments of the pilot-scale ABR for Phase IV operation with a linear regression between time and sludge load. The linear regression was highly significant (**Table 5.3**) and the regressed slope of sludge volume accumulation was found to fall in the 95% confidence interval from 1.6 to 3.3 ℓ/d.

For Phase IV, total solids data were also available for the average concentration in the upflow side of each compartment.

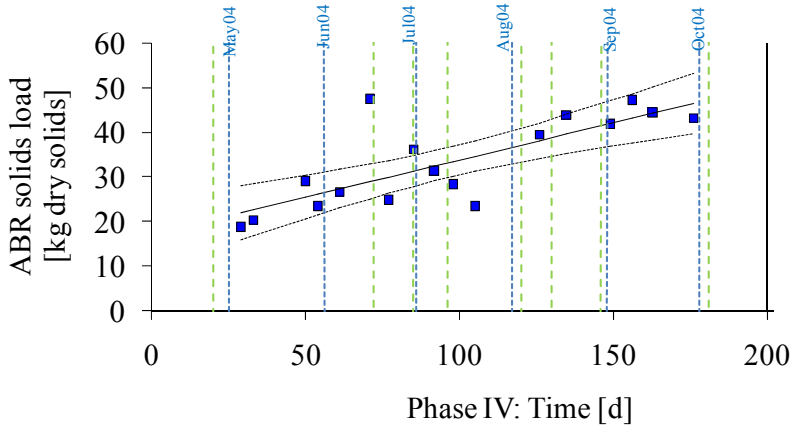


Figure 5.18: Phase IV: Mass of total solids (dry) in the upflow compartments of the pilot-scale ABR showing linear regression line with 95% confidence interval on the regression (------).

A total solids load was calculated for the pilot-scale ABR analogous to the total settled sludge volume according to Eq. 5-4:

$$TS_t = \left(\sum_i \left(C_{TS,i} \cdot Y_{i,\max}\right)\right) \cdot (\text{c/s area of up - flow compartment}) = \left(\sum_i \left(C_{TS,i} \cdot Y_{i,\max}\right)\right) \cdot (w \cdot l)$$

Eq. 5-4

Where $C_{TS,i}$ is the average total solids concentration in a mixed core sample from compartment i, $Y_{i,max}$ is the maximum height of compartment i and w and l are compartment width and length respectively.

As with the total volume of sludge calculated with Eq. 5-3, this quantity only takes into account the solids content of the upflow side of each compartment and does not reflect the total mass of solids in the reactor and therefore cannot be used for mass balance purposes.

Figure 5.18 shows that total solids accumulated in the upflow side of each compartment at a rate between 34 and 88 kg dry solids/year (**Table 5.3**).

5.6.1.4 Rate of sludge accumulation: dependence on organic loading rate

It was surprising that although the OLR in Phase III was double that of Phase IV, the Phase III sludge accumulation rate appears to be significantly more than double that of Phase IV. The ratio of the sludge accumulation rate in Phase III to that of Phase IV had a 95% confidence interval of 2.6 to 6.1 (**Table 5.3**), calculated using Fieller's theorem, Appendix A2.4) i.e. it can be said with 95% confidence that the rate of sludge accumulation in Phase III was between 2.6 and 6.1 times greater than that of Phase IV despite the fact that the OLR was approximately double.

Table 5.4: Calculation of sludge accumulation rates normalised by organic loading rate for Phase III and Phase IV.

	Sludge accumulation rate	**Organic loading rate**	**Normalised sludge accumulation rate**
Units	m^3 sludge/year	kg COD applied / m^3 reactor volume.year	ℓ sludge accumulated / (kg COD applied)
Phase III	3.46	269	4.3
Phase IV	0.901	146	2.1
Phase IV (kg dry solids)	60.7[kg dry solids/year]	146	0.14 [kg dry solids/kg COD applied]

To explain this observation, the sludge accumulation rate was normalised using OLR[1]. Using average values only, the normalised sludge accumulation rates (i.e. amount of sludge accumulated per kg COD applied) were calculated and are presented in **Table 5.4**.

Initially, it was expected that the *normalised* sludge accumulation rate would be *higher* in Phase IV than in Phase III since the lower feed flow rate resulted in lower upflow velocity in the upflow compartments, and this was expected to lead to better sludge retention. However, **Table 5.4** shows a significantly *lower* normalised sludge accumulation rate in Phase IV than in Phase III.

The most obvious explanation for the difference in normalised accumulation rates is that the additional solids accumulated in Phase III were undigested biodegradable particulate material originating from the feed material. The implication is that the resident anaerobic biomass population in the pilot-scale ABR in Phase III was either not sufficiently concentrated or not sufficiently active to convert

[1] The sludge accumulation rate per organic loading quantity was calculated roughly for the purposes of simplifying the calculations and therefore confidence intervals were not calculated.

biodegradable particulates at the rate at which they entered the ABR. Conversely, at the lower flow rate applied in Phase IV, the rate of influx of biodegradable particulates did not exceed the rate of their conversion to the same extent as in Phase III. This would have been partly due to the fact that biodegradable particulates were supplied at half the rate in Phase IV, but may also have been due to the establishment of a more stable anaerobic community at the lower upflow velocities applied in Phase IV. The latter point will be shown to be the case in **Section 5.7.2**, where microbiological studies on pilot-scale ABR sludge are presented. Furthermore examination of the average amount of free and saline ammonia produced due to digestion in the ABR is greater in Phase IV than in Phase III, indicating that a greater extent of treatment has been achieved (**Table 5.6**).

5.6.1.5 Mechanism of sludge carryover between compartments

Further analysis of the compartment dynamics with respect to settled sludge bed level and total solids concentration revealed an interesting insight into the mechanism of solids carryover between compartments.

Figure 5.19 presents solids variations in compartments 4, 5 and 6 of the pilot-scale ABR during Phase IV (2004). In **Figure 5.19 (a)**, sludge level is calculated according to Eq. 5-5.

$$Sludge\ level = \frac{\sum_i \left(x_i Y_{i,\max}\right)}{\sum_i Y_{i,\max}} \cdot 100\% \qquad \textbf{Eq. 5-5}$$

Where

$$x_i = \frac{y_i}{Y_{i,\max}} = \frac{height\ \ of\ \ settled\ \ sludge\ \ bed}{\max .\ comp .\ height}$$

Compartment 4 showed relatively constant settled solids level and total solids concentration with time. When fluidised, the sludge blanket produced filled the entire compartment. Between day 50 and 60, an increase in both settled solids level and total solids concentration is seen in compartment 5, which is followed by a similar increase in compartment 6 around day 100.

These data indicate that once the fluidised solids bed in a compartment increases until its height matches the compartment height, solids overflow from that compartment to the next. Before this overflow occurs, solids increase in the following compartment is probably mainly due to biological growth. However, once the previous compartment solids begin to overflow, solids accumulation rates exhibit a significant increase due to solids displacement from the previous compartment. Before compartments become filled with sludge, the population dynamics are controlled by the growth rates of different micro-organism species under the substrate concentration and up-flow velocity conditions. However, once the compartments become filled by the fluidised sludge bed, overflow from one compartment to the next results in the next compartment having a population that is dependent both on substrate-related kinetics within the compartment, and on the population characteristics of the sludge overflowing from the previous compartment.

Figure 5.20 presents sludge level and total solids concentration for compartments 7 and 8. The scale is the same as in **Figure 5.19 (a)** and **(b)** for purposes of comparison. It can be seen that although from day 100 onwards compartments 4, 5 and 6 were essentially full and overflowing, the sludge accumulation rate in compartments 7 and 8 remained relatively low.

It is not clear why the rate of accumulation and overflow in compartments 4, 5 and 6 are so high in comparison to those of compartments 7 and 8 (except that accumulatable material was exhausted by the time flow reaches these compartments). However, these results suggest that total reactor sludge load, and especially overflow of solids from one full compartment to the next play an important role in determining where the maximum sludge accumulation rate occurs in the reactor at any point in time.

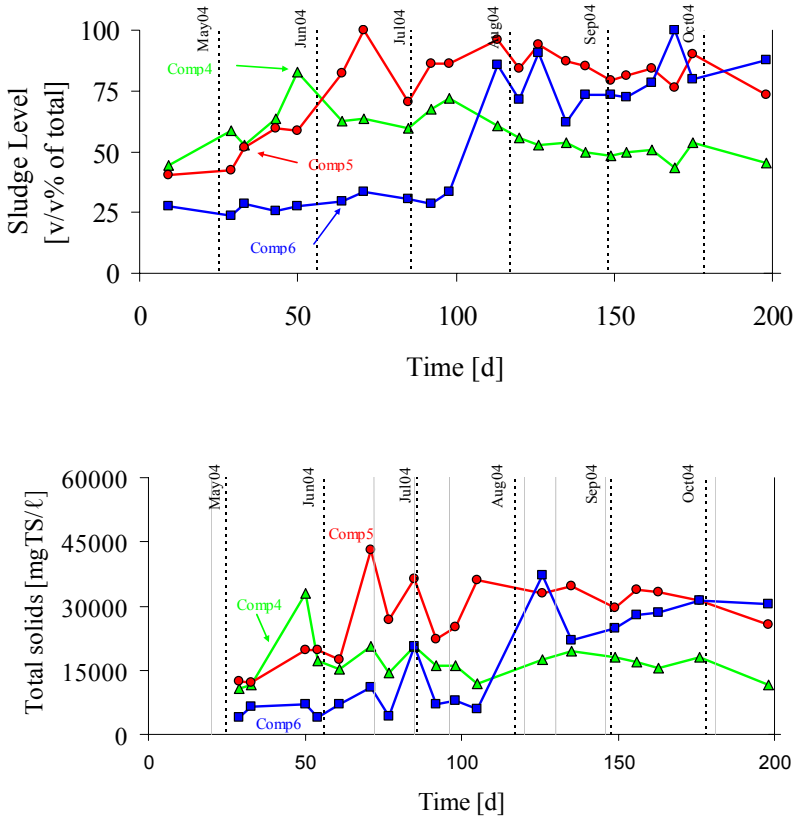


Figure 5.19: **Phase IV: Solids accumulation and carryover in compartments 4, 5 and 6. (a) Sludge level in each compartment as a percent of total compartment height; (b) Average total solids concentration in each compartment**

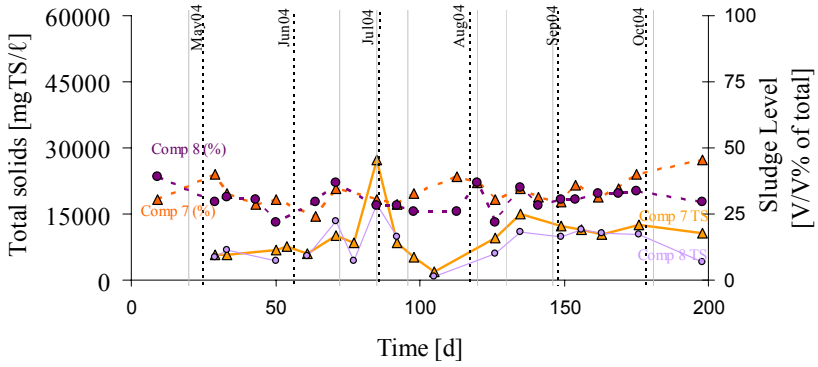


Figure 5.20: **Phase IV Total solids data for compartment 7 and 8. Compartment 7 data are plotted in orange; data for compartment 8 are purple. Sludge level is plotted with a dashed line and totals solids with a solid line.**

5.6.2 pH value and soluble COD concentration

Figure 5.21 (a), **(b)** and **(c)** show the minimum, maximum and median values of pH measured during each of the three phases of operation at Kingsburgh WWTP. The range from which the minimum value is calculated for **Figure 5.21** (Phase II) did not include the pH measurements recorded during the ABR souring incident. Only normal operation is reflected.

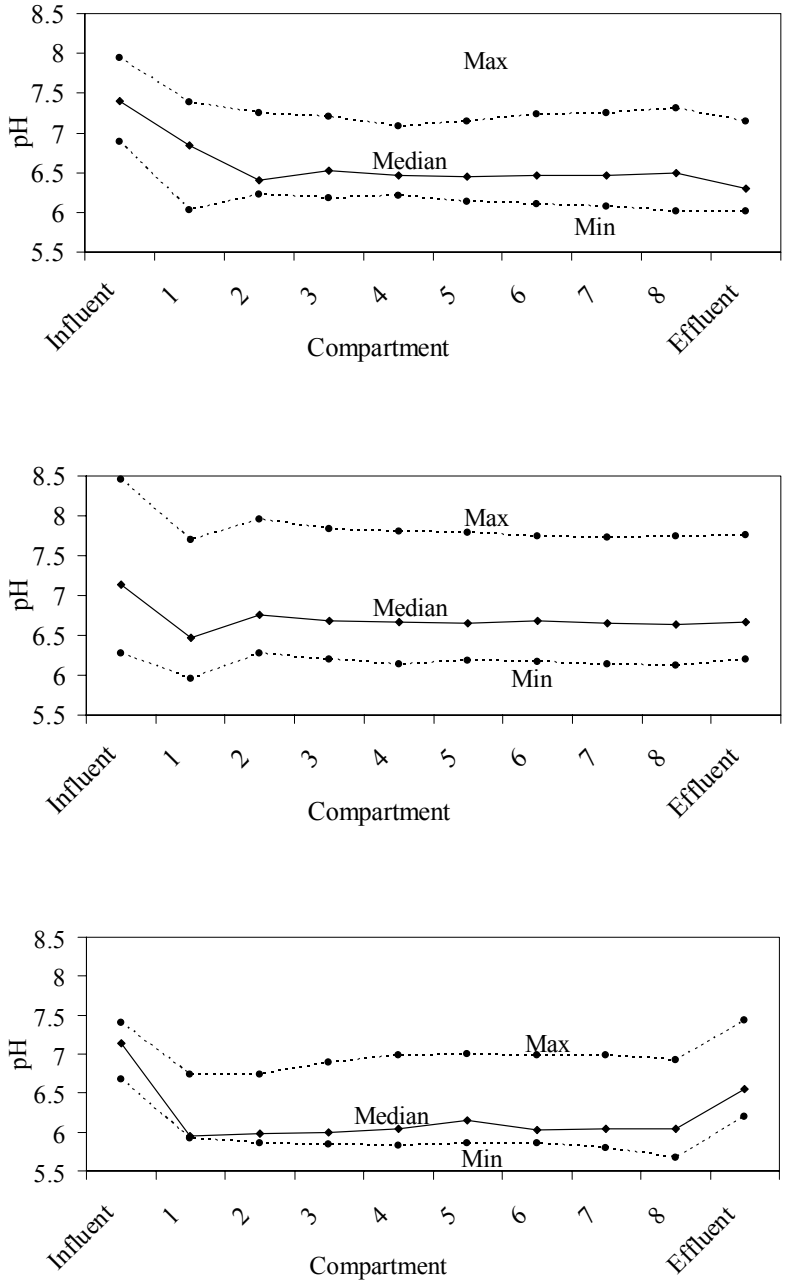


Figure 5.21: pH values measured in inflow, outflow and upflow compartments in Phase II (a), Phase III (b) and Phase IV (c). Median, minimum and maximum pH values are shown for each set.

Speece (1996) states that the *proper pH* for anaerobic digestion should fall between values of 6.5 and 8.2. In Phase II, 6 on-site measurements of pH were made for each of the compartments (excluding those recorded during and immediately after the souring incident. The median recorded pH value in

most compartments was exactly 6.5, with a median value of 6.4 in the first compartment. In Phase III, 14 measurements of pH were recorded for samples that had been returned to the UKZN laboratory. The median pH value was between 6.6 and 6.7 for most compartments except the first, where pH values were usually slightly below 6.5. In Phase IV, 6 on-site measurements of pH value were recorded. Median values for each compartment were well below 6.5, with values regularly decreasing below 6 in the early compartments.

All data sets show an initial decrease in pH value followed by a slight increase or stabilisation in pH value. The increase between compartment 8 and the outflow in Phase III and Phase IV data is a result of aeration of the outlet stream and subsequent release of CO_2 gas to the atmosphere at the effluent screen. (Effluent screens were installed early in Phase III, **Section 5.1**). Similar increases between compartment 8 and outflow pH value were observed in Phase III when considering a data set for a particular day in which measurement was made, but the magnitude was less than in Phase IV due to the higher overall pH values. This is not seen clearly in Phase III due to the selection of overall minimum, median and maximum data points

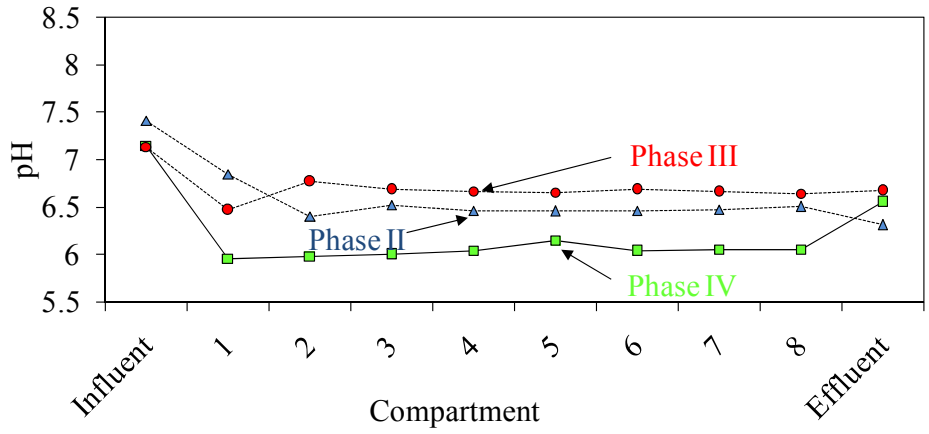


Figure 5.22: Median pH values for inflow, compartments and outflow in the ABR obtained in the Phase II (blue), Phase III (red) and Phase IV (green).

Figure 5.22 compares median values for pH obtained during the different years of operation. Caution should be employed in drawing conclusions from these data since relatively few data points were available in all operating periods, especially in Phase II and Phase IV. However, it is significant, and initially surprising that the lowest median pH values are observed in the Phase IV, which also has the lowest organic load of the three phases. Further the only data that fall into the recommended operating range for pH value are those from Phase III, which had the highest continuous loading rate. In **Section** 5.6.1.4 it was proposed the lower flow rate in Phase IV resulted in better overall extent of treatment of particulate biodegradable material through the establishment of a more stable anaerobic population. Using this hypothesis, it may be possible to explain the differences in median pH value per compartment between Phase III and Phase IV if it is proposed that poor retention of anaerobic micro-organisms in Phase III resulted in generally lower rates of anaerobic digestion occurring in compartments 2 to 8 than in Phase IV; thus after the initial decrease in pH in compartment 1, pH recovery ensued in subsequent compartments due to low acidogenic activity, or to CO_2 stripping due to gas exchange with the atmosphere at low gas production rates (**Section** 4.4.3 and **4.5.3**). Conversely in Phase IV, the fact that the pH value in compartment 2 did not increase may indicate that acidogenic activity has continued at reasonable rates.

Figure 5.23 shows the mean soluble COD profile calculated from measurements of soluble COD in the inflow, outflow and mixed upflow side of each compartment for Phase IV. Soluble COD concentration increased between the inflow and compartment 1 as a result of hydrolysis and acidogenesis in compartment 1; acid production caused a shift of COD from the particulate to the soluble phase, with a corresponding dip in pH value (**Figure 5.21 c**). Methanogenesis in this compartment was unable to remove all the soluble COD that was produced. Hence the pH remained lower and the soluble COD was higher in compartment 1 than the feed. In subsequent compartments, the pH value recovered slightly, and the concentration of soluble COD decreased. This implies that there was a shift in rate-limiting step from methanogenesis in compartment 1 to hydrolysis in later compartments; i.e. remaining particulate COD was hydrolysed slowly to soluble COD and acid, which then underwent methanogenesis at the rate at which it is produced. Consequently, a roughly constant pH value and soluble COD concentration was observed from compartments 2 to 8, and in the outflow.

It should be noted that the significant increase in soluble COD in compartment 1 relative to other compartments is a function of the fact that most particulate COD is retained in compartment 1 due to the baffled design of the pilot-scale ABR. Thus by far the highest rates of particulate solubilisation must occur in this compartment, resulting in the highest observed soluble COD levels.

Soluble COD data for Phase IV show that VFA did accumulate to some extent in compartment 1; however for the same operating period, although VFA concentration appears to decrease in subsequent compartments (**Figure 5.23**), the pH value does not increase in compartment 2 (**Figure 5.22**). It is concluded that the mechanism observed here is not true phase separation where different micro-organisms are dominant in different zones of the system due to different pH conditions, but rather that different solids concentrations in the different zones result in different net rates of COD solubilisation. The similar pH values indicate that the probability of different micro-organisms dominating as a result of pH inhibition is not large. In other words, true phase separation where methanogenic micro-organisms are protected from low pH values by creation of a methanogenic zone does not exist in this case. However, the ratio of methanogenic to acidogenic micro-organisms may be higher after compartment 1 than in compartment 1 because of the different ratios of substrate concentrations in the different zones.

Two other conclusions were drawn from the pH and soluble COD data:

- pH data that are consistently below a value of 7, and often below 6.5 indicate that anaerobic digestion in the ABR was poorly buffered during treatment of dilute wastewater. This was a function of the low inflow alkalinity concentration and low alkalinity generation potential of the relatively dilute wastewater.

- Despite the fact that anaerobic digestion in the pilot-scale ABR was found to be poorly buffered, and pH values were observed to decrease significantly between the inflow and first compartment, pH values remained above a value of 6 in the later compartments for all the flow conditions tested (except for the souring incident in Phase II). The implication is that the baffled configuration assisted in protecting later compartments, and thus the outlet stream from low pH conditions that may have developed as a result of acidogenesis in the first compartment.

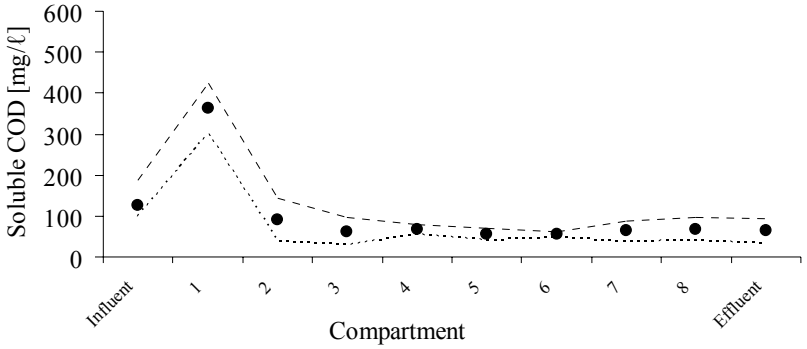


Figure 5.23: Phase IV: COD concentrations in compartments of the pilot-scale ABR. Data points indicate mean values for each compartment. The dashed lines show upper and lower 95% confidence limits on the mean. n=18 except for compartment 1 (n=15) inflow (n=16) and compartments 3 and 4 (n=17).

5.6.3 pH values during souring incident

During Phase II, a souring incident occurred where low pH values were observed in all compartments and the outflow of the ABR. It is theorised that the initial cause of the souring was illegal dumping of septic tank contents into the Kingsburgh incoming wastewater, a practice that is known to occur from time to time in the middle of the night. **Figure 5.24** shows a series of pH profiles on different days during Phase II around the souring incident. The dumping would have caused a slug of low pH, high COD wastewater to enter the ABR, causing organic overload, and inhibition of methanogenesis[1].

The pH values during normal operation, souring and 3, 4, 5, 9 and 10 days after souring are shown in **Figure 5.24** Sour anaerobic conditions resulted in pH values around 4.5, the pK_a value of acetic acid. It is expected that souring occurred first in compartment 1, and was propagated to subsequent compartments. Measurements on the day of souring (0) were taken at around 13h00. Illegal dumping is reported to occur between 20h00 and 04h00 suggesting that between 9 and 17 h had passed between souring and pH measurement, hence sour compartment liquors would have been washed down to later compartments and replaced with lower strength partially treated liquors by the time the measurements were made. Data for the day of souring (day 0) showed that pH values as low as 4.5 were only seen in compartments 7 and 8, indicating that the first 6 compartments had already begun to recover. Assuming the apparent HRT to be 16 h , the time for wastewater flow to reach the sample valve in compartment 7 is 7/8 x 16 h = 14 h (**Section** 5.5.2.3). Consequently, it is supposed that a high COD load was delivered to compartment 1 before 23h00 the previous evening, and by the time the reactor was sampled, the first 6 compartments had already begun to recover. Three days after souring the

[1] It is reported that security guards are bribed to open the gates to contractors who engage in septic tank emptying. They then drive their vacuum tankers to the head of works and discharge the contents into the influent channels before the coarse screens. The only evidence of these activities is sludge splash marks in the influent channels. The Works manager did not appear to be concerned by the practice since it did not appear to have any significant effect on the operation of the plant due to dilution in the mixed activated sludge units. When plant workers were questioned retrospectively by one of the project team, they were uncertain but thought that the sludge splash marks may have been present on the day that reactor souring was observed, indicating that illegal dumping may have occurred the previous night.

reactor had essentially recovered, although low (<6) pH values were still observed in the later compartments. Ten days after souring, complete recovery was observed.

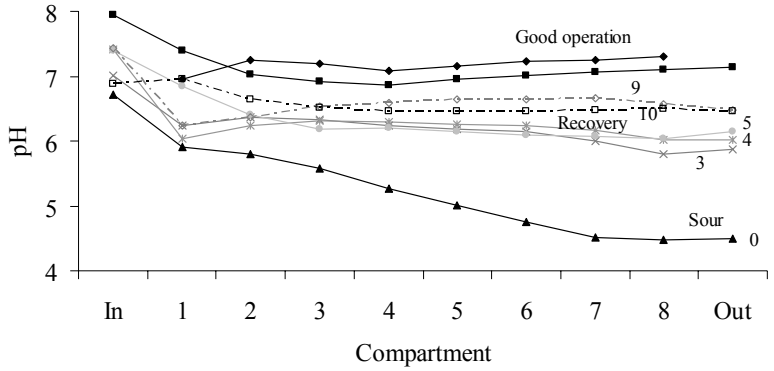


Figure 5.24: pH profiles in the ABR compartments showing good operation (♦ and■), pH profile shortly after souring (▲), labelled 0, and then profiles 3 days after souring (×), 4 days(☼), 5 days(●), 9 days (◊) and 10 days (□) during Phase II.

Operational problems in the recovery period resulted in little flow during this time, which will have accelerated recovery. However, the immediate increase in pH value in the early compartments on the day of souring (with normal flow) implies that rapid recovery is also possible under continuous flow conditions. It is hypothesised that the mechanism of the rapid recovery observed was based on the pseudo-plug-flow nature of the baffled reactor. Acid residues and untreated organics originating from the shock load were washed out of each compartment at a much faster rate than would be the case in a mixed system while sufficient biomass for further recovery was retained. Therefore, rapid recovery was seen to occur sequentially in each compartment. This is clear in **Figure 5.24** where considerable recovery is seen only a few hours after the event in the early compartments.

5.7 MICROBIOLOGICAL STUDIES OF THE PILOT-SCALE ABR

In this section, various microbiological studies that were undertaken on the pilot-scale ABR are presented. This work was achieved through collaboration with the School of Biological and Conservation Sciences at UKZN and the Department of Biotechnology, Durban Institute of Technology.

5.7.1 Pathogen indicator organisms

A portion of the MSc research undertaken by Pillay (2006) investigated the fate of pathogen indicator organisms in the pilot-scale ABR during Phase II. Summary results are presented **Table 5.5**. Detailed results of this study are presented in **Appendix A3.3.2** and are summarised here.

The removal efficiency for of the pathogen indicator organisms where all found to be significantly greater than zero (Student's t-test, $P < 10^{-3}$). From all indicator organisms tested, the greatest reductions were observed for Ascaris eggs. The greater removal helminth eggs is probably due to eggs having a larger mean residence time within the reactor due to sedimentation.

Although the reduction of the various indicator organisms was significant, none of the microbial and parasitic parameters met the requirements for discharge, either to water resources or for irrigation purposes in agriculture. It is therefore likely that the outflow may harbour a wide range of microbial

pathogens and parasites, which may present a potential health risk to humans and water supplies. These results indicate that outflow produced by a baffled septic tank or ABR would not be safe for discharge to water course or for agricultural use without a post-treatment step for the removal of pathogenic contaminants.

Table 5.5: Phase III: Pathogen indicator organism concentrations in inflow and outflow from the pilot-scale ABR

		Average/ Median	Std Deviation	Number of observations	Min.	Max.
Total Coliforms	In	7.3		25	7.0	7.7
[log(cfu/100mℓ)]	Out	6.6		25	5.8	7.1
E. Coli	In	7.7		25	7.2	8.1
[log(cfu/100mℓ)]	Out	6.8		25	5.9	7.3
Coliphage	In	4.1		24	3.6	4.8
[log(pfu/100mℓ)]	Out	3.5		24	2.0	4.2
***Ascaris* spp.**	In	772	341	13	347	1 500
[Number eggs/ℓ]	Out	17	15	13	2	56
Mean HRT: 22 h				**Total flow treated: 352 658 ℓ**		

5.7.2 Microbial community studies

Two Master level student projects were undertaken to investigate microbial community structure and dynamics on the pilot-scale ABR.

- During Phase III, Lalbahadur undertook a MTech research project through Durban Institute of Technology (DIT, now Durban University of Technology, DUT) looking at quantifying different classes and genera of micro-organisms in sludge taken from each of the compartments using a variety of microbial techniques (Lalbahadur, 2005).

- In Phase III and Phase IV, Pillay investigated microbial community structure by examining dispersed and granular sludge taken from different compartments by scanning electron microscopy (SEM) and epi-fluorescence microscopy (EFM) as part of his Master of Science dissertation through the School of Conservation and Biological Sciences at the University of KwaZulu-Natal. (Pillay, 2006)

The main findings of these projects are summarised in this section.

5.7.2.1 Characterisation of microbial communities using molecular techniques

Three molecular techniques were used for the identification and enumeration of microbial consortia in the samples. The details of the methods for these techniques may be found in Lalbahadur (2005). The three methods were

- total cell counts using 4'6-diamidino-2-phenylindole (DAPI) staining: binds with intact DNA, thereby providing an indication of the density of intact micro-organisms (although this technique does not indicate the activity of these organisms).

- fluorescent in situ hybridisation (FISH): binds with specific sites on ribosomal RNA in target cells. This technique allows identification and enumeration of live micro-organisms at a class, family or genus level.

- Polymerase Chain Reaction (PCR) technique: provides a qualitative positive identification of the presence of micro-organisms at a class, family, genus or species level.

The results of this study were reanalysed in detail in Foxon et al. (2006) significantly extending the conclusions that were presented in the original thesis by Lalbahadur (2005). The findings are summarised as follows:

- It was found that the full range of micro-organisms that effect anaerobic digestion were present in the pilot-scale ABR in Phase III.

- No spatial separation of micro-organisms with specific functionality (e.g. hydrolytic, acidogenic, acetogenic, methanogenic) was observed in the samples studied.

- The presence of hydrolytic and acidogenic bacteria throughout the reactor indicates that hydrolysable material was present in all compartments, i.e. that initial breakdown of particulate and polymeric material was the rate-limiting steps in digestion of complex particulate wastewater.

- Surprisingly low numbers of Archaea, particularly acetoclastic methanogens, were obtained by FISH throughout the compartments of the pilot-scale ABR; although Archaea were found in all compartments in all samples analysed, no *Methanosaeta spp*. were detected by FISH at all and *Methanosarcina spp.* were only found in compartments 1 to 4 in significant amounts at the beginning of Phase III (days 36 & 57), and in compartments 1 and 2 later in Phase III (Days 85, 101 & 127). DNA sequencing confirmed that *Methanosarcina spp.* were present but did not identify any DNA from *Methanosaeta spp.* Furthermore, *Methanobacterium* and *Methanococcus spp.* (probably hydrogenotrophic methanogens) were identified by DNA sequencing, although these had not specifically been probed for in the FISH study.

- The ratio of Eubacteria to Archaeal spp. (indicative of the ratio of hydrolytic and acidogenic bacteria to methanogenic Archaea) was large in all compartments (**Figure 5.25**) with significant microbial diversity being observed among Eubacterial genera in all samples. There did not appear to be any overall changes in the ratio of Eubacterial spp. to Archaeal spp. in different compartments. Thus this study did not provide any additional evidence to support the hypothesis that significant phase separation occurred in the pilot-scale ABR.

- No conclusions about the relative numbers of micro-organisms in each compartment could be satisfactorily drawn since details of the methods used to concentrate samples were not adequately recorded.

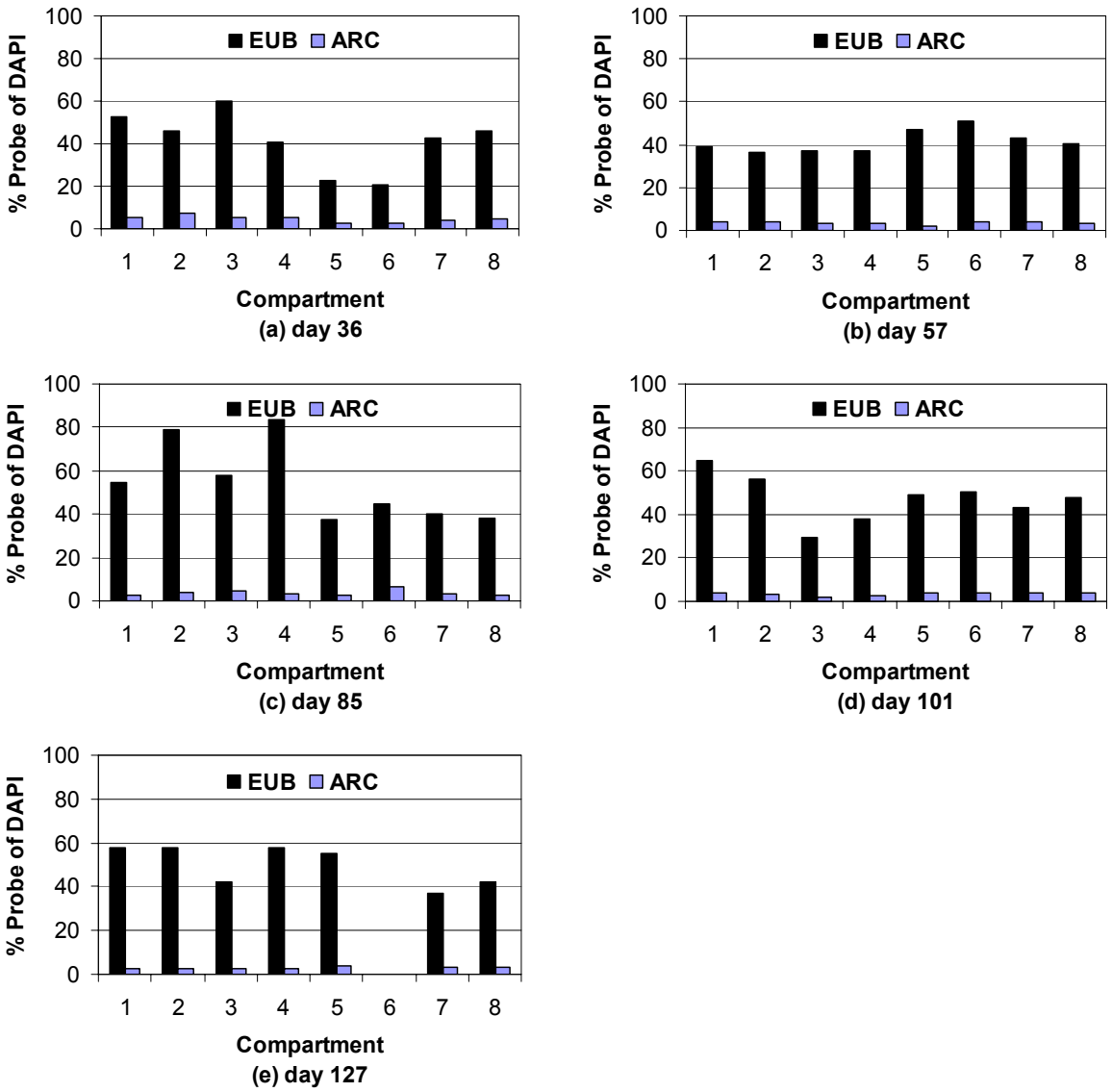


Figure 5.25: Domain-specific probe counts (Eubacteria and Archaea) as a fraction of total cell counts in each compartment for samples obtained on day 36, 57, 85, 101 and 127 respectively during Phase III. The sample from compartment 6 on day 127 was lost. (Reproduced from Foxon et al., 2006)

5.7.2.2 Microbial community structure characterisation

During Phase III and Phase IV, samples of dispersed anaerobic sludge, and granules from each of the compartments of the pilot-scale ABR were analysed using *scanning electron microscopy* (SEM). Individual granules were studied using *epi-fluorescent microscopy* (EFM) in Phase III only. Details of the analytical methods and results of this study may be found in Pillay (2006). Pillay (2006) cautioned that while SEM allows scientists to tentatively identify micro-organisms in environmental samples from their size and morphology, positive identification cannot be made without the assistance of direct molecular techniques of identification. However, certain of the methanogenic micro-organisms have characteristic morphologies that have been well documented, and can therefore be identified with some confidence.

In Phase III, the pilot-scale ABR was operated at an A-HRT of 22 h. Samples of sludge from each compartment were obtained using the coring technique described in **Section** 3.2.1.3. Samples of dispersed sludge and individually identified and isolated granules were prepared for examination by SEM. Pillay (2006) found that the diversity of microbial species of both dispersed sludge and granules was somewhat less than expected for a sludge treating a complex wastewater. Specifically, *Methanosaeta spp.* (characteristically rod shaped acetoclastic methanogens, which are thought to play an important role in the development of anaerobic granules, **Section** 2.1.7.3) were only infrequently observed in a few of the granules harvested from compartments 2 and 3, and not at all in dispersed sludge.

A few small granules were only found in compartments 2 and 3 in Phase III. Granules were typically small (diameter < 2 mm), were found to have a loosely packed interior, full of cavities, and were brittle with a tendency to crumble when handled (**Figure 5.26**). Examination of the surface and interior of the granules identified only two morphotypes, thought to be either *Methanococcus spp*. or *Methanosarcina spp*., and *Methanospirillum spp*. Observation of granules under epi-fluorescent excitation failed to identify any *Methanosaeta spp*. at all. These results supported those of Lalbahadur (2005) who had also failed to observe any *Methanosaeta spp*. in any samples of sludge from any of the compartments using molecular techniques. Furthermore, no stratification of layers within the granules was seen in these samples, as has often been reported for well-developed granules.

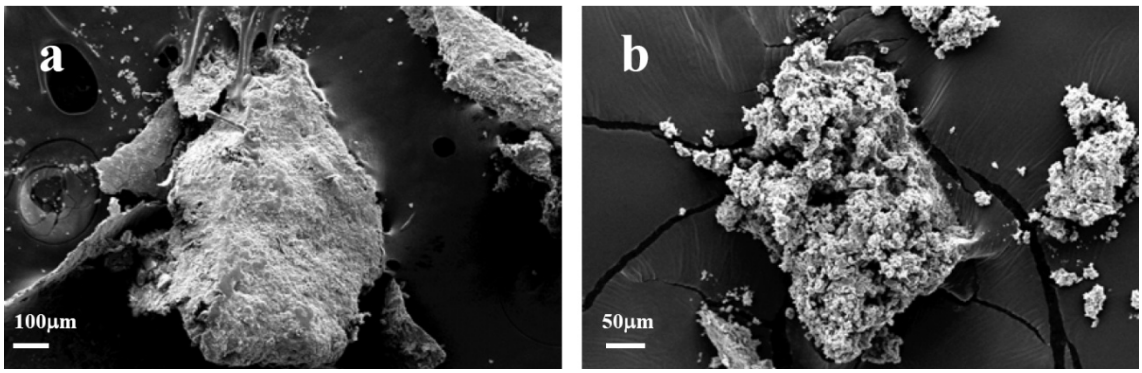


Figure 5.26: Scanning electron micrographs showing (a) the surface topology of an entire granule taken from compartment 3 of the pilot-scale ABR during Phase III; and (b) Cleaved granule showing loosely packed, cavity filled structure of the granule interior. (Reproduced with permission from Pillay, 2006)

It was concluded that the small range of methanogenic micro-organisms observed in these samples was due to limited microbial diversity caused by *selection pressure*. The concept of selection pressure describes the selection of certain micro-organisms in a culture through environmental conditions that preclude the establishment of other micro-organisms, as a result of the latter not being able to grow at a rate faster than their washout rate. Pillay was able to convincingly show that at the relatively low pH values observed in the pilot-scale ABR and at the relatively high upflow velocities applied in Phase III, a number of factors could have resulted in the failure of *Methanosaeta spp*. in particular to establish in the reactor. This was concluded to have been a significant cause of the little and poor granulation in Phase III (Pillay, 2006).

Finally, Pillay (2006) reported that the predominance of acetoclastic methanogens was not greater in later compartments than in earlier compartments, as would be expected in a system with complete phase separation (**Section 2.5.1.4**); The acetoclastic methanogenic genus *Methanosarcina* was tentatively identified in samples in compartments 1, 2, 3, and 5 and was not identified in samples from

later compartments at all. Furthermore, it was recorded that the numbers and diversity of different types of micro-organisms was greatest in compartment 1, and that the numbers and diversity decreased in later compartments. However, it should be stressed that this does not necessarily mean that no phase separation occurred since no identification or quantification of eubacterial species was made, and relative numbers of different types of micro-organisms were not determined in this study; thus the ratio of hydrolytic and acidogenic bacteria to methanogens could not be determined.

In Phase IV the A-HRT was increased to between 40 and 44 h, i.e. the flow rate was reduced. Samples of dispersed sludge and granules from each compartment were examined by SEM to investigate changes to the microbial community structure brought about by the change in flow conditions.

The first observation recorded was an increase in microbial diversity with many more different types of morphology observed in all samples. Significantly, bamboo-shaped rods and filaments characteristic of the morphology of *Methanosaeta spp.* were observed in all compartments, except the first with prevalence apparently decreasing in later compartments (**Figure 5.27**)

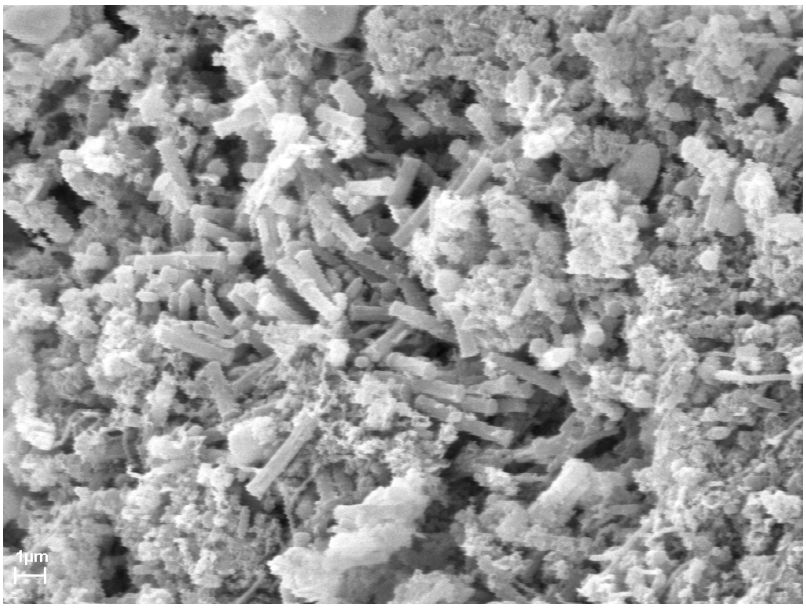


Figure 5.27: Bamboo-shaped bacteria that closely resemble the acetoclastic methanogen, *Methanosaeta* observed in dispersed sludge during Phase IV (Reproduced with permission from Pillay, 2006)

Granulation was observed to have occurred in compartments 2 and 3 and to a lesser extent in compartment 4. Granules observed in Phase IV were distinctly different in appearance to those studied in Phase III. Granules were more uniform in shape with spherical or oval shape and a much smoother surface than those from Phase III (**Figure** 5.28). In Phase IV, granules had distinct two-layered structure with a thin outer surface layer and a large interior core. The microbial diversity of the outer surface was greater than observed in Phase III with tentative identification based on morphology suggesting the presence of acidogenic bacteria, hydrogenotrophic and acetoclastic methanogens on the surface of the granule.

The granule interior was found to be made up largely of *Methanosaeta*-like cells embedded within a complex matrix of what appeared to be extracellular polymer (**Figure** 5.29). The interior of the granules also had a system of well defined cavities (**Figure** 5.29), often surrounded by cells whose

morphology suggested that they may have been acidogenic bacteria. Similar bacteria were observed in layers around aggregates of polymer-bound *Methanosaeta*-like cells (not shown).

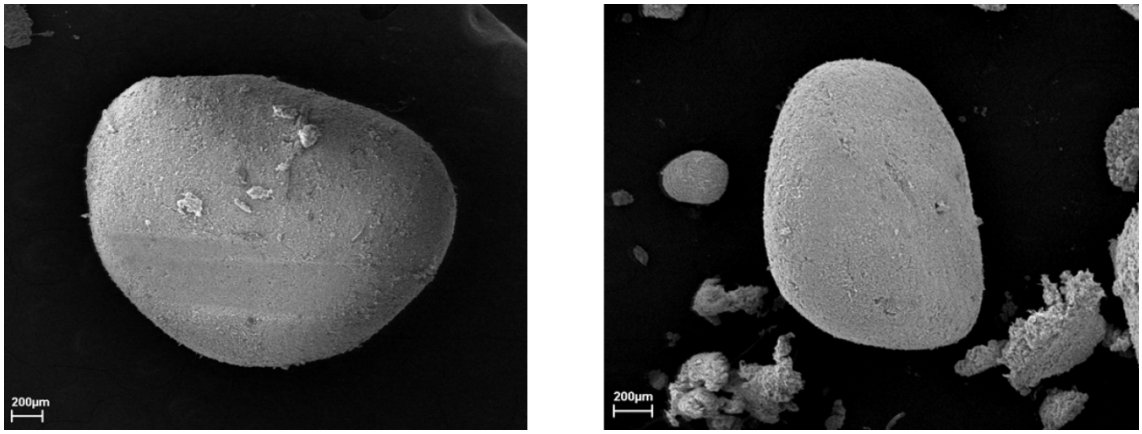
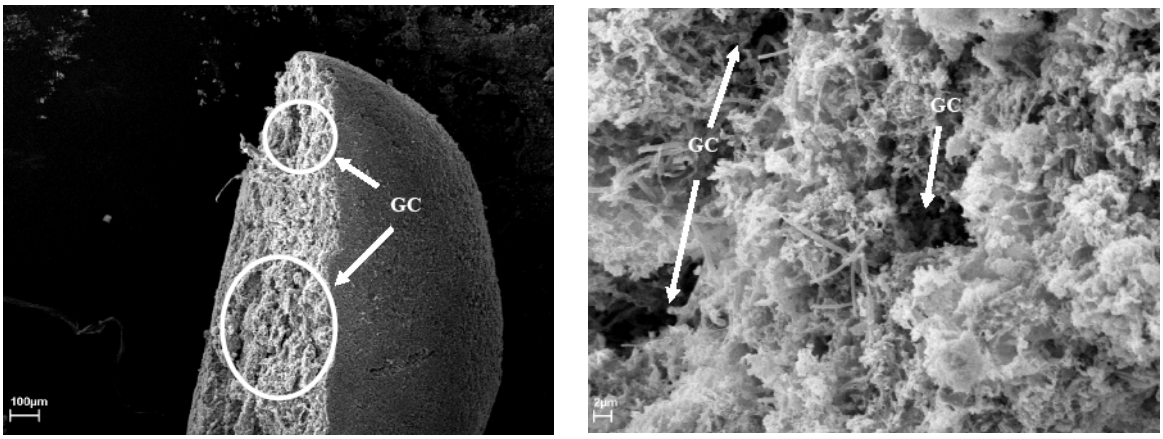


Figure 5.28: Scanning electron micrographs of granules from compartments 2 and 3 harvested from the pilot-scale ABR during Phase IV (Reproduced with permission from Pillay, 2006)



(a) (b)

Figure 5.29: Scanning electron micrographs of (a) a dissected granule showing cavities (GC) in the granule interior and (b) interior matrix composed of rod and filamentous forms of Methanosaeta-like morphotypes in a matrix of extracellular polymer. (Reproduced with permission from Pillay, 2006)

5.7.2.3 Conclusions of microbial community studies

In Phase III, two independent studies on microbial community structure showed that acetoclastic methanogenic populations in the pilot-scale ABR at an A-HRT of 22 h were not well established. This was most clearly shown in the failure to systematically observe *Methanosaeta spp.* in either dispersed sludge or within granules by either SEM or direct molecular techniques. Furthermore, granules were small, few and had a brittle, unstratified structure with a tendency to crumble.

In Phase IV, at a flow rate approximately half that of the previous phase, granulation was observed with much larger, two-layered granules that were found to have large populations of Methanosaeta-like micro-organisms present in the granule interior. Much greater microbial diversity was observed than in Phase III. A significant observation was the fact that Methanosaeta-like micro-organisms and granules were not observed in the first compartment at all, but both were found in compartments 2, 3

and 4, with decreasing observations of these micro-organisms in dispersed sludge of later compartments. This implies that partial phase separation has occurred between the first and subsequent compartments with hydrolysis and acidogenesis predominating in the first compartment and establishment of acetoclastic methanogenesis thereafter.

These results confirm the hypotheses presented in **Sections** 5.6.1.4 and 5.6.2 that proposed that high solids accumulation rates (normalised for OLR) in Phase III compared to those in Phase IV were due to the establishment of a stable anaerobic population at the lower flow rate of Phase IV. As the concentrations of soluble components in the pilot-scale ABR are relatively low during the treatment of domestic sewage, the cause of the differences in population stability between the different flow rates must be due to the higher upflow velocities employed in Phase III resulting in the washout of anaerobic micro-organisms. Thus it may be concluded that the upflow velocity in each compartment is a critical design factor that must be considered in conjunction with A-HRT and OLR when sizing a baffled reactor for the treatment of domestic sewage.

5.8 SUMMARY OF OPERATION AT KINGSBURGH WASTEWATER TREATMENT PLANT

Table 5.5 and **Table 5.6** (pages 117 and 127) are summaries of all inflow and outflow measurements averaged for operation of the pilot-scale ABR at Kingsburgh WWTP for Phase II, III and IV. In the sections that follow, the main conclusions drawn from the data presented in this chapter are summarised.

5.8.1 Outflow characteristics

Substantial reductions in COD were observed in all operating periods, with lowest values (around 130 mgCOD/ℓ) measured in Phase IV. A number of factors contributed to the best performance being observed in Phase IV:

- The ABR was operated at low flow rates, resulting in low upflow velocity and therefore good solids retention characteristics and development of a balanced anaerobic population to facilitate digestion.
- The low feed rate in Phase IV corresponded to a low OLR.
- Significant improvements in management and control of the pilot-scale ABR resulted in fewer sludge loss incidents

Anaerobic treatment of domestic wastewater in the ABR caused a net increase in alkalinity and free and saline ammonia concentrations, and a slight decrease in TKN. A small amount of sulphate in the feed stream was removed by the ABR. Nitrate present in the wastewater was expected to be completely removed within the first one or two compartments of the system. Phosphate in the inflow stream was not expected to change as a result of treatment in the ABR. The ABR outlet stream therefore contained increased concentrations of alkalinity and free and saline ammonia and similar concentrations of phosphate compared to the influent wastewater. No sulphate or nitrate was detected in the outlet stream.

Significant removal of pathogen indicator organisms was observed in Phase IV. However, outflow coliform, coliphage and *Ascaris* spp. concentrations in the outlet stream were sufficiently high that the effluent from an ABR treating domestic wastewater should be considered a risk to human health.

5.8.2 Diurnal variation in inflow characteristics

Measurements of inflow and outflow COD and alkalinity concentrations and pH values made hourly for 24 h showed that there was significant variation in COD and alkalinity concentration, with higher values for both determinands observed during the day time and lower values at night.

The normal times at which regular samples were taken (i.e. between 08h00 and 13h00) corresponded to higher COD concentrations in the ABR feed stream than the average value for the 24 h period. Since the pilot-scale ABR was fed at a relatively constant rate (i.e. with no diurnal variation in hydraulic load), the average OLR actually applied to the system may have been lower than predicted from the regular samples obtained.

Although alkalinity concentration also showed some diurnal oscillation, there was no apparent difference between measurements recorded during the 24 h campaign and during regular sampling.

Examination of inflow pH values indicated that a value of 6.8 might more accurately describe the feed condition than the median value of 7.0 observed in regular sampling data.

Comparison of inflow and outflow alkalinity data led to the tentative estimation that the hydraulic retention time experienced by the bulk of the fluid flow was approximately 16 h at an A-HRT of 22 h. This corresponds to a value of 27% of the reactor that is not available, or not readily available (as a result of the presence of sludge) for liquid flow.

5.8.3 Sludge dynamics within compartments

As was observed for operation while installed at Umbilo WWTP, interpretation of data for determinands in compartments and in the pilot-scale ABR outflow hinges on understanding of sludge dynamics in each compartment. **Section** 5.6.1 showed that the average amount of sludge in upflow compartments of the pilot-scale ABR increased during Phases III and IV, periods characterised by good control of the feeding system. In contrast, no significant increase in amount of sludge was reported for Phase II, and this was attributed to poor feed system control with regular periods of down-time and sludge washout[1]. These results confirm that when solids are not being subjected to washout incidents, increases in the amount of sludge present in compartments can be expected; i.e. solids are not removed by the biological or physical processes within the ABR at the same rate at which they are being added, indicating that inert particulates and possibly also biodegradable particulates accumulate with time. Thus it is inevitable that the ABR will ultimately require desludging to remove accumulated solids.

[1] Note that this is not the same as saying the design is not stable to hydraulic shocks, since the conditions prevailing during washout incidents were more like flood conditions than high hydraulic load conditions, and would presumably be eliminated by appropriate sizing and design of a full-scale system.

- The mechanism of sludge build-up within a compartment has been shown to be a combination of growth of micro-organisms on biodegradable material within the compartment, and sludge carry-over from the previous compartments. In **Section 5.6.1.5**, it is seen that the sludge bed in a compartment can build up until the entire compartment is full. In the event of this occurring in the last compartment, considerable carry-out of sludge will occur, and desludging in all compartments, but especially the last, may be required. However, this did not occur in the 5 years of operation of the pilot plant.

- The rate at which solids accumulate (considering only the upflow side of each compartment was found to be 2.1 (ℓ settled sludge)/(kg COD applied) or 0.14 (kg dry solids)/(kg COD applied) at an A-HRT between 40 and 44 h, while this increased to 4.31 (ℓ settled sludge)/(kg COD applied) at a lower A-HRT of 22 h.

- Two independent microbial community analysis studies indicated that few Archaeal genera were well established during Phase III. Poor granulation was observed and low microbial diversity, but most importantly, few acetoclastic methanogens were observed in this phase with A-HRT of 22 h.

- In comparison, in Phase IV many granules were harvested and found to be much larger than the few observed in Phase III, with a two-layered structure that incorporated significant amounts of the acetoclastic methanogens *Methanosaeta spp.*. It was concluded that the lower upflow velocities applied in Phase IV allowed the establishment of stable methanogenic populations and resulted in overall better anaerobic digestion of the feed wastewater.

- These results suggest that the upflow velocity is the limiting factor that determines the organic and hydraulic loading rates that may be applied in a baffled reactor design. Upflow velocity appears to control the specific sludge accumulation rate, and thus, ultimately the required desludging interval for any particular system.

5.8.4 Soluble COD and pH dynamics within compartments

In **Section** 5.6.2, analysis of the shape of the pH value profile in each operating period was used to understand the relationship between acidogenic and methanogenic processes in the ABR.

Anaerobic digestion of domestic wastewater in the ABR occurred with little *reserve alkalinity*, causing operating pH values to regularly drop below 6.5. Significant inhibition of methanogenesis would therefore have occurred. Since hydrolysis was been identified as the rate limiting process in all but the first compartment, it was not expected that methanogenesis inhibition would have reduced the overall COD removal. However, inhibition of methanogenic micro-organisms by low pH values may have inhibited Archaeal growth rates and therefore increased the risk of methanogen washout. This would have compromised the ability of the ABR to withstand, and recover from, shock loads, either hydraulic or organic, and therefore lessens the advantage of installing an ABR over simpler technology, such as a septic tank.

The pH value profile for Phase IV surprisingly showed lowest pH values in compartments 2 to 8 when compared to the more highly loaded Phases II and III. It was concluded that the establishment of more stable anaerobic consortia at the lower upflow velocity of Phase IV meant that greater anaerobic activity was occurring throughout the pilot-scale ABR in Phase IV resulting in significant acidogenesis throughout the reactor, and therefore lower pH values throughout.

Table 5.6: Inflow and outflow characteristics, Phase II, III and IV. Data are presented as mean value ± 95 % conf. Interval, [min, max] (number of observations)

		In[1]	Out Phase II	Out Phase III	Out Phase IV
COD	mgCOD/ℓ	**680 ± 25** [246, 1749] (258)	**299 ± 57** [125, 674] (20)	**212 ± 37** [107, 1 202] (57)	**130 ± 29** [63, 339] (18)
Soluble COD	mgCOD/ℓ	**154 ± 40** [69, 395] (20)	**204 ± 37** [132, 298] (8)	**71 ± 8** [27, 121] (26)	**61 ± 50**[2] [18, 427] (19)
Alkalinity	mg$CaCO_3$/ℓ	**243 ± 6** [43, 369] (269)	**226 ± 35** [59, 353] (15)	**268 ± 20** [185, 316] (13)	**246 ±52** [168, 286] (4)
Free and saline ammonia	mgN/ℓ	**39 ± 2** [3.8, 204] (273)	n.d.[3]	**34 ± 2** [30, 39] (7)	**51 ± 14** [20, 90] (10)
Total Kjeldahl nitrogen	mgN/ℓ	**45 ± 2** [40, 51] (8)	n.d.	n.d.	**38 ± 3** [32, 45] (8)
Soluble phosphate	mgP/ℓ	**15 ± 0.8** [0.95, 64] (245)	n.d.	**5.6 ± 0.5** [4.7, 6] (5)	**22 ± 4** [10, 26] (7)
Total solids	mgTS/ℓ	**673 ± 66** [253, 1 076] (43)	**450 ± 52** [310, 675] (15)	**225 ± 50** [80, 390] (14)	**378 ± 39** [135, 556] (26)
Volatile solids	mgVS/ℓ	**417 ± 66** [125, 705] (25)	**268 ± 102** [118, 605] (11)	**127 ± 42** [5, 290] (14)	n.d.
pH	(median value reported)	**7.1** [4, 7.94] (289)	**6.2** [4.5, 7.1] (7)	**6.5** [6.2, 6.7] (9)	**6.5** [6.2, 7.4] (6)
VFA	mgCOD/ℓ	**30 ± 31** [0, 74] (4)	n.d.	n.d.	**0** [0, 0] (4)
Sulphate	mgSO_4/ℓ	**4.5 ± 1.1** [2.4, 5.8] (5)	n.d.	n.d.	**0.42 ± 0.28** [0.20, 0.97] (5)
Sodium	mgNa/ℓ	**150 ± 104** [87, 362] (5)	n.d.	n.d.	**132 ± 123** [40, 380] (5)
Potassium	mgK/ℓ	**21 ± 3** [16, 26] (6)	n.d.	n.d.	**25 ± 4** [19, 31] (6)

[1] Overall average conditions are reported for the inlet stream (i.e. all inflow data Phase II, Phase III and Phase IV are considered in the reported descriptive statistics.

[2] Note: these data are not normally distributed as a result of a few high outlet values.

[3] Not determined

6 DISCUSSION: IMPLICATIONS FOR DESIGN AND OPERATION

"Reports that say that something hasn't happened are always interesting to me, because as we know, there are known knowns; there are things we know we know. We also know there are known unknowns; that is to say we know there are some things we do not know. But there are also unknown unknowns - the ones we don't know we don't know" – Donald Rumsfeld

The purpose of this study was to investigate the performance of the pilot-scale ABR to (1) understand the mechanisms of treatment of domestic wastewater; (2) identify potential advantages and disadvantages of the reactor design for this application; (3) identify critical parameters for effective design of an ABR for domestic wastewater treatment; and (4) to develop a dynamic biochemical model of the pilot ABR to assist with design. It was hypothesised that the critical design parameter was A-HRT and that phase separation would ensure stability of operation; consequently the pilot-scale ABR was designed to have variable feed flow rates.

This chapter considers the experimental results in terms of the four objectives of this study.

6.1 PILOT-SCALE ABR STUDY – IMPLICATIONS FOR DESIGN

Apart from generally presenting and interpreting data describing performance of the pilot-scale ABR during treatment of domestic wastewater at two municipal WWTPs, this study has in particular developed an understanding of specific mechanisms of treatment that control start-up, sludge build-up, solids retention and microbial community dynamics. In this section, three issues that affect design and applicability of the technology, vis.

- pH buffering and phase separation
- Sludge accumulation and microbial community dynamics
- Pathogen removal and effluent characteristics

Analysis of these topics fulfils the requirement of the first objective of the study i.e. to understand the mechanisms of treatment in an ABR treating domestic sewerage.

6.1.1 pH buffering and phase separation

Two important and interrelated measures of the condition of anaerobic digestion are the pH value and the alkalinity concentration. The former affects microbial metabolic rates and the speciation chemistry of several liquid-phase components and the latter affects the system's ability to resist change in pH value.

6.1.1.1 pH value

An important observation was that for most of the experimental study, the pH value measured in the outlet stream was less than the pH value of the inlet stream of the pilot-scale ABR. This was initially surprising since in many applications, anaerobic degradation of a waste stream results in the production of significant amounts of alkalinity and free and saline ammonia, and a net increase of pH is observed. However, examination of the literature showed that the effect was not unprecedented;

Behling et al. (1997) studied the performance of a UASB treating domestic wastewater and also observed a net decrease in pH value although stable digestion conditions were maintained. These authors concluded that pH and alkalinity measurements are not as important as the VFA:alkalinity ratio and showed that for their system, this ratio was less than the critical value of 0.4. It was proposed that the expected pH value that should be achieved through digestion of domestic wastewater be investigated and this is dealt with in **Section** 6.4.

The pH values recorded in the compartments of the pilot-scale ABR were low, usually below 7.0. Many anaerobic processes experience inhibition when the pH values decrease below 7.0 and it was probable that biochemical conversion rates were low in the pilot-scale ABR study due to pH inhibition effects.

6.1.1.2 Alkalinity

The alkalinity concentrations measured within the reactor and in the outlet stream were substantially lower than values conventionally observed in anaerobic digestion processes. However, these values were expected since the dilute feedstock was known to have a low initial alkalinity concentration and a low alkalinity generation potential. The problem of low alkalinity in the feed wastewater was exacerbated by the fact that eThekwini Municipality is known to have low alkalinity water sources (**Section** 3.1.3)

As a result, the pH buffering capacity in the ABR was not large, and therefore the ability to withstand hydraulic and organic loads was compromised. This was evident during an incident in Phase II when dramatic and sudden depression of pH values was observed throughout the pilot-scale ABR, allegedly as a result of a single high organic load incident.

One approach to handling low alkalinity conditions is through alkalinity dosing e.g. with lime. In this way, operators of the system can be assured that there is sufficient pH buffering capacity since it is possible to dose lime or carbonate salts in excess of the strict requirements of the system (Speece, 1996). There are however two disadvantages to this approach: firstly alkalinity supplementation constitutes an additional cost to system maintenance, and moreover requires daily maintenance by a dedicated operator. In many applications these may not be available. Secondly, there may be consequences of excess supplementation such as precipitation of calcium carbonate.

A second approach is to continue to operate at low pH values with low pH buffering capacity, since this study has shown that stable digestion can be achieved under these conditions provided the hydraulic and organic loads are sufficiently low. Speece (1996) indicates that this is a practical approach to reduce alkalinity requirements, but requires monitoring and control instrumentation to prevent process upsets, and is characterised by microbial inhibition and sub-optimal treatment rates.

Depending on the location and hydrogeology of a system, and whether or not there are pH and load sensitive post-treatment steps, a certain amount of process instability may be tolerable: in **Section** 5.6.3 it was concluded that rapid recovery from a process upset appeared to occur without any operator intervention since the pseudo-plug-flow design of the system resulted in sour liquors washing out of the ABR without excessive loss of sludge. Therefore, if the receiving environment is able to tolerate small amounts of sour effluent very occasionally, then provided due care is taken in the design to reduce the probability of this occurring, it should be possible to operate without any alkalinity supplementation. It is noted that most of the BORDA DeWaTS systems (**Section** 2.5.2.3) operate without any alkalinity supplementation.

Nevertheless, **Section** 6.4 provides an analysis of alkalinity supplementation requirements for dilute domestic wastewater.

The low alkalinity in eThekwini waters relative to the rest of South Africa mean that these results and conclusions apply to a poor case (if not worst case) scenario, since in many other regions, the alkalinity concentration in the inflow will be substantially higher than observed in this study. Under these conditions, higher pH buffering capacity and therefore higher pH values may be expected, with improved degradations rates and therefore higher biomass growth rates.

6.1.1.3 Phase Separation

One of the claimed advantages of the baffled design of an ABR is the spatial separation of acidogenic and methanogenic stages in different compartments of the ABR (**Section** 2.5.1.4); it was hypothesised that a baffled reactor design would be more beneficial in sewage treatment than a single stage system because of this effect. However, in the general anaerobic digestion literature, several authors suggest that the advantages of phase separation in the digestion of particulate material are uncertain (Hanaki et al., 1987; Leitão et al., 2006). This point was also made by Barber and Stuckey in their review of ABR applications with reference to an unpublished study by Hassouna and Stuckey.

In the present study lower pH values were seen in the first compartment during certain phases of operation, and in Phase IV, when stable anaerobic digestion was known to have established, acetoclastic methanogen species *Methanosaeta* was observed in compartments 2, 3 and 4, but not in the first compartment. These results indicate that partial phase separation did occur in that methanogenesis was limited in the first compartment. However, hydrolytic and acidogenic processes were not limited to the first compartment; hence true phase separation, i.e. where methanogenesis is negligible in the first phase, and dominant in the second or later stages (Leitão et al., 2006) did not occur (**Section** 5.8.4).

From the results of the pilot-scale ABR study, the following understanding of phase separation was developed for the ABR treating domestic wastewater:

- The extremely high load of biodegradable organic material accumulated in the first compartment led to the establishment of a large number of anaerobic micro-organisms which degrade the particulate organics. The high organic load and relatively low pH buffering capacity of the wastewater and the reactor liquors resulted in pH depression in compartment 1, where the rate of acidification exceeded the rate at which acids were removed through methanogenesis. Consequently VFA exited compartment 1 with other soluble components.

- The higher flow of VFA to compartment 2 but overall lower organic load in compartment 2 compared to compartment 1 resulted in the kinetic selection of different microbial communities in the two compartments. Acetoclastic methanogens made up a larger fraction of all active micro-organisms in compartment 2, while the fraction of hydrolytic and acidogenic bacteria were proportionately fewer than in compartment 1 (**Section** 5.7.2.2). However, the pH values were not always significantly higher in these compartments; therefore true phase separation had not occurred.

- Subsequent compartments had similar microbial consortia to one another since the biodegradable substrates in these compartments would be similar, i.e. particulate components washed through by pulsing flow or soluble components that had not made contact with micro-

organisms in previous compartments. Thus a certain amount of differentiation between compartment micro-organisms could be expected, although, as was observed, the most significant differences should have been between compartments 1 and 2.

In **Section** 4.4.3, it was further observed that pH depression extended to compartment 2 in the 20 h THRT period, but that the outlet stream pH values were not affected. This implies that the physical reactor design is particularly robust in terms of pH buffering since low pH values can extend to more than one compartment without adversely affecting overall reactor operation.

In conclusion, there was some evidence that partial phase separation had occurred between the first and subsequent compartments, but that this was not true phase separation since low pH values were observed at times in compartments subsequent to the first, and since hydrolysis continued outside of the first compartment. Therefore the original hypothesis that phase separation in an ABR treating sewage is a benefit of the design over a single phase system is not strictly supported; however, the baffled design provides other advantages as elaborated in 6.1.2. Furthermore, Since the ABR is a solids accumulating system, the spatial location of the acidogenic phase was not fixed to the first compartment. This implies that a reactor design specifying three or more compartments would be appropriate for this kind of application.

6.1.2 Sludge accumulation and microbial community dynamics

The key difference between the design of an ABR and many conventional anaerobic digesters such as some UASB and the expanded granular sludge bed reactors is that there is no provision for continuous sludge removal. Consequently, they accumulate solids, causing continuous change in sludge loads and associated microbial dynamics. Understanding of sludge accumulation mechanisms and rates is therefore critical for developing design and operation guidelines.

From the results of the pilot-scale ABR study, the following understanding of sludge accumulation was developed for the ABR treating domestic wastewater (**Figure 6.1**):

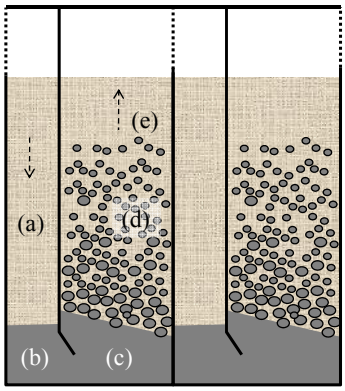


Figure 6.1: Location of solids in downflow (a and b) and upflow (c, d and e) sections of each compartment: (a) downflow clear zone; (b) compacted sludge bed; (c) settled sludge bed; (d) fluidised sludge bed; and (e) upflow clear zone.

The downflow side of each compartment consisted of (a) an upper *clear* zone derived from the overflow of the previous compartment and (b) a lower zone consisting of a compacted sludge bed. Each upflow section consisted of three parts analogous to a UASB (**Section** 2.2.1): (c) a settled sludge bed containing large anaerobic granules, dense debris and entrapped finer solids; (d) a fluidised sludge

blanket consisting of smaller granules, biomass flocs and dispersed sludge, and (e,) a *clear* zone consisting of liquid flow with entrained fine solid particles. Once the compartment became filled with sludge, the clear zone would be eliminated by the sludge blanket.

The sludge bed and fluidised sludge blanket increased the treatment potential of the system by (i) creating a filter through which fluid flow had to pass, causing entrapment of less dense particulate material and colloidal material and (ii) by creating a matrix in which anaerobic micro-organisms could establish, and through which flow had to pass, ensuring that good contact between biomass and degradable components of the feed occurred. Soluble components originating from the feed or generated through hydrolysis of particulate material passed through the sludge bed and fluidised sludge blanket with the liquid flow. Inert soluble components passed through the reactor and exited with the outflow from compartment 8. Degradable soluble components may have degraded *if* they came into contact with appropriate anaerobic micro-organisms. The probability of this occurring would have been highest in the sludge bed or blanket since this is where the greatest concentration of micro-organisms may be found. However, small anaerobic consortia or free suspended anaerobic micro-organisms may have been entrained in the liquid flow; thus digestion of soluble or colloidal components may also have occurred in the relatively clear liquid above the sludge bed/blanket.

As sludge beds developed in later compartments, the contact time between biodegradable material and micro-organisms increased since the probability that a biodegradable molecule would come into contact with appropriate anaerobic micro-organisms increases with increased amount of time spent in the sludge bed or blanket. The overflow from each compartment (including the last, and thus the outlet stream) therefore contained a combination of completely degraded material i.e. the distribution of degradability of material in the outlet of a compartment depended on the distribution of contact times of material in the compartment (**Section** 2.3.1.2).

Sludge retention has two components: firstly, retention of biodegradable particulate material increases the residence time of the material and therefore the probability that it will be degraded within its residence time in the reactor; Secondly anaerobic micro-organisms, the active agents of anaerobic digestion form part of the solid sludge, therefore retention of sludge implies retention of anaerobic micro-organisms and enhancement of anaerobic digestion.

Two critical questions arise from this analysis:

- What is the relationship between organic load, upflow velocity and sludge accumulation rate (specifically, micro-organism retention rate)? i.e. how do the upflow velocity and organic load affect the degree of solids retention, and the rate of microbial growth?

- What is the relationship between contact time and sludge accumulation? i.e. what is the probability that the actual contact time experienced by a particle of biodegradable organics at a particular OLR and with the prevailing sludge load exceeds the ultimate contact time required for complete degradation?

6.1.2.1 Upflow velocity vs. specific sludge accumulation rate

The relationship between the rate of solids carry-over from one compartment to the next, and upflow velocity was not well known. Technically this information could be inferred by comparing sludge loads per compartment (such as compartment total solids or compartment sludge height) to solids concentration in the liquid passing over the standing baffle of each compartment at different upflow

velocities. Unfortunately, these two measurements were not performed simultaneously during this study and thus no comparable data sets are available to inform a model of upflow velocity vs. sludge accumulation rate.

The analysis of the pilot-scale ABR data concluded that acetoclastic methanogens were not well established, and poor granule formation was observed (assumed to be due to a high washout rate of slower growing anaerobic micro-organisms) at the upflow velocities applied in Phase III; conversely, at the lower hydraulic and OLR of Phase IV, good granulation and acetoclastic methanogenesis were both observed. With the assistance of rigorous statistical tests, it was shown that the amount of sludge accumulating per kg COD applied was higher in Phase III than in Phase IV despite the higher upflow velocities in Phase III, and this was attributed to the much enhanced digestion rates in Phase IV due to the establishment of more stable and diverse microbial consortia.

These two operating periods provided only two points in a 4-dimensional space (organic loading, upflow velocity, inflow alkalinity concentration and specific sludge accumulation rate), although conveniently, one point (Phase III) fell in a region of hydraulic overload and the other (Phase IV) in a region of stable digestion.

The mean upflow velocities for these two phases were calculated as 0.55 m/h in Phase III and between 0.27 m/h and 0.30 m/h in Phase IV (**Table 5.2**). In their review of UASB processes, Lin and Yang (1991) state that the design of the settling region of a UASB reactor should not exceed a superficial velocity of 0.7 m/h; Sasse (1998) recommended a design upflow velocity not exceeding 2 m/h for an ABR treating domestic wastewater (**Section** 2.5.2.3) and Hulshoff Pol et al. (2004) indicated that the settling velocity of anaerobic granules was commonly in the region of 60 m/h i.e. it appears that the performance of the pilot-scale ABR was poor in terms of sludge retention compared to other anaerobic applications.

The maximum upflow velocity (the lowest upflow value at which acetoclastic methanogens fail to establish due to high selection pressure) is not a fixed value for all systems, but is related to growth rates of micro-organisms and settling properties of the sludge.

- At high growth rates, and particularly when the rate of growth of sludge (in an easily settlable form) is high, higher upflow velocities may be achieved.

- Conditions that cause low microbial growth rates such as low OLRs and low pH conditions (or high concentrations of toxicants) will result in wash-out of slowest growing microbes at lower upflow velocities than in some more highly loaded systems.

  It was noted that the local municipal region, eThekwini, experiences low alkalinity concentrations in water resources and in potable water (**Section** 3.1.3), and that this is a determining factor in the low reactor pH values observed in this study (**Section** 6.1.1.2). Thus low inflow alkalinity concentration resulted in inhibition of microbial growth in this study, and therefore a lower estimation of critical up-flow velocity than may have been inferred in a different region with higher alkalinity waters.

- At low biodegradable COD concentrations, especially low soluble COD concentrations, the development of granules is not favoured since the rate of diffusion of organic substrate into sludge granules is low due to small concentration gradients. The flow rate required to wash out

small granules or ungranulated sludge flocs would therefore also be higher, compounding the problem of low growth rates.

In this study, both low OLRs and low pH conditions can be expected to have negatively impacted on microbial growth rates and thus increased the susceptibility to wash-out of slow growing microbes.

It is concluded that the ability to treat domestic wastewater at higher OLRs than tested in this study will depend on designing the upflow section of each compartment such that the superficial upflow velocity is below the critical value at which slower growing micro-organisms will not establish. A value of 0.3 m/h for critical up-flow velocity is proposed at the design OLR for low alkalinity applications. Sufficient biomass retention may be achieved at up-flow velocities of 1 m/h or higher if there is no inhibition due to low pH values.

Low up-flow velocities for a fixed reactor volume may be achieved by using shallow reactors with large upflow area, or by having few compartments. Alternatively, additional measures to retain sludge may be considered such as the addition of packing media to reduce the rate of removal of micro-organisms through washout (Rajinikanth et al., 2008).

It should be noted that there is a difference between the effect of peak up-flow velocities and sustained up-flow velocities on biomass growth rates:

- Garuti (2004) concluded that short bursts of flow at high flow rates resulted in better overall sludge retention than longer periods of flow at a lower flow rate, (but overall equal average hydraulic load), since the maximum sludge bed expansion achieved during short bursts of flow was less than during sustained low flow periods. (**Section** 2.5.1.5). It follows that short periods of higher (peak) flow will not have the same effect on micro-organism washout as sustained high up-flow velocities since the sludge bed will not expand to the same extent.

- If the prevailing up-flow velocity is sufficiently low to allow establishment of slow-growing micro-organisms, then occasional slightly elevated flows will strip out a portion of these, but not all, unless all of the solids are washed out.

- If the prevailing up-flow velocity is sufficiently low to allow establishment of slow-growing micro-organisms, this study has shown that anaerobic granules will develop. These are known to be more resistant to washout at elevated up-flow velocities due to their better settling properties than dispersed sludge (Hulshoff Pol et al., 2004; Zhou et al., 2006).

Consequently, it is possible to operate an ABR at peak upflow velocities greater than 0.3 m/h, provided the lengths of the bursts of elevated upflow velocity are not too long. This study does not provide any indication of how long these bursts can be, or how much higher than the critical upflow velocity the peak upflow velocity should be.

6.1.2.2 Contact time vs. Sludge accumulation rate

In an ABR treating domestic wastewater, contact time refers to the amount of time a biodegradable component spends in direct contact with active anaerobic micro-organisms such that microbially-mediated reaction can occur. There is no direct method of measuring contact time (Nauman and Collinge, 1968b). However, it is assumed that contact time will be inversely related to flow rate and directly related to the proportion of the liquid flow path filled with active micro-organisms.

Thus for a fixed sludge load, at high flow rates, contact time will be low and therefore the relative extent of treatment will be low; or similarly, at fixed flow rates, contact time should increase with increasing sludge load. However, not all sludge is made up of active micro-organisms; it may be inferred from the high sludge accumulation rates in Phase III, but relatively low microbial diversity and relatively high outlet COD concentration, that the proportion of active micro-organisms in the sludge was lower than in Phase IV. Thus contact time and consequently extent of treatment will increase with increasing active micro-organism load, rather than with increasing sludge load.

This concept is illustrated by the fact that after start-up, outlet COD concentrations (**Figure 5.8**) did not decrease with time in Phase III and IV, despite the fact that overall sludge load did increase (**Figure 5.16** and **Figure 5.17**). This implies that a pseudo-steady-state exists where although the amount of solids in the reactor increases with time, the number and activity of micro-organisms do not. For this to be true, the bulk of the accumulated material must be inert or slowly biodegradable particulates.

However, increasing contact time (i.e. amount of time in contact with active micro-organisms) should result in reduced sludge accumulation rates as a result of greater extent of treatment being achieved. Thus for long contact times, the accumulated sludge contains a large fraction of inert material, while for shorter contact times, it contains a greater proportion of biodegradable material.

Finally, in an ABR, increased flow rates result in increased washout of slower-growing micro-organisms. Thus at increased flow rates, the micro-organism load may be low (as observed in Phase III) and thus the contact time is low as a result of both short retention times, and low micro-organism loads.

6.1.2.3 Critical design parameter for sludge digestion

It was hypothesised that the most important design parameter for managing effluent quality and retained sludge digestion rates was the applied hydraulic retention time (A-HRT) since this controls the amount of time that wastewater spends inside the digester. However, it has been proposed that the average contact time (defined as the amount of time that a package of fluid spends in contact with active micro-organisms during its residence time in the digester) controls effluent quality and digestion rates, and that this depends on concentration of active biomass in the sludge bed and amount of time spent in the sludge bed.

The first factor, the biomass concentration has been shown to be strongly dependent on the upflow velocity in compartments of the ABR since at higher upflow velocity, poor biomass retention was inferred, while at lower velocities, a stable, active granulating sludge developed.

The second factor, the amount of time spent in the sludge bed, depends on the volume of the sludge bed, and the residence time in the reactor. At higher flow rates (at fixed feed concentration), sludge bed volume increased more rapidly than at a lower flow rate, but lower active biomass concentrations meant that higher sludge volume did not equate to higher contact time.

This analysis indicates that although residence time in the reactor affects digestion rates and effluent quality, design should consider upflow velocity and residence time (A-HRT) as both are critical in the control of digestion rate and effluent quality. Furthermore, OLR and inflow alkalinity concentration will affect the critical upflow velocity that determines whether slow-growing micro-organisms can establish, or are washed out.

6.1.3 Pathogen removal and effluent characteristics

The first objective of this work was to determine whether the ABR design had application in the treatment of domestic wastewater. To achieve this, it is necessary to define the performance criteria and then to compare the system performance to these criteria.

It was understood that anaerobic technology alone could not treat any domestic wastewater stream to meet the South African General or Special limits for discharge to water resources (DWAF, 1999) (**Appendix** A1) since anaerobic digestion makes no provision for the removal of free and saline ammonia and phosphorus, and in fact increases the free and saline ammonia concentration through liberation of organically bound nitrogen. Therefore, except when a treatment system is designed for a specific application where treated wastewater will be used for irrigation of agricultural land, it must be assumed that the ABR would form part of an integrated treatment unit that removes additional constituents.

The experimental study fulfilled these expectations, with significant COD and suspended solids reduction observed in the treatment of domestic wastewater, but no significant change in phosphate concentrations, and a net increase in free and saline ammonia.

Enumeration of pathogen indicator organisms (total coliforms, *E. coli*, coliphage and helminth eggs) in the influent and effluent of the pilot-scale ABR in Phase IV in each case showed significant removals at 95 % confidence levels. However all indicator organisms were observed in all effluent samples indicating that further disinfection is required before ABR effluent can be reused.

6.1.4 Gap analysis

There were several shortcomings in the pilot-scale ABR experimental study that negatively impact on the ability of the research to fulfil the objectives. These can be divided into two categories: (i) Physical design of the pilot-scale ABR and (ii) missing experimental data.

6.1.4.1 Design of the pilot-scale ABR

The project budget did not have provision for fundamental changes in reactor design after initial construction. This section lists problems identified with the physical design of the pilot-scale reactor that restricted the information that could be obtained from the study.

Upflow velocity vs. applied hydraulic retention time: The hypothesis for this research was that the critical design parameter was the A-HRT, and that the most important variable to be manipulated was the feed flow rate. Therefore the pilot-scale ABR was designed with little consideration for internal flow velocities. However, it emerged that the upflow velocity that was experienced when operating at relatively low hydraulic loading rates was too high for a stable anaerobic microbial consortium to establish at the low OLRs. Consequently, the pilot-scale ABR performance was relatively poor at A-HRT values approaching the operational target of 20 h. This research therefore cannot predict performance of baffled reactors with width:height ratios that allow lower upflow velocities than obtained in the experimental study at similar OLRs, or of baffled reactors with similar external dimensions but fewer compartments.

Feed condition: The feed for the pilot-scale ABR was obtained by a submersible pump from the influent channels at head of works at the two WWTP. This was necessary since the flow to the WWTP had to be subsampled (the entire flow could not be treated in the experimental rig) and gravity

sampling was not possible since in both cases, the influent channels were well below ground level. The problem with the feeding system was that the submersible pumps macerated the feed; therefore the feed to the ABR was not identical to the influent WWTP wastewater (although the chemical composition was of course not changed). In addition, the pumps used were not well suited to the pumping of raw sewage and the impellors were regularly entangled with strings and rags, resulting in low flow and reactor down time.

Gas measurement: It proved impossible to obtain accurate measurements of gas production rates from the pilot-scale ABR system as a result of the pressure buffering provided by the standing baffle system. Gas production in the reactor displaced liquid within the reactor as well as in a liquid displacement gas measuring system. To overcome internal pressure buffering, all compartment gas production needed to be collected simultaneously to prevent the increased pressure in each compartment being redistributed to neighbouring compartments. This was not possible with the available equipment. Without CH_4 production data, it was not possible to complete a mass balance to determine the extent of treatment achieved in the pilot-scale ABR.

6.1.4.2 Missing experimental data

Although the experimental data yielded much information relating to mechanism and rate of domestic wastewater treatment in the ABR, accurate measurements of the feed and outflow stream biodegradability were not made. At the time, it was assumed that the feed wastewater was similar to wastewater in any other facility in the region, and that the residual biodegradability of the outflow stream would be negligible. However, the absence of these data has two consequences;

- Firstly, since there were no gas flow and composition measurements, it is not possible to accurately calculate the amount of COD converted during treatment in the ABR, and therefore how much CH_4 was produced or how much particulate biodegradable COD accumulated; and

- Secondly, it is not clear how close the system approached to completely removing degradable COD from the wastewater, and therefore what the outlet stream characteristics (COD, alkalinity, free and saline ammonia) would have been if near-complete treatment had been achieved.

To address these gaps, two exercises were undertaken: (i) Mass balance analysis of all COD data to determine probable CH_4 production rates (**Section** 6.2), and (ii) Estimation of outflow stream characteristics from stoichiometric principles (**Section** 6.4).

6.2 MASS BALANCE TO DETERMINE PROBABLE SLUDGE/METHANE PRODUCTION RATES

In this section, mass balance principles were used to determine the composition and quantity of the three outputs of the ABR, i.e. the liquid outflow stream, biogas produced and accumulated sludge. The pilot-scale ABR is considered to be a non-steady-state system with continuous feed, liquid outflow and biogas streams, but with accumulating solids that are removed discontinuously (**Figure 6.2**)

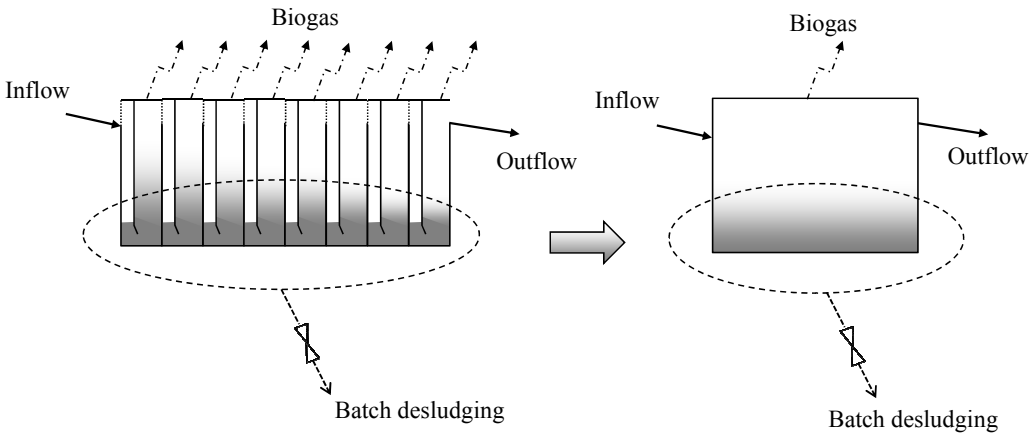


Figure 6.2: Black box representation of ABR for mass balance modelling

6.2.1 COD mass balance

The COD mass balance is described by Eq. 6-1

$$COD_{\text{inflow}} - COD_{\text{outflow}} - COD_{\text{accum.sludge}} - CH_4 = 0 \qquad \textbf{Eq. 6-1}$$

Measurements of inflow and outflow COD concentration were available for all operating periods. In Phase III and Phase IV, values for sludge accumulation rate in terms of volume of settled sludge in each upflow compartment are available and in Phase IV, a value for sludge accumulation rate in terms of mass of dry solids in each compartment is available (**Table 5.4**). All values are reported with 95% confidence limits.

These values cannot be used directly in a mass balance since the units are dissimilar. Therefore, the following assumptions were made:

- The relationship between the volume of settled sludge and its mass is fixed; therefore the mass of the settled sludge can be determined for Phase III from the ratio of settled sludge volume to mass from Phase IV.

- The relationship between the mass of COD of settled sludge and its mass is fixed.

6.2.1.1 Interpretation of sludge accumulation data

Sludge accumulation rates were calculated from data obtained in the upflow side of each compartment only, so the reported values cannot be used directly in the mass balance since they underestimate the total amount of sludge accumulated. Since upflow compartments constituted 2/3 of the total cross-sectional area of the compartment, it was proposed that the actual sludge load was 1 ½ times that calculated. Accumulation rates reported in **Table 5.4** were increased accordingly for use in mass balance calculations.

6.2.1.2 Selection of COD:TS ratio

There were limited data for simultaneous TS and VS measurements on the same samples of compartment sludge. A single data set for compartment sludge was available for Phase III, and a number of data sets for inflow and outflow stream TS and VS concentrations were available for Phase II and Phase III. From these data, the ratio of VS to TS was estimated at 0.57 g VS/g TS.

Wentzel (2006) recommends a COD to VSS ratio of 1.48 gCOD/gVSS for municipal sewage, based on large amounts of experimental data. Comparison of COD and VS data when both were available from this study yielded a value of 1.6 gCOD/gVS, while the ratio between COD and TS was 1.0 gCOD/gTS.

A value of 0.9 gCOD/gTS was chosen in the following calculations since it was in the middle of the range of the values predicted by all combinations of the above data.

A summary of the mass balance calculations is presented in **Table 6.1**.

6.2.1.3 Calculation of CH_4 production for Phase IV

The CH_4 production rate was calculated from the following general mass balance equation:

$$CH_4 = COD_{\text{inflow}} - COD_{\text{outflow}} - COD_{\text{accum. sludge}} \qquad \textbf{Eq. 6-2}$$

where all terms were calculated in units of kg COD/year.

The accumulated sludge consists of three components, accumulated active biomass (HBA), accumulated biodegradable particulate material (SBCOD) and accumulated unbiodegradable particulate material or inert solids (UPCOD). However, the experimental data do not differentiate between these three fractions.

The calculated CH_4 production rate for Phase IV was 258 kg COD/year, which amounts to 90 m^3 CH_4/year at standard temperature and pressure (STP)

Since there was no clear indication of the error and variance on the TS to COD conversion factor, or the scaling factor used to predict overall sludge accumulation rates, it was not possible to rigorously determine confidence limits for this value. CH_4 production rate was calculated from the minimum and maximum combination of inflow, outflow and sludge accumulation values, and fell in the range of 190 to 330 kg COD/year, or 65 to 120 m^3 CH_4(STP)/year.

Table 6.1: Mass balance determination of CH_4 production rates in Phase III and Phase IV.

Quantity	Unit	Phase III	Phase IV
Upflow sludge accumulation rate	m^3 settled sludge/year	*3.46*[1]	*0.901*
Upflow sludge accumulation rate	kg TS/year	233	*61*
Total Sludge accumulation rate	kg TS/year	350	92
Total Sludge accumulation rate	kgCOD/year	315	82
Inflow COD	kg COD/year	*820*	*420*
Outflow COD	kg COD/year	*260*	*80*
CH_4 production rate	kg COD/year	250	260
CH_4 production rate	m^3 CH_4/year	87	90
Fraction of Total COD removed	%	69	81
Fraction of inflow COD converted to CH_4	%	30	61
Fraction of inflow COD accumulated as sludge	%	38	19
Sludge accumulation rate	m^3 settled sludge/kgCOD	0.0063	0.0032
Sludge accumulation rate	kgTS/kgCOD	0.43	0.22
Conversion factors			
Volume to TS	kgTS/m^3	*67*	
COD to TS	kg COD/kg TS	0.9	
COD to volume CH_4 (STP)	m^3 CH_4 (STP)/kg COD	0.35	
Upflow comp load to total load	Total load/load in upflow comp	1.5	

[1] Values in italics were determined directly from experimental data. Other values were inferred from experimental data and calculated using the reported conversion factors.

6.2.1.4 Calculation of CH_4 production for Phase III

In Phase III, there were no TS data for sludge in individual compartments. However, in both Phase III and IV sludge accumulation data were available as volume of settled sludge in compartments. A conversion factor for mass of TS per volume of settled sludge was calculated from the two data sets in Phase IV generating a value of 67 kg TS/m^3 settled sludge. This value was used to convert the accumulation rate of settled solids to kg TS for Phase III data, generating a value of 350 kg TS/year .

From this value, the estimated CH_4 production rate was 250 kg COD/year or 87 m^3 CH_4(STP)/year with a probable range of between 23 and 150 m^3 CH_4(STP)/year (**Table 6.1**)

6.2.1.5 Significance of CH_4 production rate estimates

The values presented in **Table 6.1** provide an indication of the amount of CH_4 produced in Phase III and Phase IV, but cannot be validated against experimental data due to difficulties in obtaining measurements of gas production. However, the numbers support the conclusions about biological activity drawn in **Section** 6.1.2, i.e. that higher sludge accumulation rate per OLR values were obtained in Phase III than in Phase IV due to the failure of anaerobic micro-organisms (particularly acetoclastic methanogens) to establish. In spite of the uncertainty in the calculated values, it is clear from the mass balance calculations that the amount of CH_4 produced (per mass of COD fed to the system) in Phase III was significantly lower than in Phase IV, despite the fact that there was more degradable material present at the higher loading rates of Phase III.

Table 6.1 also presents values for fraction of COD removed from the wastewater stream, and fraction converted to CH_4.

In Phase III, although nearly 70% of COD was removed, only 30% of the COD was likely to have been converted to CH_4, while the balance was accumulated as biodegradable solids. In contrast, around 60% of the COD applied in Phase IV was converted to CH_4, with an overall COD removal of 80%. The overall mechanism of treatment was fundamentally different between the two phases, with solids retention dominating in Phase III and solids digestion dominating in Phase IV.

These results do not indicate what the overall extent of treatment in the pilot-scale ABR was; i.e. there is no indication of what the residual biodegradability of the outflow stream was.

Clearly, since the outlet COD concentrations were lower in Phase IV than in Phase III, at least 10% of the inflow COD was potentially degradable, but not removed from the wastewater stream in Phase III. In Phase IV, the outflow had a COD concentration that was 20% of the inflow COD concentration. It is expected that there was very little biodegradable material in this stream since the residence time in the ABR was long.

6.2.2 Nitrogen mass balance

The fate of reduced nitrogen in anaerobic digestion at low pH values (< 7) may be described as follows:

Reduced nitrogen enters an anaerobic system as free and saline ammonia or organically bound nitrogen. The latter is nitrogen associated with organic compounds such as the amino group in peptides and proteins. Influent free and saline ammonia remains relatively unchanged (although ammonia may be used in small amounts for growth processes, if reduced nitrogen is not present in a more utilisable form). Organically bound nitrogen is liberated as free and saline ammonia during the

anaerobic digestion of the organic compound to which it is bound. At pH values below 7, very little ammonia will enter the gas phase as the predominant species under these conditions is the associated ammonium form (NH_4^+), which does not exchange directly with the gas phase.

Considering reduced nitrogen in an accumulating anaerobic system; free and saline ammonia from the feed will pass through the digester relatively unchanged and exit in the outflow stream. Organically bound nitrogen has three possible fates (i) a portion passes through the digester unchanged; (ii) a portion is liberated as free and saline ammonia during anaerobic digestion, and this exits the digester with the free and saline ammonia from the feed; (iii) the remainder remains undigested, bound to particulate organic compounds that are retained in the digester due to sludge accumulation.

There are limited nitrogen data available for Phase I, Phase III and Phase IV of the experimental study. These are presented in **Table 6.2**.

Table 6.2: Free and saline ammonia and Total Kjeldahl Nitrogen (TKN) concentrations in inflow and outflow streams of the ABR. Average values ± confidence interval and number of observations (n) are presented.

	Inflow TKN	Inflow free and saline ammonia	Outflow free and saline ammonia
Phase I day 206-255	42 ± 5[1] (n=21)	23 ± 1 (n=271)	31 ± 8 (n=8)
Phase I day 262-409	42 ± 5 (n=21)	23 ± 1 (n=271)	50 ± 11 (n=14)
Phase III	45 ± 3 (n=8)	39 ± 2 (n=273)	34 ± 3 (n=7)
Phase IV	45 ± 3 (n=8)	39 ± 2 (n=273)	51 ± 7 (n=10)

[1] Inflow values for TKN and free and saline ammonia are calculated from all data available for each installation (Umbilo WWTP – Phase I; and Kingsburgh WWTP – Phase III and Phase IV)

It can be seen that outflow free and saline ammonia values were lower than inflow TKN values for Phase I (day 206 to 255) after increasing feeding rates, and for Phase III, but similar to inflow TKN values for Phase I (Day 262 to 409) and Phase IV.

In Phase III, there is no apparent increase in free and saline ammonia, indicating that little digestion has occurred. This concurs with the proposal that during Phase III, the principle mechanism of COD removal was through solids accumulation.

It has already been shown that improved digestion rates were inferred for the second part of the 20 h THRT period of Phase I and for Phase IV. The nitrogen data presented here suggest that either all inflow TKN appears in the outflow as free and saline ammonia (indicating complete liberation of

organically bound N) or that N accumulated from earlier phases characterised by poor digestion rates is digested and liberated in these two periods. The latter explanation may be valid for Phase I, but is unlikely to hold for Phase IV since a long standing period existed between Phase IV and previous operating periods. Thus in Phase IV, it is assumed that little nitrogen is accumulated with the accumulated sludge. This suggests that most of the accumulated sludge is inert particulate material.

6.3 DEVELOPMENT OF A SLUDGE AGE MODEL FOR AN ACCUMULATING SYSTEM

The concept that methanogenic micro-organisms can fail to establish in the pilot-scale ABR while simultaneously, high rates of sludge accumulation are observed is initially difficult to accept. The literature indicates that the ability of methanogenesis to establish in a steady-state system such as a UASB depends on the SRT achieved in the system; In their review of available data, Zeeman and Lettinga (1999) reported that methanogenesis could be achieved during digestion of manure when the SRT was 100 days, but no methanogenesis was observed for SRT of 50 days. These authors published a model for calculating the HRT required to give a certain SRT. Zeeman and Lettinga went on to state that a system should be designed by selecting SRT values that are large enough for methanogenesis to occur.

The model of Zeeman and Lettinga (1999) is presented in **Appendix** A5.1. This model defines the SRT as the ratio of the sludge concentration in the reactor X [g COD/ℓ] to the (reactor) net sludge production X_P [gCOD/ℓ.d] (where sludge, X, is any particulate matter in the reactor):

$$SRT = X / X_P \qquad \textbf{Eq. 6-3}$$

However, when this definition is applied to a sludge accumulating system such as the ABR, the calculated sludge age is greater than the operating period of the system (working shown in **Appendix** A5). Thus the Zeeman and Lettinga definition of sludge age does not apply to an accumulating system.

A model of sludge age in an accumulating system was developed and is presented in its entirety in **Appendix** A5.2. It can be readily shown that in an accumulating system, the sludge age increases continuously; i.e. there is no characteristic sludge age that describes the system.

For a *package* of sludge generated in the time interval $[t_{i-\Delta t}, t_i]$ (by retention of influent solids or ultimately by generation of biomass through growth), at t_e the sludge age of this *package* of sludge is $(t_e - t_i)$. If the amount of sludge generated in the time interval $[t_{i-\Delta t}, t_i]$ that ultimately accumulates in the reactor is X_P, and if X_P is approximately constant, then the average sludge age at the end of a period, time t_e depends on the sludge load at start-up, X_0, the sludge production rate X_P and the length of operation t_e and can be approximated by Eq. 6-4

$$SRT = \frac{\frac{1}{2} t_e^2 X_P}{X_0 + X_P \cdot t_e} \qquad \textbf{Eq. 6-4}$$

The complete derivation is presented in **Appendix** A5.2.

In this analysis it can be seen that the SRT is related to the integral of the sludge production X_P: in an ABR, X_P depends on the rate of solids retention through settling and the rate of solids hydrolysis through biological activity. If the rate of generation of micro-organisms is low micro-organisms may washout since the inflow of micro-organisms to the system is negligible. It is therefore quite possible that high solids accumulation rates X_P are due to low rates of biological activity, despite poor biomass retention, *but only if the net rate of accumulation (in – out + generation – degradation) of biodegradable solids is larger than the rate of accumulation of the micro-organisms that should degrade them*.

Thus design guides that indicate that methanogenesis will establish at SRT values above some critical value are empirical and are only valid for the type of systems from which they were determined.

It is further concluded that neither the classically defined SRT nor the SRT defined in this analysis are appropriate design parameters for an accumulating system.

6.4 CALCULATION OF RANGE OF OUTFLOW CHARACTERISTICS

The outlet stream pH value was consistently below that of the inflow to the ABR, despite increases in measured alkalinity values between the inlet and outlet. Initially, it was thought that this indicated process instability as a result of inhibition of acid-removing micro-organisms. However, similar studies on anaerobic digestion also found that pH values decreased without apparently affecting process stability (Behling et al., 1997).

Sötemann et al. (2005) demonstrated that for an anaerobic process operated at steady state where hydrolysis was the rate limiting step, the effluent characteristics could be accurately predicted if the feed characteristics were sufficiently well understood. It was proposed that the stoichiometry of Sötemann et al. (2005) could be used to predict the range of effluent characteristics that could be expected from an anaerobic system treating domestic wastewater, and that this information would assist in understanding the condition of the outflow (and digestion as a whole) in this study.

6.4.1 Sötemann et al. (2005) model of steady-state anaerobic digestion

Steady-state models are based on the principle of the *rate-determining step*: in a steady-state system, the overall rate of treatment will depend on the slowest process that occurs in the system. Provided the conditions of the system do not change such that another process becomes rate-limiting, a calibrated steady-state model will give a reasonably quick basis for designing a system and determining operating parameters, or estimating system performance under slightly different conditions.

Sötemann et al. (2005) developed a steady-state model for anaerobic digestion of sewage sludges, based on the assumption that hydrolysis of macromolecules is the rate-limiting step in this process. This is a three step model consisting of (i) a kinetic part for determining COD removal and gas production, (ii) a stoichiometric part that calculates free and saline ammonia, alkalinity production and digester gas composition and (iii) a weak acid-base section that calculates the digester pH from the gas composition and alkalinity.

The COD of the feed in the steady-state model is assumed to be a combination of particulate biodegradable organic materials (SBCOD, **Section** 2.4.1) with a known average elemental composition $C_XH_YO_ZN_A$, VFA (represented by acetic acid) and a fraction of unbiodegradable material.

All VFA is consumed in the process, and a portion of the remaining biodegradable organic material is converted to CH_4, CO_2, alkalinity and free and saline ammonia. The extent of biodegradation, i.e. the amount of SBCOD degraded, depends on the sludge age or length of contact time in the system.

6.4.2 Steady-state model implementation

The purpose of this exercise was not to simulate the experimental data, but to determine what the probable outflow conditions would be for a range of inflow characteristics and at different values for the extent of treatment achieved. Therefore, kinetics of biodegradation were not initially considered to be important. However, the kinetic part of the Sötemann steady-state model could not be ignored completely since in the original form, the sludge age and kinetic constants (endogenous respiration rate) determine the apparent yield of the process, E, defined as the fraction of removed COD converted to sludge through microbial growth. A value for E is necessary for determining overall stoichiometry.

6.4.2.1 Sötemann model of growth-death-regeneration

The original Sötemann model makes use of a growth-death-regeneration model, called the hydrolysis model; i.e. biodegradable particulate material (denoted S_{bP}) and VFA (S_{bVFA}) are converted to CH_4, CO_2 and biomass; biomass undergoes endogenous decay, where more S_{bP} is a product of the endogenous decay.

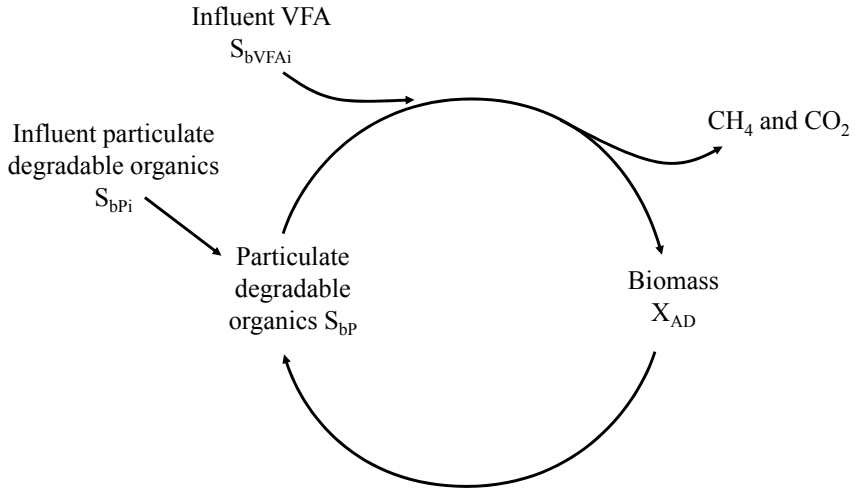


Figure 6.3: Growth-Death-Regeneration scheme used in the hydrolysis part of the steady-state model of Sötemann et al. (2005)

Thus the initial biomass yield from degradation of feed biodegradable organic material may be high (Sötemann et al. (2005) recommend a value of 0.113 gCOD/gCOD for Y_{AD}, the yield co-efficient), but the overall yield calculated by mass balance over the system operated at steady state (E) may be significantly lower. Longer sludge ages and higher endogenous rates result in lower values of E, while for short sludge ages and low endogenous rates, E will be close to Y_{AD}.

Sötemann et al. (2005) reviewed a considerable amount of data on primary sludge digestion in order to determine values for rate constants for the hydrolysis model. As there were no experimental data available for calculating the apparent yield, E, it was proposed that the hydrolysis model with kinetic constants proposed by Sötemann et al. (2005) with appropriate corrections for temperature be employed to estimate a value for E.

The original steady-state model describes a CSTR with no unsteady sludge accumulation. Consequently, the HRT and sludge age are identical and are calculated from the ratio of the volume of the reactor V and feed flow rate Q. As with Zeeman and Lettinga's definition of sludge age (**Section** 6.3), this definition is not appropriate for an accumulating system. Thus a new hydrolysis model had to be developed to describe the effect of sludge retention on apparent yield in an accumulating system.

6.4.2.2 Hydrolysis model for a solids retention system

The same general methodology as published in Sötemann et al. (2005) was followed, with the following assumptions and changes:

- A pseudo-steady-state condition was assumed. This implies that the concentrations of reacting species in the digester did not change with time, i.e. biomass and SBCOD concentration were approximately constant. (The corollary of this assumption is that only inert solid material accumulated in the digester. This is in line with observations from the nitrogen balance that little nitrogen accumulates with solids in the digester, **Section** 6.2.2)

- The concentration of solids exiting the digester (X_{ADe}) was a fixed fraction f_X of the total solids concentration in the digester (X_{AD}):

$$X_{ADe} = f_X X_{AD} \qquad \textbf{Eq. 6-5}$$

Using these assumptions, an expression for the apparent system yield, E was derived:

$$E = \frac{f_X \cdot Y_{AD}}{f_X + \frac{V}{Q} \cdot b_{AD}(1 - Y_{AD})} \qquad \textbf{Eq. 6-6}$$

Where

f_X = Ratio solids COD conc. in outflow: average solids conc. in reactor
Y_{AD} = Yield co-efficient for acidogenic micro-organisms
V = Volume of digester [m^3]
Q = Feed flow rate [m^3/d]
b_{AD} = Endogenous decay rate [d^{-1}]

The complete derivation is presented in **Appendix A6**.

No differentiation was made between unbiodegradable particulate material (UPCOD) that is retained in the digester, and that which passes out with the effluent. Therefore, it is not possible to accurately predict outflow COD concentrations (denoted S_{te} in the model) since these values are inflated by the concentration of UPCOD that should be retained in the ABR.

The pseudo-steady-state assumption that implies that reacting species concentrations are approximately constant and that only UPCOD accumulates, seem reasonable when considering data from Phase IV where outlet species concentrations do not appear to change significantly with time.

However, these assumptions are clearly not valid for conditions such as prevailed in Phase III where undegraded SBCOD appears to have accumulated.

The stoichiometric portion and weak acid-base chemistry of Sötemann's steady-state model were directly incorporated into this solids retention model without alterations. These are presented in **Appendix A6**.

6.4.3 Inputs into the steady-state model of the ABR

Anaerobic digestion feedstocks are conventionally described in terms of their carbohydrate, protein and lipid components (Batstone et al., 2002), since each of these categories may be represented by characteristic elemental compositions; i.e. carbohydrates have elemental compositions similar to $(CH_2O)_n$; proteins contain nitrogen, and lipids have high C:O and H:O ratios. Although it is not usually practical to characterise the feed by measuring these constituents, it is useful to represent the feed in terms of these since it is easy to visualise changes in feed composition in terms of the relative contribution of each of these categories.

The average elemental composition of generic carbohydrate, lipid and protein compositions taken from Henze et al. (1992) were used to calculate the overall average elemental composition of the biodegradable organic material in the feed as follows:

Carbohydrate:	$C_{10}H_{18}O_9$	fraction = i (mol %)
Lipid:	$C_8H_6O_2$	fraction = j (mol %)
Protein:	$C_{14}H_{12}O_7N_2$	fraction = k (mol %)

Table 6.3: Inflow composition for model components for steady-state model taken from Kingsburgh WWTP inflow wastewater characteristics. These values were used as a base case for the sensitivity analysis.

Component	Unit	Value
COD	mgCOD/ℓ	680
Unbiodegradable COD	mgCOD/mgCOD	0.08[1]
Alkalinity	$mgCaCO_3/ℓ$	242
free and saline ammonia	mgN/ℓ	39
pH	-	7.0
VFA	mgCOD/ℓ	35
Temperature	ºC	25
Protein	% of inflow SBCOD	20
Carbohydrate	% of inflow SBCOD	35

[1] Not measured. This value was taken from Sötemann et al. (2005)

A filter for converting feed SBCOD composition from the carbohydrate-lipid-protein characterisation to the elemental composition ($C_XH_YO_ZN_A$) is presented in **Appendix** A6.2.

The wastewater characterisation presented in **Table 6.3** represents an average composition of Kingsburgh WWTP and was used as a base case for investigation of effluent characteristics using the steady-state model.

An input variable, the extent of treatment f_E was defined in order that outflow conditions may be calculated for ranges of treatment efficiency. **Table 6.4** presents values used in the *kinetics* part of the steady-state model that is used for predicting the apparent sludge yield of the pseudo-steady-state model.

Table 6.4: Model parameters used in or calculated by the steady-state model

Parameter	Unit	Value
Volume (V)	ℓ	3 000
Flow rate (Q)	ℓ/d	1 800
Extent of treatment (f_E)	mgCOD/mgCOD	0.80
Fraction of average reactor solids concentration exiting with outflow stream (f_X)	mgCOD/mgCOD	0.01
Apparent sludge yield (fraction of removed COD converted to sludge through microbial growth) (E)	mgCOD/mgCOD	0.016
Endogenous rate constant (b_H)	d^{-1}	0.041
Acidogen yield (Y_{AD})	mgCOD/mgCOD	0.113

6.4.4 Comparison of steady-state model predictions with experimental data

Several of the parameters in **Table 6.3** were estimated and may not have been a true reflection of the wastewater characteristics (VFA, protein, carbohydrate, unbiodegradable COD). Therefore, it was not expected that the steady-state model would precisely match the measured properties of the experimental study.

The steady-state model with sludge retention developed in **Appendix** A6 was used to calculate an apparent yield, E. This value represents the reduction in sludge yield due to endogenous respiration of biomass resulting from the long retention of solids in the reactor. The value obtained was 0.016 mgCOD/mgCOD; i.e. of the SBCOD removed from the wastewater stream, approximately 1.6% is converted to biomass. Since the model is derived assuming that the concentrations of SBCOD and biomass do not change significantly with time, this value implies that 1.6% of the influent SBCOD exits the reactor as washed out biomass.

Figure 6.4 shows the agreement between the measured and simulated outlet characteristics for pH value, alkalinity concentration and free and saline ammonia concentration. Relatively good agreement is obtained between simulated and measured values for all three categories, with the simulated value for free and saline ammonia and alkalinity falling within the confidence limits of the means of the measured values.

The calculated pH value is compared to the median of all measured values on samples drawn from the last compartment of the pilot-scale ABR. It is significant that the pH values are similar. The measured value is slightly higher than the simulated value, but even the small difference observed is not unexpected since evolution of CO_2 from samples that are exposed to the atmosphere after sampling and before measurement is expected, and widely reported as a source of error in pH measurement in closed systems (Stumm and Morgan, 1996; Sötemann et al., 2005). These results indicate that a final digester pH value close to 6.0 is to be expected from digestion of domestic wastewater with characteristics described by **Table 6.3**.

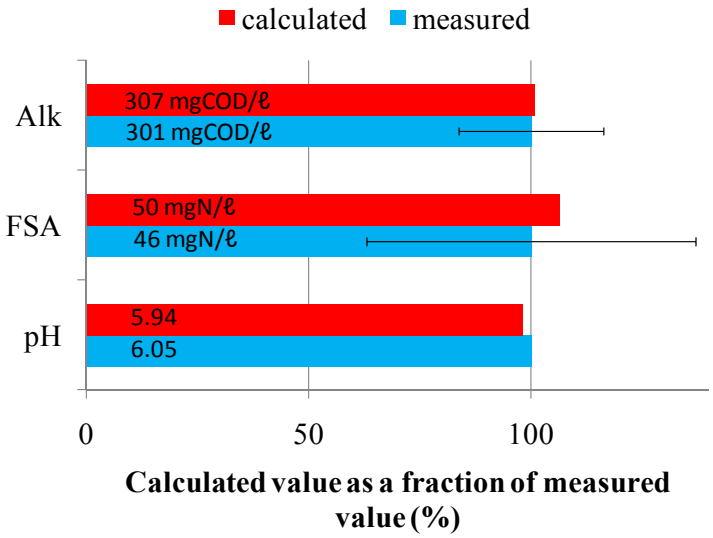
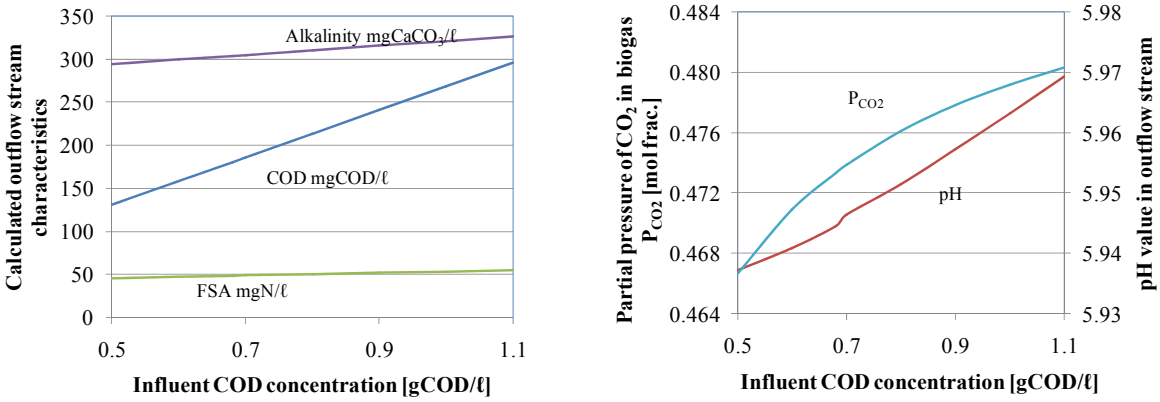


Figure 6.4: **Steady-state model predictions of outlet characteristics alkalinity, free and saline ammonia (NH_3+NH_4^+) and pH value compared to average or median values obtained from Phase IV operation.**



(a) **(b)**

Figure 6.5: **Steady-state model predictions of outlet characteristics (a) Alkalinity, outflow/accumulating COD, free and saline ammonia and (b) P_{CO2} and pH value for increasing influent wastewater strength**

Figure 6.5 shows the results of simulations estimating the effect of increasing influent wastewater strength at the same extent of treatment (e.g. for 80% of SBCOD removed, $f_E = 0.80$) on conditions in the digester. Higher feed concentrations result in more digestion occurring, resulting in increased

amounts of free and saline ammonia, and therefore alkalinity being generated. For the wastewater composition, it can be seen that increasing the wastewater strength increases P_{CO2} and pH values in the digester.

Sensitivity analyses were performed in which the effect of changes in feed conditions from the base case described in **Table 6.3** on digester pH was investigated. It was found that pH values were most sensitive to changes in feed alkalinity, and to a lesser extent, to feed composition in terms of the ratio of carbohydrate, lipid and protein components. Digester pH values were not sensitive to feed pH values.

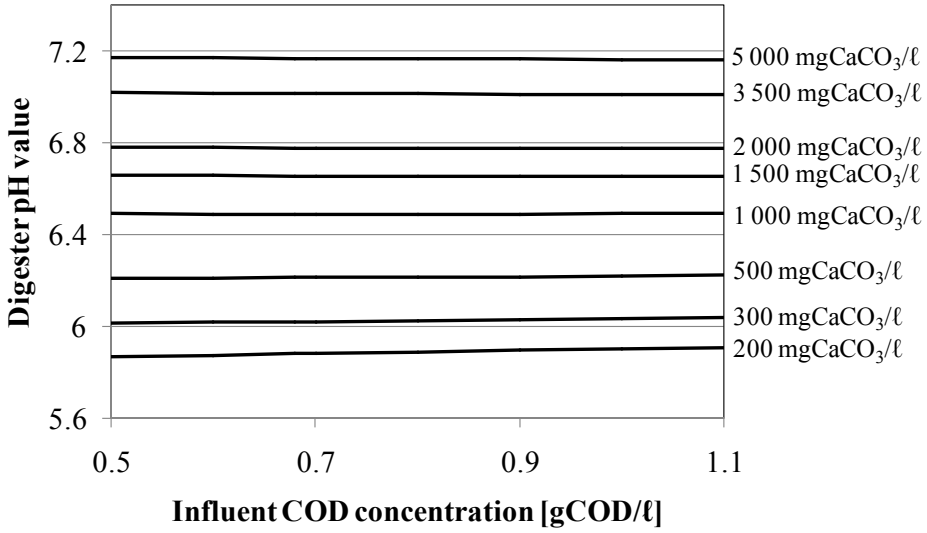


Figure 6.6: **Steady-state model sensitivity analysis: Estimated effect of influent alkalinity concentration and COD concentration on digester pH values for fixed wastewater composition (but varying strength).**

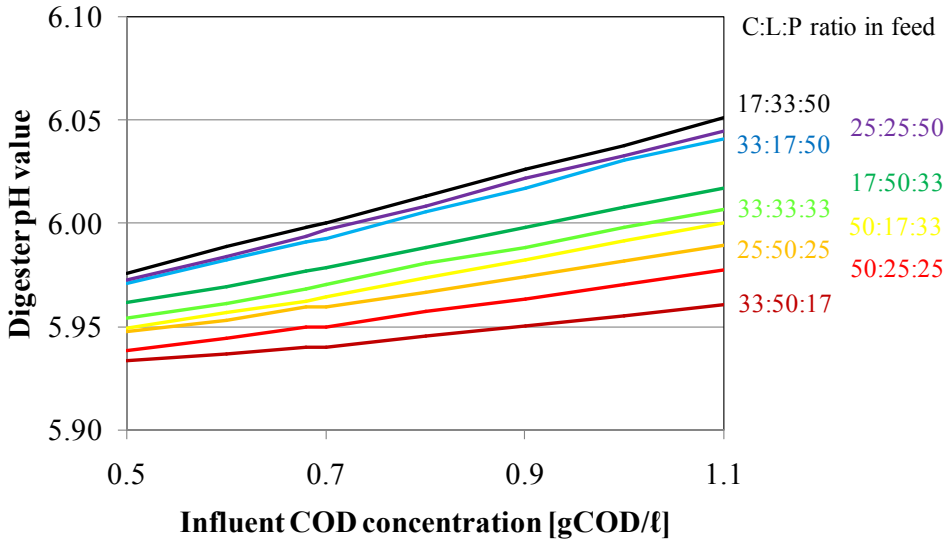


Figure 6.7: **Steady-state model sensitivity analysis: Estimated effect of ratio of carbohydrate (C), lipid (L) and protein (P) and COD concentration on digester pH values for fixed feed alkalinity.**

Figure 6.6 shows the relationship between feed strength and pH at a range of influent alkalinity concentrations (from 200 $mgCaCO_3/\ell$ to 5 000 $mgCaCO_3/\ell$). This analysis predicts that increasing

influent alkalinity will increase digester pH. **Figure 6.6** may be used to estimate alkalinity supplementation requirements should a system treating domestic wastewater be required to operate at a particular pH (e.g. above a value of 6.5) to prevent inhibition of methanogenesis and to ensure process stability. For example, **Figure 6.6** implies that inflow alkalinity would have to be increased to approximately 1 000 $mgCaCO_3/\ell$ to maintain digester pH values at 6.5[1].

Figure 6.7 shows the predicted effect of differences in feed composition in terms of ratio of carbohydrate, lipid and protein components. It can be seen that higher proportions of protein fractions (e.g. C:L:P = 17:33:50 or 25:25:50) result in higher digester pH values. This is because only the modelled protein component is defined as having a nitrogen fraction, and it is the generation of free and saline ammonia through digestion of the protein fraction that principally results in the production of alkalinity in anaerobic digestion processes (Speece, 1996; Sötemann et al., 2005).

Changes in the value of f_E (i.e. amount of influent COD removed through accumulation or digestion) result in changes to the amount of sludge accumulated and CH_4 produced. Changing f_X (Fraction of solids COD that leaves the reactor) results in changes to the value of E (apparent yield). Changing the extent of digestion or sludge age in this model results in predicted outputs that are similar to scenarios in which the feed SBCOD concentration changes, since in all of these cases, it is the amount of SBCOD digested that changes in the steady-state model.

6.4.5 Significance of steady-state modelling study

The steady-state modelling exercise demonstrated that the relatively low digester pH values observed in the experimental study were a result of the low feed strength and low alkalinity generation potential of the wastewater treated.

In **Section** 2.1.6 the importance of alkalinity and pH buffering capacity on stability of anaerobic digestion processes was discussed. Should insufficient acid neutralisation capacity exist in an anaerobic system, low pH values could affect the establishment of stable anaerobic micro-organism consortia, the degree of treatment that could be obtained in a system and thus the quality of the effluent produced. The alkalinity depends on three factors:

- Alkalinity of the incoming wastewater,
- Alkalinity generation potential of the wastewater, and
- Operating conditions

The sum of the incoming alkalinity concentration and the alkalinity generation potential should reach a value that places the reactor pH in a region where pH inhibition effects do not result in poor methanogenesis. The steady-state model predicted that a digester alkalinity concentration of around 1 000 $mgCaCO_3/\ell$ would maintain digester pH values at 6.5. The alkalinity of the incoming wastewater may not necessarily affect process stability if there is sufficient alkalinity generation potential in the wastewater to achieve these alkalinity levels.

[1] Note that corrections for high ionic strength on equilibrium constants have not been included in this analysis, and therefore values at high alkalinity concentration should be interpreted with caution.

Wastewater composition affects the alkalinity generation potential; wastewater with high protein content will result in the production of ammonium ions during the anaerobic digestion of proteinaceous organics, thereby causing an increase in overall alkalinity. However, digestion of high carbohydrate or lipid containing wastewaters results in a smaller net increase in alkalinity across the process.

Large oscillations in hydraulic and organic load can lead to the development of transients in VFA concentration which could precipitate a souring event and hamper the development of stable anaerobic consortia, especially in the form of granular sludge. Therefore, when operating conditions are not stable, it is necessary to have a higher digester alkalinity concentration to provide additional buffering capacity to neutralise potential VFA accumulation

It is concluded that for the hydrolysis-limited case, *the alkalinity, and alkalinity generation potential are the most important variables for maintaining reactor stability*. The alkalinity generation potential is determined by the relative compositions of proteins, carbohydrates and lipids in the feed and the wastewater strength. (The VFA concentration also affects the alkalinity generation potential, but the concentration of VFA in domestic sewage is not expected to be high). It may be inferred that where low pH values may be resulting in pH inhibition of methanogenesis, increasing alkalinity will also result in improved COD reduction by causing an increase in the rate of methanogenesis.

Finally this study has indicated that anaerobic digestion of domestic wastewater without alkalinity supplementation can be achieved, as seen in the experimental study, but this is likely to result in significant inhibition of methanogenic micro-organisms (**Section** 2.1.5.2) and associated low growth rates may affect the maximum upflow velocity that can be achieved in the system (**Section** 6.1.2.1)

6.5 BIOCHEMICAL MODEL OF THE ABR

The last objective of this study was to develop a dynamic mathematical model of the biochemical processes of an ABR as a design tool. However, there were two major obstacles to the achievement of this objective.

- *Hydrodynamics and solids retention:* the ABR design poses a severe challenge in terms of hydraulic modelling. The hanging and standing baffles divide each compartment into a downflow / settling zone and an up-flow / fluidised bed zone. There is considerable solids retention in each compartment, but also some solids carry-over between compartments. The solids build-up in each compartment must have some effect on the available volume for liquid flow in each compartment, and hence the residence time distribution of soluble and particulate species within each compartment. The solids are effectively the catalyst for the biological conversion. Thus, in order to accurately model the system, and to be able to predict the behaviour of the system under different operating conditions, an understanding of the relationship between the solids retention characteristics and flow rate is required.

- *Biochemistry*: Secondly, the experimental study was not able to tease out the relationships between inflow alkalinity concentration, organic loading rate, upflow velocity and washout rate of anaerobic micro-organisms. Thus a detailed hydrodynamic/biochemical model could not be satisfactorily calibrated to be able to predict system behaviour under conditions different to those tested.

The original objective to develop a dynamic mathematical model of biochemical processes for the purposes of designing ABR systems was over-optimistic, given the quality of the experimental data available. These data were mined for design information about process limitations (upflow velocity in this case) and for CH_4 and sludge production estimates. More sophisticated design tools can only be developed with the assistance of additional experimental data describing the relationships between observed sludge accumulation or retention and applied upflow velocity, alkalinity and organic loading rate.

Nevertheless, a model was developed using the WEST[1] platform to identify what further information would be required to build a biochemical model for design purposes.

6.5.1 Model construction

The model was constructed as a series of 8 continuous stirred tank reactors with a solids retention factor (**Figure 6.8**) i.e. a small fixed fraction of the concentration of particulate species was allowed to leave each compartment.

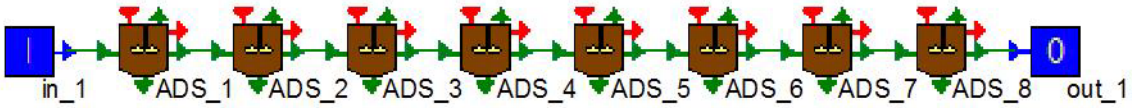


Figure 6.8: WEST® representation of the ABR flow configuration. Each element represents a continuous stirred tank reactor with sludge retention.

The biochemical model of Siegrist et al. (1993) was used to describe anaerobic conversions in the modelled ABR. **Figure 6.9** shows the flow diagram for carbon catabolism in the Siegrist model.

The Siegrist model has a number of limitations:

- Only one category of influent particulate COD is described, represented as *Biopolymers* in **Figure 6.9** Hydrolysis of this component yields amino acids, sugars and fatty acids in a fixed ratio. The stoichiometry of this process is fixed in the Siegrist model, although it is possible to manually alter the stoichiometry. There is no mechanism for allowing the ratio between the different hydrolysis products to vary during a simulation.

- The model was constructed and calibrated for mesophilic sewage sludge digestion, and therefore default parameter values for kinetic constants and stoichiometry may not be appropriate for domestic wastewater

- Protolysis and deprotolysis of volatile fatty acids are not included in the model; therefore it is not possible to simulate extreme acidification (pH < 6) of a digester (Siegrist et al., 1993).

[1] WEST: Worldwide Engine for simulation, Training and Automation. This software is built and supported by Hemmis.

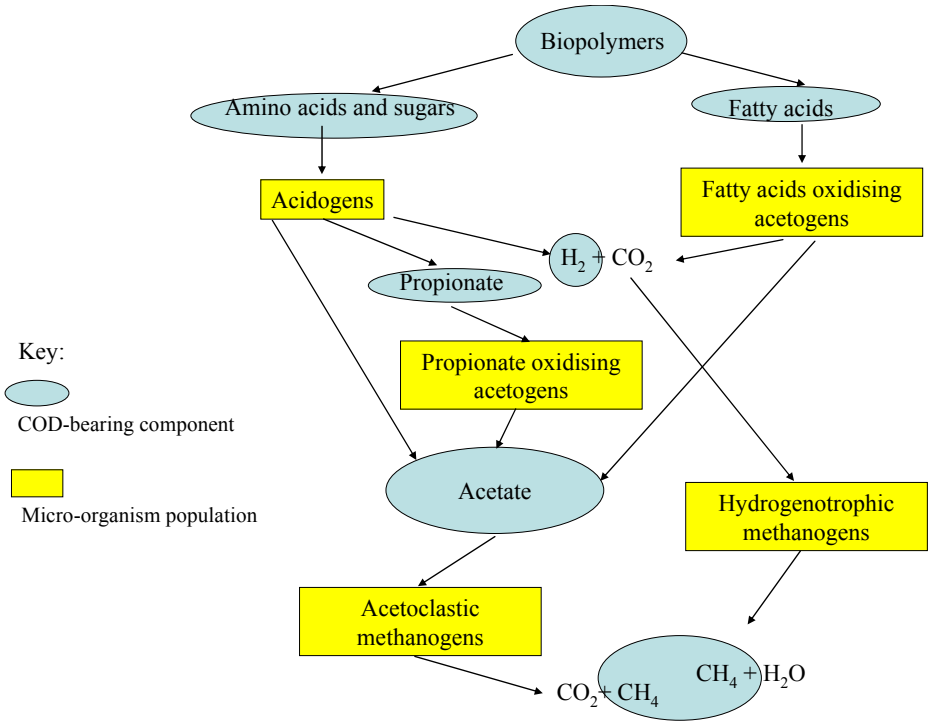


Figure 6.9: Flow diagram showing route taken during catabolic degradation of COD in the anaerobic digestion model proposed by Siegrist et al. (1993).

Details of implementation and results of the model were presented in Foxon et al. (2004) and Foxon and Buckley (2007). Copies of these conference papers are included in Annexure 1 on the CD enclosed in the back cover of this thesis.

6.5.2 Results of biochemical modelling

The modelling exercise was extremely useful in that the process of building and manipulating the model identified concepts that proved important in analysing and understanding the experimental data. These included the following:

- Except during start-up and when the reactor is already full of solids, a pseudo-steady state exists where the ABR effluent does not change substantially with time, despite ongoing accumulation of solid material
- During stable digestion, the bulk of accumulating material is inert particulate material
- It is possible to continue accumulating solid material, while washing out active micro-organisms

6.5.2.1 Simulation results

In both the 2004 and 2007 studies, the biochemical models were able to describe ammonia and alkalinity generation, when appropriate feed characteristics were specified. However, the simulations were not able to predict pH and solids dynamics in the compartments of the ABR with any reliability:

- They did not accurately depict the dynamics of soluble COD components, and components that have an effect on the pH value in each compartment (Foxon and Buckley, 2007)

- It was not possible to simulate solids accumulation with any accuracy since the experimental data were extremely noisy, and the kinetic parameters were not well understood.

- The composition of the feed in terms of alkalinity and ammonia generation potential (which can be inferred from a carbohydrate/ protein /lipid fractionation, or from elemental analysis of the feed) was required for accurate prediction of effluent characteristics. Some of these quantities were not available from the pilot-ABR experimental study (Foxon et al., 2004)

- Representation of hydrolysable COD as a single component in the Siegrist model resulted in poor agreement between predicted and measured soluble COD concentrations (Foxon et al., 2004)

6.5.2.2 Recommendations for model development

The following changes were proposed to the model structure:

- It was concluded that a biochemical model with protolysis and deprotolysis of weak acids and bases is necessary to adequately describe soluble COD and pH value dynamics in the ABR treating domestic wastewater at low inflow alkalinity concentrations. (Foxon and Buckley, 2007)

- One particulate biodegradable COD fraction is insufficient to represent decreasing average hydrolysis rates; a subdivision of particulate biodegradable COD is proposed.

Both of these changes are implemented in the Anaerobic Digestion Model No. 1 (ADM1, Batstone et al., 2002), so it seems appropriate to adopt the ADM1 model structure in future modelling.

The following additional data would improve the model's predictive performance:

- Model inputs
- Particulate and soluble organic nitrogen in the feed
- Particulate and soluble inert COD in the feed
- Total gas production
- Organic acid concentration in the feed and in compartments
- Some measure of the biomass seeding rates, or indication of where the model is sensitive to the seeding rate

- A sludge retention model that describes the relationship between upflow velocity and sludge carry-over for granular and dispersed sludge

6.6 DESIGN OF BAFFLED DIGESTERS FOR TREATING DOMESTIC WASTEWATER

The analysis presented in this chapter has indicated that the principle parameters that must be fixed for design of an ABR treating domestic wastewater are upflow velocity in compartments and overall A-HRT. This is used as the basis for presenting a design methodology for a classic (hanging and standing baffle) ABR design.

6.6.1 Design objective

In engineering terms, an ABR functions as a series of mixed reactors, in which the biological catalyst, the biomass in the sludge of each compartment is retained in that compartment when the liquid flow passes out of the compartment. The first one or two compartments have the added function of retaining solids originating from the feed.

The design objective is to increase the amount of contact time between suspended or dissolved contaminants and the biomass and decrease the amount of sludge washout in the ABR effluent. This is achieved by finding a compromise between

- increasing HRT (the treatment time);
- increasing the number of passes through the sludge bed (i.e. number of compartments);
- reducing the upflow velocity to reduce solids carry-over; and
- reducing space requirements and capital cost.

Low upflow velocity can be achieved by either selecting a reactor geometry that has a short flow path for a specified HRT (e.g. a low, wide reactor, or few compartments), or by reducing the flow to a specific reactor size, i.e. increasing HRT.

In the analysis that follows, the parameters in the process design are described, indicating the effect on process performance of the choice of parameter value.

6.6.2 Design Parameters

The classic ABR process design consists of a number of equally dimensioned compartments. For a specific wastewater flow, the design is fully specified by fixing the following 6 independent parameters: (i) design HRT, (ii) number of compartments, (iii) peak upflow velocity, (iv) compartment width to length ratio, (v) reactor depth and (vi) compartment upflow to downflow area ratio. The civil design of the reactor interior also requires values for hanging baffle clearance, headspace height, baffle construction and inlet and outlet construction. All other internal features such as length and width of individual compartments are dependent on the first six parameters.

6.6.2.1 Principle design parameters

This thesis argues that the hydraulic retention time and the up-flow velocity are the most important parameters in the design of an ABR treating domestic wastewater.

- *Hydraulic retention time:* The mean HRT affects the contact time in which wastewater treatment may occur, and indirectly, the upflow velocity, that controls solids/sludge retention. It is also the parameter that dictates the size of the reactor (working volume) and therefore has a significant effect on the capital cost of the system. Although not shown in this study, a design A-HRT of 20 h could be achieved if the upflow velocity is sufficiently low.

- *Peak upflow velocity* is the maximum permitted upflow in the reactor that does not cause an unacceptable entrainment and washout of sludge (**Section** 6.1.2.1). The peak up-flow velocity is the design velocity increased by a peak flow factor. The latter is the ratio of the peak flow expected to the average daily feed flow rate. Studies on simplified sewerage (small bore sewer systems) in poor communities in Brazil found a peak flow factor of 1.8 to be adequate for design purposes (Mara et al., 2000). It is proposed that a peak upflow velocity of 0.5 m/h be

employed for a medium strength wastewater, with low alkalinity concentration if alkalinity supplementation is not employed, and if no additional sludge retention devices are included in the design. This gives a design upflow velocity of 0.28 m/h for a medium strength wastewater (COD = 680 mgCOD/ℓ) with an alkalinity concentration of approximately 300 mg$CaCO_3$/ℓ. (In systems with higher alkalinity concentration, this value may be much higher).

Number of compartments, reactor depth, and compartment upflow to downflow area ratio all define the *peak upflow velocity* within the reactor. Independently either increasing the number of compartments, the reactor depth or reducing the compartment upflow to downflow area ratio results in an increase in upflow velocity. Except in the case of the upflow to downflow area ratio, the increased liquid velocity is caused by the lengthening of the overall path that wastewater has to traverse through the reactor (working height of reactor x number of compartments x 2 [m]).

6.6.2.2 Secondary design parameters

The number of compartments and the reactor width to length ratio are considered to be secondary design parameters.

Intimate contact between sludge and wastewater ensures efficient use of treatment volume. This means that a greater number of passes through the sludge blanket, achieved by increasing the *number of compartments* will increase the overall contact time between wastewater and sludge, and therefore increase COD removal. Analysis of COD concentration in the overflow from each compartment however showed that, after a certain number of compartments, the added benefit in each additional compartment becomes progressively less (**Section** 4.4.2).

The *number of compartments* should be selected to be equal to or greater than the number of zones within the reactor that can develop microbial consortia with significantly different characteristics. Boopathy (1998) showed that for 4 ABRs with 2, 3, 4 and 5 compartments respectively, and with all other dimensions identical, more compartments resulted in better solids retention and overall greater extent of treatment for a swine manure feed. This implies that repeated passes through the sludge bed has a greater beneficial effect in increasing extent of treatment than maintaining a low upflow velocity, although Boopathy's findings were for constant flow-rate conditions. However the results from the pilot-scale ABR study reported here indicate that there must be a cross-over point where increasing the number of compartments will increase the upflow velocity to a point where washout of sludge occurs to the detriment of the biological processes, resulting in poorer COD removal performance than for a smaller number of compartments.

It is proposed that a system should be designed with at least three, but up to 5 compartments, to ensure good contact between wastewater flow and sludge beds, and to provide for development of an acid zone in the first one or two compartments.

The number of compartments may be increased to provide additional sludge storage volume to increase the desludging interval. However, if this is the purpose of the additional compartments, then the A-HRT should be increased accordingly, i.e.

$$\text{Design HRT} = \text{Reqd HRT} \cdot \frac{\text{Reqd. no. of compartments} + \text{additional storage compartments}}{\text{Reqd. no. of compartments}} \qquad \textbf{Eq. 6-7}$$

where the required HRT and number of compartments are those values that are understood to be necessary to achieve the required effluent specifications.

Appropriate hydraulic design, particularly the length of each compartment (distance between successive standing baffles) is important to ensure that wastewater is not able to bypass large portions of the sludge bed. *Reactor width to length ratio* does not have a direct effect on the superficial upflow velocity. However, a compartment that is too long will experience channelling and by-passing effects; more liquid flow will pass up through the sludge blanket near to the hanging baffle than near the following standing baffle, effectively by-passing much of the sludge bed and under-utilising reactor space.

6.6.3 Example of design parameters

Table 6.5 presents guidelines for selecting design parameters for an ABR using the findings of this study.

Table 6.5: Example of design parameters for an ABR treating domestic wastewater

Parameter	Symbol	Unit	Recommended parameter range or equation
Flow rate	Q	m^3/d	-
Hydraulic Retention Time	A-HRT	h	12[1] to 20 but 40 to 60 during start-up
Reactor working volume	V_W	m^3	Q×HRT/24
Peak upflow velocity	v_p	m/h	0.5
Design upflow velocity	v_d	m/h	$v_p/1.8 = 0.28$
Number of compartments	N	-	4 to 6
Hanging baffle clearance	d_h	m	0.15 to 0.20
Compartment upflow area	A_U	m^2	$Q/(v_D \times 24)$
Upflow to downflow area ratio	$R_{U:D}$	m^2/m^2	2 to 3
Compartment width to length ratio	$C_{W:L}$	m/m	3 to 4
Total compartment area	A_C	m^2	$A_U \times (1+R_{U:D})/R_{U:D}$
Reactor depth	r_D	m	1 to 3 (The reactor depth will largely be governed by the cost of excavation)
Reactor width	r_W	m	$\sqrt{\frac{V_W \cdot C_{W:L}}{N \cdot r_D}}$
Reactor length	r_L	m	$N \times r_W / C_{W:L}$

[1] See e.g. Lettinga (Lettinga, 2001) for justification of A-HRT values lower than attempted in this study

6.6.4 Alternative baffle design

This study has considered the performance of an ABR with a hanging and standing baffle design. However, a number of studies (e.g. Sasse, 1998; Garuti et al., 2001) have looked at baffled reactors where transport from one compartment to the next was achieved by downcomer pipes that fed the overflow from one compartment directly to the bottom of the next compartment. It is not possible to state whether one or the other design results in better performance of the ABR in wastewater treatment, although good performance of either system will depend on achieving a relatively uniform distribution of liquid flow at the bottom of each upflow section. Selection of either the classical hanging/standing baffle or downcomer pipe for compartment separation therefore will probably depend on the ease and cost of construction of each of the designs.

6.7 OPERATION OF BAFFLED DIGESTERS TREATING DOMESTIC WASTEWATER

Part of the first objective of this study (**Section** 1.4) was to investigate the performance of an ABR for treatment of domestic wastewater. The perceived application of the research was for rural or semi-rural communities with land available for treated effluent disposal, with some kind of community or municipal involvement for management of the system. From an operational point of view, the performance of the ABR is measured in terms of the outflow characteristics and the desludging requirements of the system.

6.7.1 Effluent characteristics

An ABR treating domestic wastewater will convert a large amount of wastewater COD to CH_4 gas, and will reduce pathogen loads in the wastewater. However, there is no nutrient removal, and the amount of pathogen removal obtained is insufficient to render the effluent safe for human contact. The presence of significant amounts of free and saline ammonia and phosphorus in the effluent mean that it cannot be discharged to surface or ground water, but theoretically could be used in irrigation of agricultural land, (provided metal concentrations are not too high) or disposed of in a soak-away. The pathogen indicator organism load measured in the pilot-scale ABR outflow indicates that secondary treatment is required before any conventional irrigation methods may be used, although subsurface irrigation could be acceptable.

Therefore, except in the case where sufficient area and infrastructure is available to build a sub-surface soak-away system, some post-treatment of the effluent is required before it can be reused. It has been recommended that the use of membranes in conjunction with the ABR be considered since an ultrafiltration membrane would remove virtually all COD and pathogens, while allowing nutrients, which have a real economic value as a fertiliser, to be retained for use in agriculture. Research in this area is continuing. Another post-treatment option is a constructed wetland.

6.7.2 Sludge build up rates

After a prolonged period of operation, changes in indicator measurements (especially increases in total solids and COD) would be observed in the outflow from the last compartment as the probability decreases that fluidisable solids from the last few compartments would be retained. At this point, solids would wash out of the digester. Initially, these solids would consist largely of inert particulate material and biomass, but eventually would also include significant amounts of biodegradable matter. At the point when indicator measurements in the effluent increase above acceptable levels, it will become necessary to desludge the ABR.

Development of an effective sludge management plan depends on understanding the following:

- How rapidly does sludge accumulate in the digester?
- What is the degree of stabilisation of the sludge that resides in the sludge beds and sludge blankets of the ABR?

From this information it would be possible to predict sludge accumulation rates, desludging intervals and to make recommendations for the management of the accumulated sludge after desludging.

6.7.2.1 Amount and condition of accumulated sludge

From this study, values for the amount of sludge accumulating in the upflow compartments of the ABR per amount of COD fed to the ABR were measured. If it is assumed that total sludge load is approximately 1.5 times the load residing in the upflow compartments, two values are proposed (**Table 6.1**):

- When high selection pressure results in poor digestion and the system operates as a sludge accumulator (Phase III), sludge accumulates at a rate of 0.43 kg dry solids/ kg applied COD (nearly 40% of influent COD). This sludge will contain significant amounts of biodegradable material since anaerobic digestion was not well established under the operating conditions.

- When stable anaerobic digestion is established and the system operates as a solids digester (Phase IV), a sludge accumulation rate of approximately 0.11 kg dry solids/ kg applied COD was observed. This sludge is expected to be fairly stable with low concentrations of residual SBCOD (except in the first compartment)

6.7.2.2 Desludging interval

From **Table 6.1** it was calculated that sludge accumulated in upflow compartments at a rate of 0.0064 m^3 settled sludge/kg COD applied in Phase III and 0.0017 m^3 settled sludge/kg COD applied. It is proposed that the maximum desludging interval should be calculated as the amount of time for the volume of settled sludge to equal one third of the total reactor working volume, i.e.

$$\text{Max desludging interval} = \frac{\frac{1}{3} \cdot \text{reactor volume}\,[\text{m}^3]}{\text{organic load}\,[\text{kgCOD/d}] \cdot \text{sludge accumulation rate}\,[\text{m}^3/\text{kgCOD}]}$$

Eq. 6-8

A value of ⅓ reactor volume is used since the fluidised sludge bed will be significantly greater than the settled sludge bed volume, and it is the fluidised bed volume that dictates when sludge overflow occurs.

At the sludge accumulation rates (**Table 6.1**) and OLR (**Table 5.2**) for Phase III and Phase IV, Eq. 6-8 predicts desludging intervals of 105 days for Phase III and 405 days for Phase IV. The advantage of operating the reactor such that the accumulated solids are effectively digested is clear; at half the organic load, the desludging interval is nearly 4 times longer for a fixed reactor design in Phase IV than in Phase III.

6.7.3 Monitoring requirements

The monitoring requirements for any wastewater treatment system depend on the purpose for which the information is required. These may be divided into two categories: (i) compliance; and (ii) diagnosis.

6.7.3.1 Compliance monitoring

Generally monitoring requirements for compliance purposes are specified in a licence or authorisation for the operation of a treatment system. These may include COD, suspended solids, nitrogen, phosphorus, pH, metals and pathogen indicator species. In South Africa, the major environmental contaminants of concern are nitrogen, phosphorus and pathogens, although other components are considered important in specific applications.

6.7.3.2 Diagnosis

There are three main objectives in diagnostic monitoring; i.e. to determine (i) what the condition of digestion in the system is; (ii) what alkalinity supplementation could be recommended; and (iii) what the desludging requirements of the system are.

Condition of digestion

- The condition of digestion depends on the prevailing pH and alkalinity values and the presence of active anaerobic micro-organisms. Simple measurements of pH and alkalinity in feed and effluent will indicate whether the reactor pH is sufficiently high that methanogenesis is not inhibited by low pH.

- Thereafter it is recommended that analyses of TKN, free and saline ammonia and COD are performed on the inflow and streams. COD provides the easiest measurement for identifying the fate of organic material in an anaerobic system; but the fate of COD is not easy to determine; it may leave the system in the outflow stream, as CH_4 gas, or as accumulated sludge. Measurements of COD associated with accumulated sludge are difficult to obtain and imprecise, while measurements of CH_4 production require that a gas seal is maintained on the ABR, and equipment for monitoring gas production and composition are available. Thus it is difficult to obtain an understanding of the relationship between accumulation and bioconversion of organic solids by looking at COD concentrations.

  Organically-bound nitrogen is associated with slowly biodegradable or particulate COD, and thus significant reductions in TKN between inflow and outflow stream indicate that significant quantities of particulate material are accumulating. However, if inflow and outflow TKN values are similar, and free and saline ammonia concentrations are significantly higher in the outflow than in the inflow, it may be inferred that particulate COD is being solubilised.

  Therefore, it is proposed that by monitoring the fate of both COD and nitrogen species in inflow and outflow streams, it is possible to understand what the predominant mechanism of COD removal is.

Alkalinity supplementation

- If it is desired that the ABR should achieve reasonable solids stabilisation rates and is operated at upflow velocities higher than 0.3 m/h, it is recommended that alkalinity supplementation be

employed to increase if the inflow stream alkalinity concentration to a value of 1 000 mg$CaCO_3$/ℓ, if the concentration is substantially less than this value. Speece (1996) describes the relative merits of different types of alkalinity supplementation.

Desludging requirements

- In order to monitor the amount of sludge accumulated in an ABR, it is recommended that core samples of upflow section of each compartment be obtained and the fluidised and settled sludge bed heights be obtained. The fluidised bed heights provide an indication of how *full* the reactor is, i.e. how soon it will have to be desludged, while the settled sludge bed heights indicate how much sludge will have to be removed.

6.8 ADVANTAGES AND DISADVANTAGES OF ABR TECHNOLOGY

The final objective of this study was to determine whether there was any advantage in the baffled design of the ABR over other anaerobic technologies used in the treatment of domestic wastewater (**Section** 1.4). This section presents a summary of the advantages and disadvantages identified and anticipated in the use of this technology. This study did not undertake a direct experimental comparison between different reactor designs; therefore the tabulated advantages and disadvantages are implied, rather than proven.

6.8.1 Advantages

Besides all the well-documented advantages of anaerobic technology over aerobic technology (lower energy input, lower operation costs, lower sludge production, CH_4 generation), the following specific advantages of the baffled design were identified:

- Good solids retention can be achieved due to the baffled design. Specifically, the first compartment retained a significant portion of the inflow solids, initially through settling, and subsequently through the development of a thick sludge bed that acted as a filter for incoming solids. This is a significant benefit in suspended solids management; Aiyuk (2006) recommended that suspended solids be removed from domestic wastewater before treatment in a UASB to preserve integrity of sludge granules. However, the first compartment of the ABR filters out solids, creating an up-front sludge digester, thereby reducing the need for pre-settling and separate treatment of settleable solids.

- The baffled design resulted in good contact between biomass and wastewater by forcing flow through sludge beds and sludge blankets (6.1.2.2 and 6.6.2.2). This advantage is amplified by the fact that flow is forced through a number of beds; multiple passes through beds decreases probability of that channelling and by-passing effects will result in slugs of fluid passing out of the reactor untreated. This is a significant benefit over the performance of most septic tanks, since most septic tank designs allow the bulk of the wastewater to pass from the inflow to the outflow without passing through a sludge bed (USEPA, 2002).

- Partial phase separation may have resulted in the development of zones in the reactor that were separated from and therefore protected from transient low pH values exerted due to acidogenesis in the first compartment. Therefore, although low pH conditions prevailed in compartment 1 and at times, compartment 2, pH values were usually slightly higher in later

compartments (**Section** 6.1.1) pH values in the effluent were similar to values predicted for digestion of domestic wastewater by a stoichiometric pseudo-steady state model, indicating that low pH values in the outflow were not caused by inhibition of methanogenesis.

- As there were more than 2 compartments, the acidogenic zone was able to increase to occupy more than just the first compartment as the total amount of sludge in the reactor built up, without compromising overall process stability (**Section** 6.1.1).

- There were indications that the nature of flow in the reactor (pseudo-plug-flow with solids retention) allowed rapid removal of sour liquors or toxicants after going sour, and that this may have assisted in rapid recovery (**Section** 5.6.3).

- The ABR is a sludge accumulating device, and therefore does not require continuous sludge removal. This has two advantages: firstly only infrequent solids handling is required; secondly, there is no requirement for continuous energy supply for sludge pumps as there would be in a non-accumulating UASB system.

6.8.2 Disadvantages

The ABR design has many of the same disadvantages as other anaerobic systems treating dilute particulate wastewater:

- low influent wastewater alkalinity concentration and low alkalinity generation potential of the wastewater result in poor pH buffering without alkalinity supplementation (**Section** 6.1.1.2)

- Without alkalinity supplementation, anaerobic digestion proceeds at low metabolic rates resulting in low biomass generation rates and increased vulnerability to sludge washout.

- The low organic load in dilute wastewater results in low biomass generation rates and therefore long start-up times. This also contributes to increased vulnerability to sludge washout.

In addition, there are two specific disadvantages associated with the design of the ABR for this application:

- There is inefficient use of reactor volume during early days of operation where most biological activity takes place in the first few compartments (**Section** 4.4.2). (However, this is compensated by the long desludging intervals that can be allowed as the remainder of the reactor space becomes active through sludge accumulation.) Thus the rate of treatment will be lower per unit volume than systems with continuous sludge removal.

- There is a higher capital cost associated with the ABR than with a standard septic tank. The cost of a multi-compartment digester would be higher than in a single-stage digester.

7 CONCLUSIONS AND RECOMMENDATIONS

"It is easier to get into something than to get out of it." – Donald Rumsfeld

Given the heterogeneous nature of domestic wastewater, investigations on domestic wastewater treatment are best performed at pilot- or full- scale since laboratory scale experimentation is beset with difficulties related to supply and condition of the feed wastewater. Few studies have been published on the performance of ABR technology at pilot- or full- scale. Thus there was a gap in the understanding of how the technology would perform on a feed of real wastewater.

An 8-compartment, 3 000 ℓ pilot-scale ABR was built and operated at two municipal wastewater treatment plants, Umbilo and Kingsburgh WWTP over a period of 5 years and chemical and microbiological data were collected on samples from inflow and outflow streams and from within compartments of the pilot-scale ABR. This thesis presents an analysis of these data. While pilot-scale research is less controlled and therefore more difficult to undertake and understand, these disadvantages were compensated for by obtaining information on the function of such a system on real domestic wastewater, including all those features that one would rather not have to contend with!

Research into anaerobic baffled reactor (ABR) technology for treating domestic wastewater was undertaken to achieve the following objectives (**Section** 1.4):

- To investigate the performance of a pilot-scale ABR in the treatment of wastewater of domestic origin and understand the mechanisms of treatment therein
- To identify the critical parameters in the design of an ABR sanitation system
- To determine whether the baffled design has any significant benefits over other anaerobic technologies in the treatment of domestic wastewater
- To develop a dynamic mathematical model of the biochemical processes in an ABR treating domestic wastewater

It was hypothesised that (i) phase separation in an ABR treating sewage is a benefit of the design over a single phase system, by allowing development of acidogenic and methanogenic zones in the ABR; and (ii) the critical parameter controlling effluent quality and sludge digestion rates in an ABR treating sewage was the applied hydraulic retention time (A-HRT) and that low effluent COD concentrations could be achieved at an A-HRT of 20 h.

This chapter presents conclusions and recommendations that have arisen from this research.

7.1 CONCLUSIONS

This section presents general conclusions made in the course of the research about mechanisms and observations related to the performance of the pilot-scale ABR treating domestic wastewater and more

generally about anaerobic digestion of domestic wastewater. Conclusions about the project objectives and hypotheses are also presented.

7.1.1 General conclusions: observations and mechanisms

In general, it was found that the pilot-scale ABR used in this study did not perform well relative to other anaerobic systems treating domestic wastewater reported in the literature. This was attributed variously to poor design resulting in high up-flow velocities at low applied hydraulic retention times and to the fact that surface waters and potable water in eThekwini Municipality have low inherent alkalinity, resulting in low pH buffering capacity, low reactor pH values and inhibition of anaerobic processes. In the sections that follow, conclusions relating to observations and mechanisms discussed in this thesis are presented.

7.1.1.1 Understanding the fate of biodegradable organic matter in the ABR

It was found that differences between COD concentration in inflow and outflow streams alone did not provide useful information on digestion rates occurring in the ABR (particularly during start-up) since COD removal may have been due to solids retention, or microbiologically mediated digestion. It was necessary to undertake COD mass balances to estimate digestion rates (e.g. amount of methane produced). It was also found that free and saline ammonia and alkalinity concentrations in the effluent provided an indication of the extent of digestion that occurred since both these quantities increased as a result of anaerobic digestion (**Section** 4.5.2).

7.1.1.2 Treatment mechanism during start-up

The predominant mechanism of COD removal during start-up was solids retention, although some digestion occurred (**Section** 4.5.2). Significant digestion (inferred from increases in alkalinity and free and saline ammonia concentration) was observed when sludge beds were well established (**Sections** 4.3.4 and 4.4.4). Sludge beds were seen to accumulate faster at higher OLR (lower HRT) during start-up (**Section 4.4.1** and 4.5.2).

During Phase I, most COD removal was observed to occur in the first compartment. Little difference in COD concentration in the overflow between subsequent compartments was observed after the first compartment for the start-up period. During the 20 h T-HRT period, when sludge beds had accumulated to a measurable extent in all compartments, the relative contribution to COD removal of compartments subsequent to the first increased; however, compartment 1 still effected greatest removals (**Section** 4.4.2). Two conclusions are drawn: firstly, COD removal is associated with sludge beds; secondly, much of the reactor volume does not contribute significantly to the treatment of the wastewater during start-up and soon thereafter; however, as sludge accumulates, reducing the available reactor volume in the first compartments, activity can shift to later compartments, thereby extending the desludging period.

7.1.1.3 pH values

Low pH values (almost always <7.0 and usually <6.5) were observed in compartments and outflow streams of the pilot-scale ABR. It was inferred that methanogenic micro-organisms experienced inhibition of metabolic processes as a result of low pH values (**Section** 4.4.3)

pH values were observed to be significantly lower in the first compartment than in the feed. Generally, pH values increased with compartment number (i.e. pH values at the outflow end of the reactor were higher than in compartment 1). Although there was some increase in pH values between

compartment 1 and 2, the increase was not sufficient to support the claim that phase separation occurred in the pilot-scale ABR where acidogenic reactions dominated in the first compartment, and methanogenic processes dominated in later compartments. It was clear that hydrolytic and acidogenic processes occurred in all compartments. However, spatial separation of compartments allowed regions characterised by slightly different pH values to develop, and this may have improved overall digestion rates and process stability (**Section** 6.1.1).

7.1.1.4 Solids accumulation rate vs. feed flow rate

Solids were observed to accumulate faster at an A-HRT of 22 h than at an A-HRT of 42 h. The solids accumulation rate per kg COD applied was also higher in the 22 h A-HRT period. Mass balance calculations indicated that approximately 30% of influent COD was removed as CH_4 at the higher flow rate, while 60% was removed as CH_4 at the lower flow rate. In addition, the free and saline ammonia concentration increase between inflow and outflow was greater at the lower flow rate. These data indicate that better digestion, termed *extent of treatment* of influent COD occurred at lower flow rates. A study using scanning electron microscopy confirmed this conclusion in that it observed low micro-organism concentrations, poor microbial diversity and few acetoclastic methanogens at higher flow rates, and conversely, good granulation, good microbial diversity, and many methanogens, including acetoclastic morphotypes at the lower flow rate. It was concluded that at the higher flow rate, the washout rate of micro-organisms was of similar magnitude to their generation rate, and thus diverse and stable microbial communities failed to establish (**Section** 6.1.2). It was concluded that liquid upflow velocity was an important factor for ensuring microbial stability in an ABR treating sewage, sludge accumulation rates, and thus ultimately the required desludging interval.

It was further concluded that, since microbial respiration rates were limited by low pH values and low substrate concentrations, the critical upflow velocity is not a global value, but system specific and dependent on prevailing pH conditions determined by alkalinity concentration and organic strength of the wastewater to be treated.

7.1.1.5 Sludge production rates and condition of accumulated solids

It was estimated that sludge accumulated at a rate of 0.43 kg dry solids/ kg applied COD at an A-HRT of 22 h (Phase III) and at 0.11 kg dry solids/ kg applied COD at an A-HRT of 42 h (Phase IV). These values corresponded to desludging intervals of 105 d and 405 d of uninterrupted operation respectively (**Section** 6.2.1).

It was concluded that at the higher flow rate, accumulated solids contained significant amounts of undegraded particulate organic material since this was not recovered in the effluent. It was concluded that the main function of ABR was a solids accumulator at high flow rates. However, at low flow rates, approximately 80% of incoming COD was removed, while it was calculated that approximately 60% of inflow COD was converted to CH_4. Therefore, the ABR behaved as a solids digester at these flow rates, and it is expected that the residual biodegradability of the accumulated sludge was far lower at the lower flow rate than at the higher flow rate. This proposal was supported by the nitrogen balance, which indicated that little of the influent TKN was retained as sludge (**Section** 6.2.2). The residual biodegradability of the accumulated sludge will have an impact on how removed sludge may be disposed of.

7.1.1.6 Effect of sludge age on biomass retention

It was shown that sludge age or solids retention time (SRT) is not a fixed value for an accumulating system. It is possible to have poor biomass retention, but high sludge accumulation rates if the net rate of accumulation (in – out + generation – degradation) of biodegradable solids is larger than the rate of accumulation of the micro-organisms that should degrade them (**Section** 6.3).

7.1.1.7 Effluent stream composition

A steady-state model with a sludge retention factor was used to predict the effluent composition of an anaerobic system treating domestic sewage. It was calculated that outflow pH values of between 5.9 and 6.1 could be expected for a system with no alkalinity supplementation. Increases in free and saline ammonia and alkalinity were predicted, and predicted values were found to be similar to those obtained experimentally (**Section** 6.4.1).

7.1.1.8 Diurnal variation in feed concentration

A sampling campaign was undertaken to determine the extent of diurnal variations in feed condition on the pilot-scale ABR. The pilot-scale ABR was monitored hourly for 24 hours. Significant diurnal variation of inflow COD and alkalinity concentrations was observed. The regular sampling time for normal operation (i.e. not during this sampling campaign) coincided with higher inflow COD concentrations than the daily average. Thus inflow sample measurements for regular operation may have overestimated the average inflow COD concentration to the pilot-scale ABR by up to 100 mgCOD/ℓ (**Section** 5.5.3). Additional data are required to support this observation.

7.1.1.9 Attenuation of diurnal concentration variations in outflow stream

The outflow COD and alkalinity concentration profiles showed less variation around the mean value than the inflow profile. Thus the pilot-scale ABR played some role in damping oscillations in determinand concentrations due to diurnal oscillations of feed strength. (**Section** 5.5.3)

7.1.1.10 Alkalinity supplementation

It was calculated that dosing digester feed with a source of alkalinity such as lime or Na_2CO_3 such that the feed alkalinity concentration increased to around 1 000 mg$CaCO_3$/ℓ is required to maintain the digester pH above a value of 6.5 (**Section** 6.4.5).

7.1.2 Conclusions relating to objectives and hypotheses of this study

This section specifically addresses the project objectives and hypotheses.

7.1.2.1 Phase separation

It was hypothesised that phase separation in an ABR treating sewage is a benefit of the design over a single phase system, since it allows development of acidogenic and methanogenic zones in the ABR. Examination of pH value and soluble COD concentration profiles across the compartments of the ABR showed that compartment 1, and at times compartment 2 experienced lower pH values than subsequent compartments, while significantly higher soluble COD concentrations were experienced in compartment 1 only. In addition, no observations of acetoclastic methanogens Methanosaeta were observed in compartment 1, but during Phase IV operation, these Archaea were observed in later compartments. These results suggest that a degree of phase separation did occur in the ABR, but that it was not complete phase separation since hydrolytic and acidogenic processes occurred and indeed were rate limiting in later compartments (**Section** 6.1.1.3).

Nevertheless, it is concluded that the compartmentalised structure of the ABR had advantages over UASB reactors and septic tanks in the treatment of domestic wastewater since:

- Solids retention in compartment 1 reduces the requirement for pre-settling and separate digestion of wastewater solids
- Sludge beds in later compartments were protected from transient high organic loads through solids retention in the first compartment
- The initial acidogenic zone was not contained exclusively to the first compartment when after a long period of continuous operation, undigested particulate organics overflowed to the second compartment, without any detrimental effects on the outflow stream (**Section** 6.8.1).

7.1.2.2 A-HRT as the critical design parameter

It was hypothesised that the critical parameter controlling effluent quality and sludge digestion rates in an ABR treating sewage was the applied hydraulic retention time (A-HRT) and that low effluent COD concentrations could be achieved at an A-HRT of 20 h.

It was found that the ability of the system to retain active biomass depended on the upflow velocity in the reactor and that this controlled the quality of the effluent and the rate of sludge accumulation. Although this was dependent on the feed flow rate, and therefore the A-HRT, the relationship between A-HRT and upflow velocity is dependent on reactor geometry and therefore would be different in another system. Therefore it was concluded that both A-HRT and upflow velocity should be regarded as critical design parameters for design of an ABR treating domestic wastewater. (**Sections** 6.1.2**,** 6.1.4.1 and 6.6.2.1).

It was observed that at a 22 h A-HRT, a well-balanced and stable anaerobic biomass was not able to establish in the pilot-scale ABR. However, it was concluded that this was due to the upflow velocity being too high, and was not specifically due to the low A-HRT value. It was proposed that stable digestion and acceptable effluent quality at A-HRT values of 20 h and lower could be achieved, but only with reactor geometries that resulted in lower upflow velocities.

7.1.2.3 Performance of ABR treating domestic wastewater and mechanisms of treatment

The first objective of this study was to investigate the performance of a pilot-scale ABR in the treatment of wastewater of domestic origin and understand the mechanisms of treatment therein.

Principle mechanisms of treatment in an ABR were shown to be through solids retention and anaerobic digestion of retained solids. It was shown that the baffled design assisted in solids retention and ensured good treatment of soluble components from the influent as a result of filtration through and contact with sludge beds in the compartments.

It was observed that significant COD reduction could be achieved, and that the extent of treatment depended on the upflow velocity and A-HRT of the system. Free and saline ammonia and alkalinity concentrations were observed to increase as a result of treatment in the ABR.

It was also observed that although treatment in the ABR resulted in significant removal of pathogen indicator organisms, especially of helminth eggs, the load of pathogens in the outflow stream was nevertheless too high for the effluent to be discharged to a water course or to surface irrigation.

7.1.2.4 Critical design parameters

The second objective of the study was to identify critical design parameters for an ABR treating domestic wastewater. These were shown to be upflow velocity, and A-HRT. The number of compartments was also shown to be important in ensuring low upflow velocities and good treatment rates

7.1.2.5 Benefits of ABR technology over other anaerobic technologies treating domestic wastewater

Finally, the research proposed to determine whether the baffled design has any significant benefits over other anaerobic technologies in the treatment of domestic wastewater.

It was concluded that the ABR would provide a greater extent of treatment and more consistent effluent characteristics than a septic tank. It was also found that a number of advantages over a single UASB system accrued through development of multiple sludge beds and partial phase separation. However, it was indicated that the baffled design and accumulating nature of the ABR meant that the rate of treatment was lower per unit volume than systems with continuous sludge removal. A further disadvantage was that the capital cost of a multi-compartment system would be greater than for a single stage system.

7.1.2.6 Development of a dynamic model of the biochemical processes in an ABR

Mass balance and steady state modelling exercises were able to tease out information pertinent to design from the experimental data. Dynamic simulation of processes in the ABR was undertaken using a Siegrist model of anaerobic digestion in WEST software. This model was able to describe alkalinity and ammonia generation for appropriate feed characterisation, but was unable to accurately simulate pH data and soluble COD data because of limitations of the model structure and in the available data. It was concluded that future modelling studies should be undertaken using the biochemical process model described by the Anaerobic Digestion Model No. 1 and additional experimental data should be obtained to improve the model's predictive abilities. Additional data required include feed characterisation data and a model of the relationship between upflow velocity and sludge settling characteristics.

7.2 RECOMMENDATIONS

Recommendations arising from the project fall into three categories: (i) recommendations relating to the design of anaerobic treatment systems for domestic wastewater using ABR technology; (ii) recommendations for monitoring an ABR; and (iii) recommendations for future research.

7.2.1 Design

It was recommended that a peak upflow velocity of 0.5 m/h be employed for a medium strength wastewater with low alkalinity concentration, such as obtained in eThekwini Municipality, if alkalinity supplementation is not employed, and if no additional sludge retention devices are included in the design.

It was proposed that an A-HRT of 20 h could be applied if the peak upflow velocity did not exceed 0.5 m/h. Additional reactor volume may be included to extend the desludging interval and thereby reduce maintenance costs.

It is proposed that a system should be designed with at least three, but up to 5 compartments, to ensure good contact between wastewater flow and sludge beds, and to provide for development of an acid zone in the first one or two compartments.

7.2.2 Monitoring

It was proposed that the following properties be measured in inflow and outflow streams in order to monitor the condition of digestion in an ABR: COD, alkalinity, pH, TKN and free and saline ammonia. Additionally, core samples of the upflow section of each compartment can be taken and the fluidised bed height and settled bed height recorded to monitor the rate of accumulation of solids in the ABR, and to determine desludging requirements. (**Section** 6.7.3)

7.2.3 Future research

The primary conclusion of all research is that more research needs to be done. A number of specific recommendations for future research came out of this study:

7.2.3.1 Decoupling OLR, alkalinity concentration and upflow velocity

This study was limited by the fact that the feed concentration was fixed at the value of the WWTP influent wastewater. Therefore, it was not possible to investigate the performance of the system for different wastewater strengths. Specifically, it was not possible to investigate the effects of OLR and upflow velocity on sludge accumulation characteristics independently since for this study, the two parameters were linearly related. Similarly, the inflow alkalinity concentration was not adjusted by alkalinity supplementation during the study, and therefore the relationship between upflow velocity and sludge accumulation characteristics at different feed concentrations was not investigated. Thus it is recommended that future research is designed in such a way that these three parameters may be decoupled by adjusting feed wastewater strength (e.g. by gravity concentration of wastewater) and alkalinity concentration.

7.2.3.2 Alkalinity supplementation

It was concluded that alkalinity supplementation that increased the feed concentration to around 1 000 mg$CaCO_3$ would ensure that digester pH values were maintained above a value of 6.5. Under these conditions, inhibition of methanogenic processes is likely to be low, and thus more stable micro-organism populations should be able to develop at a fixed upflow velocity. The corollary is that for digester pH values above 6.5, higher upflow velocities could be employed without washing out slower growing micro-organisms than if no alkalinity supplementation was practiced. These recommendations need to be experimentally verified.

7.2.3.3 Enhancing solids retention

It was proposed that addition of packing media in the upflow compartments of the ABR could improve solids retention at a particular upflow velocity and thus enhance solids digestion rates. This needs to be demonstrated and optimised for ABR technology.

REFERENCES

Aiyuk S., Amoake J., Raskin L., Van Haandel A. and Verstaete W. (2004). Removal of carbon and nutrients from domestic sewage using a low-cost integrated treatment concept. *Water Research,* **38** 3031-3042 (*in:* Aiyuk *et al.*, 2006)

Aiyuk S., Forrez I., Lieven D. K., Haandel A. v. and Verstraete W. (2006). Anaerobic complementary treatment of domestic sewage in regions with hot climates-A review. *Bioresource Technology,* **97**(17) 2225-2241

Akunna J. C. and Clark M. (1999). Performance of granular-bed anaerobic baffled reactor (GRABBR) treating whisky distillery wastewater. *Bioresource Technology,* **74** 257-261

Akunna J. C. and Clark M. (2000). Performance of granular-bed anaerobic baffled reactor (GRABBR) treating whisky distillery wastewater. *Bioresource Technology,* **74** 257-261

Anderson G. K., Kasapgil B. and Ince O. (1994). Microbiological study of two stage anaerobic digestion start-up. *Water Research,* **28** 2383–2392

APHA (1998). Standard methods for the examination of water and wastewater.

Ayres R. M. and Mara D. D. (1996). Analysis of wastewater for use in agriculture – a laboratory manual for parasitological and bacteriological techniques.

Bachmann A., Beard V. L. and McCarty P. L. (1985). Performance characteristics of the anaerobic baffled reactor. *Water Research,* **19**(1) 99-106

Bailey J. E. and Ollis D. F. (1986). *Biochemical Engineering Fundamentals*. 2nd edn. McGraw-Hill Book Company New York.

Baloch M. I. and Akunna J. C. (2003). Granular Bed Baffled Reactor (Grabbr): Solution to a Two-Phase Anaerobic Digestion System. *Journal of Environmental Engineering,* **129** (11) 1015-1021

Barber W. P. and Stuckey D. C. (1998). Influence of start-up strategies on the performance of an anaerobic baffled reactor. *Environmental Technology,* (19) 489-501

Barber W. P. and Stuckey D. C. (1999). The use of the anaerobic baffled reactor (ABR) for wastewater treatment: A review. *Water Research,* **33**(7) 1559-1578

Barber W. P. and Stuckey D. C. (2000a). Nitrogen removal in a modified anaerobic baffled reactor (ABR): 1, Denitrification. *Water Research,* **34**(9) 2413-2422

Barber W. P. and Stuckey D. C. (2000b). Nitrogen removal in a modified anaerobic baffled reactor (ABR): 2, Nitrification. *Water Research,* **34**(9) 2423-2432

Barbosa R. A. and Sant'Anna G. L. J. (1989). Treatment of raw domestic sewage in an UASB reactor. *Water Research,* **23**(12) 1483-1489 (*in* Seghezzo *et al.*, 1998)

Barker D. J., Mannucchi G. A., Salvi S. M. L. and Stuckey D. C. (1999). Characterisation of soluble residual chemical oxygen demand (COD) in anaerobic wastewater treatment effluents. *Water Research,* **33**(11) 2499-2510

Barker D. J. and Stuckey D. C. (1999). A review of soluble microbial products (SMP) in wastewater treatment systems. *Water Research,* **33**(14) 3063-3082

Batstone D. J., Keller J., Angelidaki I., Kalyuzhnyi S. V., Pavlostathis S. G., Rozzi A., Sanders W. T. M., Siegrist H. and Vavilin V. A. (2002). *Anaerobic Digestion Model No.1*. Report No. Scientific and technical report no. 13 London.

Batstone D. J., Keller J. and Steyer J. (2005). A review of ADM1 extensions, applications and analysis 2002-2005. *Proceedings: The first international workshop on the IWA Anaerobic Digestion Model No. 1 (ADM1),* Lyngby - Denmark,

Batstone D. J., Picioreanu C. and Loosdrecht M. C. M. v. (2006). Multidimensional modelling to investigate interspecies hydrogen transfer in anaerobic biofilms. *Water Research,* **40** 3099 – 3108

Behling E., Diaz A., Colina G., Herrera M., Gutierrez E., Chacin E., Fernandez N. and Forster C. F. (1997). Domestic wastewater treatment using a UASB reactor. *Bioresource Technology,* **61** 239-245

Bell J. (2002). *Treatment of dye wastewaters in the anaerobic baffled reactor and characterisation of the associated microbial populations*. PhD Thesis. Chemical Engineering. University of Natal. Durban, South Africa.

Bell J. and Buckley C. A. (2003). Treatment of a textile dye in the anaerobic baffled reactor. *Water SA,* **29**(2) 129-134

Bell J., Dama P., Mudunge R. and Buckley C. (2007). *Treatment of high-strength and toxic organic industrial effluents in the anaerobic baffled reactor*. Water Research Commission Report No. 853/1/07 Pretoria.

Bodík I., Kratochvíl K., Gašpariková E. and Hutňan M. (2003). Nitrogen removal in an anaerobic baffled filter reactor with aerobic post-treatment. *Bioresource Technology,* **86** 79-84

Bolle W. L., Breugel J. v., Eybergen G. C. v., Kossen N. W. F. and Zoetemeyer R. J. (1986). Modeling the liquid flow in up-flow anaerobic sludge blanket reactors. *Biotechnology and Bioengineering,* **28**(11) 1615-1620

Boopathy R. (1998). Biological treatment of swine waste using anaerobic baffled reactor. *Bioresource Technology,* **64** 1-6

BORDA (2008). Decentralized Waste Water Treatment - DEWATS. http://www.borda-net.org/modules/cjaycontent/index.php?id=29 (accessed 08/12/07)

Brownlee K. A. (1966). *Statistical Theory and Methodology in Science and Engineering*. 2nd edn. John Wiley & Sons, Inc. New York.

Cohen A., Breure A. M., Van Andel J. G. and Van Deursen A. (1980). Influence of phase separation on the anaerobic digestion of glucose: maximum COD-turnover rate during continuous operation. *Water Research,* **14** 1439-1448

Dama P. ((in preparation)). *Pilot-scale study of an anaerobic baffled reactor for the treatment of domestic wastewater*. MSc Eng Thesis. Chemical Engineering. University of KwaZulu-Natal. Durban, South Africa.

Dama P., Bell J., Foxon K. M., Brouckaert C. J., Huang T., Buckley C. A., Naidoo V. and Stuckey D. (2002). Pilot-scale study of an anaerobic baffled reactor for the treatment of domestic wastewater. *Water Science & Technology,* **46**(9) 263-270

Dama P., Govender K., Huang T., Foxon K. M., Bell J., Brouckaert C. J., Buckley C. A., Naidoo V. and Stuckey D. (2001). Flow Patterns in an Anaerobic Baffled Reactor. *Proceedings: 9th World Congress on Anaerobic Digestion,* Antwerp, Belgium, (2-5 September 2001)

Danckwerts P. V. (1953). Continuous flow systems, distribution of residence times *Chemical Engineering Science*, **2**(1) 1-13

Davies O. L. and Goldsmith P. L. (1977). *Statistical methods in research and production*. 4th edn. Longman Group Limited London.

DEFRA (2007). UK Biomass Strategy 2007. Working paper 3 - Anaerobic Digestion. Department for Environment, Food and Rural Affairs http://www.defra.gov.uk/environment/climatechange/uk/energy/renewablefuel/pdf/ukbio0507-work3.pdf (accessed 25/09/2008)

Dreywood R. (1946). Qualitative test for carbohydrate material. *Industrial Engineering Chemistry,* **18** 499

DWAF (1999). General Authorisations in terms of section 39 of the National Water Act, 1998 (Act no. 36 of 1998). *Government Gazette no. 20526*

DWAF (2008). Resource Quality Services water quality monitoring sites grouped by water management area (excluding borehole data). http://www.dwaf.gov.za/iwqs/wms/data/WMS_WMA_txt.htm (accessed 08/12/08)

Eastman J. A. and Ferguson J. F. (1981). Solubilisation of particulate organic carbon during the acid phase of anaerobic digestion. *Journal Water Pollution Control Federation,* **53** 352-366

Ekama G. A., Sötemann S. W., Ristow N. E. and Wentzel M. C. (2006). Biodegradability of activated sludge organics under anaerobic conditions. *Water Research*, **40**(19) 3587-3595

eThekwini Water Services (2008). Durban Heights WW Final 3 statistics, April - June 2008. http://www.durban.gov.za/durban/services/services/water_and_sanitation/water-quality/potable-water-statistics/DurbanHeights3AprilJune2008.pdf (accessed 2008/12/05)

Faisal M. and Unno H. (2001). Kinetic analysis of palm oil mill wastewater treatment by a modified anaerobic baffled reactor. *Biochemical Engineering Journal,* **9** 25-31

Feachem R. G., Bradley D. J., Garelick H. and Mara D. D. (1983). *Sanitation and Disease: Health Aspects of Excreta and Wastewater Management*. John Wiley and Sons Bath.

Foxon K. M., Brouckaert C. J., Remigi E. and Buckley C. A. (2004). Biochemical Modelling of the Anaerobic Baffled Reactor *Proceedings: WISA Biennial Conference and Exhibition,* Cape Town International Convention Centre, (2-6 May 2004)

Foxon K. M. and Buckley C. A. (2007). Analysis of a pilot-scale anaerobic baffled reactor. *Proceedings: IWA 11th Congress on Anaerobic Digestion 2007,* Brisbane, (23 -27 September 2007)

Foxon K. M., Buckley C. A., Brouckaert C. J., Dama P., Mtembu Z., Rodda N., Smith M., Pillay S., Arjun N., Lalbahadur T. and Bux F. (2006). *Evaluation of the Anaerobic Baffled Reactor for Sanitation in Dense Peri-urban Settlements*. Report No. 1248/1/06 Pretoria.

Garuti G., Giordan A. and Pirozzi F. (2001). Full-scale ANANOX® system performance. *Water SA,* **27**(2) 189-197

Garuti G., Leo G. and Pirozzi F. (2004). Experiments and modelling on biomass transport inside upflow sludge blanket reactors intermittently fed. *Water SA,* **30**(1) 97-106

Goldstein (1999). The health implications of efficient water use in urban areas.

Gopala K. G. V. T. (2007). *Treatment of low strength wastewater using an anaerobic baffled reactor*. PhD Thesis. Civil Engineering. Indian Institute of Technology Roorkee. Roorkee.

Grobicki A. M. W. (1989). *Hydrodynamic characteristics and performance of the anaerobic baffled reactor*. Thesis. Department of Chemical Engineering. Imperial College. London, UK (*in*: Barber and Stuckey, 1999).

Grobicki A. M. W. and Stuckey D. C. (1992). Hydrodynamic characteristics of the anaerobic baffled reactor. *Water Research,* **26** 371-378

Gujer W. and Zehnder A. J. B. (1983). Conversion processes in anaerobic digestion. *Water Science & Technology,* **15** 127-167

Halalsheh M., Sawajneh Z., Zu'bi M., Zeeman G., Lier J., Fayyad M. and Lettinga G. (2005). Treatment of strong domestic sewage in 96 m^3 UASB reactor operated at ambient temperatures: Two-stage versus single-stage reactor. *Bioresource Technology,* **96** 577-585

Hanaki K., Matsuo T., Nagase M. and Tabata Y. (1987). Evaluation of effectiveness of two-phase anaerobic digestion process degrading complex substrate. *Water Science & Technology,* **19**(311)

Hanekom D. (2006). Address by Deputy Minister Derek Hanekom, at the "Closing the Sanitation Loop" seminar in Sweden (accessed 28/11/2007)

Heertjes P. M. and Vandermeer R. R. (1978). Dynamics of liquid flow in an up-flow reactor used for anaerobic treatment of wastewater. *Biotechnology and Bioengineering,* **20**(10) 1577-1594

Hendriks L. (2007). Budget Vote 34 of 2007/08 Speech by Mrs LB Hendricks, Minister of Water Affairs and Forestry. http://www.dwaf.gov.za/Communications/MinisterSpeeches/2007/Budget17May07.doc (accessed 28/11/2007)

Henze M. (1992). Characterisation of wastewater for modelling of activated sludge processes. *Water Science & Technology,* **25**(1) 1-15

Henze M., Grady C. P. L., Gujer W., Marias G. v. R. and Matsuo T. (1987). *Activated Sludge Model No.1*. Report No. 1 London.

Henze M., Harremoës P., Jansen J. and Arvin E. (1997). *Biological and chemical processes in wastewater treatment*. Springer Verlag Berlin, Germany.

Hulshoff Pol L. W., De Castro Lopes S. I., Lettinga G. and Lens P. N. L. (2004). Anaerobic sludge granulation. *Water Research,* **38** 1376-1389

Hulshoff Pol L. W., De Zeeuw W. J., Velzeboer C. T. M. and Lettinga G. (1983). Granulation in UASB reactors. *Water Science & Technology,* **15**(8/9) 291-304

Kalogo Y. (2001). *Reactor start-up and process design for the anaerobic treatment of domestic wastewater*. PhD Thesis. Ghent University. Belgium.

Kalogo Y. and Verstraete W. (2001). Potentials of anaerobic treatment of domestic sewage under temperate climate conditions. In: *Decentralised Sanitation and Reuse: Concepts, Systems and Implementation.* Lens P., et al. (Ed.) IWA publishing London.

Kalyuzhnyi S. V. and Fedorovich V. V. (1998). Mathematical modelling of competition between sulphate reduction and methanogenesis in anaerobic reactors. *Bioresource Technology,* **65** 227-242

Kato M. T., Field J. A. and Lettinga G. (1997). The anaerobic treatment of low strength wastewaters in UASB and EGSB reactors. *Wat. Sci. Tech,* **36**(6-7) 375-382

Kennedy K. and Barriault M. (2005). Effect of recycle on treatment of aircraft de-icing fluid in an anaerobic baffled reactor. *Water SA,* **31**(3) 377-384

Kreutzer G. (2008). BORDA BNS Network at the World Water Week in Stockholm. http://www.borda-net.org/modules/news/article.php?storyid=107 (accessed 08/12/07)

Lalbahadur T. (2005). *Characterisation of the microbial communities present in an anaerobic baffled reactor using molecular techniques*. MTech Thesis. Department of Biotechnology. Durban Institute of Technology. Durban.

Langenhoff A. A. M., Intrachandra N. and Stuckey D. C. (2000). Treatment of dilute soluble and colloidal wastewater using an anaerobic baffled reactor: Influence of hydraulic retention time. *Water Research,* **34**(4) 1307-1317

Langenhoff A. A. M. and Stuckey D. C. (2000). Treatment of dilute wastewater using an anaerobic baffled reactor: Effect of low temperature. *Water Research,* **34**(15) 3867-3875

Leitão R. C., Van Haandel A., Zeeman G. and Lettinga G. (2006). The effects of operational and environmental variations on anaerobic wastewater treatment systems: A review *Bioresource Technology,* **97**(9) 1105-1118

Lettinga G. (2001). Potentials of anaerobic pre-treatment (AnWT) of domestic sewage under tropical conditions. In: *Decentralised Sanitation and Reuse: Concepts, Systems and Implementation* Lens P., et al. (Ed.) IWA Publishing London.

Lettinga G., De Man A. W. A., Grin P. and L.W. H. P. (1987). Anaerobic waste water treatment as an appropriate technology for developing countries. *Trib. Cebedeau,* **40** 21-32 (*in* Seghezzo *et al.*, 1998)

Lettinga G., De Man A. W. A., Van der Last A. R. M., Wiegant W., Knippenberg K., Frijns J. and Van Buuren J. C. L. (1993). Anaerobic treatment of domestic sewage and wastewater. *Water Science & Technology,* **27** 67-73

Lettinga G., Roersma R. and Grin P. (1983). Anaerobic treatment of raw domestic sewage at ambient temperatures using a granular bed UASB reactor. *Biotechnology and Bioengineering,* **22** 699-734 (*in* Seghezzo *et al.*, 1998)

Lettinga G., Van Nelson A. F. M., Hobma S. W., De Zeeuw W. J. and Klapwijk A. (1980). Use of the upflow sludge blanket (USB) reactor concept for biological wastewater treatment, especially for anaerobic treatment. *Biotechnology and Bioengineering,* **22** 699-734

Levenspiel O. (1999). *Chemical reaction engineering*. 3rd edn. John Wiley & sons

Lin K.-C. and Yang Z. (1991). Technical review of the UASB process. *International Journal of Environmental Studies,* **39** 203-222

Liu Y., Xu H.-L., Yang S.-F. and Tay J.-H. (2003). Mechanisms and models for anaerobic granulation in upflow anaerobic sludge blanket reactor. *Water Research,* **37**(3) 661-673

Lowry O. H., Rosebrough N. J., Farr A. L. and Randall R. J. (1951). Protein measurement with the Folin phenol reagent. *J. Biol. Chem.* **193** 265-275

Lyberatos G. and Skiadas I. V. (1999). Modelling of anaerobic digestion-A review. *Global Nest. The Int. J.* **1**(2) 63-76

MacLeod F. A., Guiot S. R. and Costerton J. W. (1990). Layered structure of bacterial aggregates produced in an upflow anaerobic sludge bed and filter reactor. *Appl Microbiol Biotechnol,* **56**(6) 1598-1607

Madigan M. T., Martinko J. M. and Parker J. (1996). *Brock's biology of microorganisms*. 8th edn. Prentice-Hall

Mahmoud N., Zeeman G., Gijzen H. and Lettinga G. (2004). Anaerobic sewage treatment in one-stage UASB reactor and a combined UASB-Digester system. *Water Research,* **38** 2348-2358

Manariotis I. D., Grigoropoulos and Sotirios G. (2002). Low-Strength Wastewater Treatment Using an Anaerobic Baffled Reactor. *Water Environment Research,* **74**(2) 170-176

Mara D. D., Sleigh A. and Taylor K. (2000). PC based simplified sewer design. http://www.efm.leeds.ac.uk/CIVE/Sewerage/sewerage_index.html (accessed 2008/11/24)

Maximova N. and Dahl O. (2006). Environmental implications of aggregation phenomena: Current understanding. *Current Opinion in Colloid & Interface Science,* **11**(4) 246-266

McCarty P. L. (2006). Thermodynamic electron equivalents model for bacterial yield prediction: Modifications and comparative evaluations. *Biotechnology and Bioengineering,* **97**(2) 377-388

McCarty P. L. and Mosey F. E. (1991). Modelling of anaerobic digestion processes (a discussion of concepts). *Water Science & Technology,* **24**(8) 17-33

McCarty P. L. and Smith D. P. (1986). Anaerobic Wastewater Treatment. *Environ. Sci. Tech.,* **20** 1200-1206

Moosbrugger R. E., Wentzel M. C., Loewenthal R. E., Ekama G. A. and Marais G. v. R. (1992). Simple Titration Procedure to Determine H2CO3 Alkalinity and Short Chain Fatty Acids in Aqueous Solutions Containing Known Concentrations of Ammonium, Phosphate and Sulphide Weak Acid/Bases. Report No. W74 Pretoria.

Mösche M. and Jördening H.-J. (1999). Comparison of different models of substrate and product inhibition in anaerobic digestion. *Water Research,* **33**(11) 2545-2554

Mosey F. E. (1983). Mathematical modelling of the anaerobic digestion process: regulatory mechanisms for the formation of short-chain volatile acids from glucose. *Water Science & Technology,* **15** 209 - 232

Mtembu D. Z. (2005). *The anaerobic baffled reactor for sanitation in dense peri-urban settlements*. MScEng Thesis. Chemical Engineering. University of KwaZulu-Natal. Durban.

Nachaiyasit S. and Stuckey D. C. (1997a). Effect of low temperatures on the performance of an anaerobic baffled reactor (ABR). *J. Chem. Technol. Biotechnol.,* **69** 276-284

Nachaiyasit S. and Stuckey D. C. (1997b). The effect of shock loads on the perfomance of an anaerobic baffled recator (ABR). 1. Step changes in feed concentration at constant retention time. *Water Research,* **31**(11) 2737-2746

Nachaiyasit S. and Stuckey D. C. (1997c). The effect of shock loads on the performance of an anaerobic baffled reactor (ABR). 2. Step and transient hydraulic shocks at constant feed strength. *Water Research,* **31**(11) 2747 2754

Nagase M. and Matsuo T. (1982). Interactions between amino acid degrading bacteria and methanogenic bacteria in anaerobic digestion. *Biotechnology and Bioengineering,* **14** 2227-2239

Nauman E. B. and Collinge C. N. (1968a). Measurement of contact time distributions in gas fluidized beds. *Chemical Engineering Science,* **23**(11) 1317-1326

Nauman E. B. and Collinge C. N. (1968b). The theory of contact time distributions in gas fluidized beds. *Chemical Engineering Science,* **23**(11) 1309-1316

Orozco A. (1997). Pilot and Full-scale Anaerobic Treatment of Low-Strength Wastewaters at Sub-Optimal Temperature (15 °C) with a Hybrid Plug Flow Reactor. *Proceedings: 8th International Conference on Anaerobic Digestion,* Sendai Japan,

Pavlostathis S. G. and Giraldo-Gomez E. (1991). Kinetics of anaerobic treatment. *Water Science & Technology,* **24**(8) 35-59

Perry R. H. and Green D. W. (1997). *Perry's Chemical Engineers' Handbook* 7th edn. McGraw-Hill

Pillay S. (2006). *Microbiological studies of an anaerobic baffled reactor (ABR) treating domestic wastewater*. MSc Thesis. School of Biological and Conservation Sciences. University of KwaZulu-Natal. Durban.

Pohland F. G. and Ghosh P. K. (1971). Developments in anaerobic stabilisation of organic wastes: the two phase concept. *Environ. Lett.,* **1** 255-266

Polprasert C., Kemmadamrong P. and Tran F. T. (1992). Anaerobic baffle reactor (ABR) process for treating a slaughterhouse wastewater. *Environmental Technology,* **13** 857-865

Rajinikanth R., Ramirez I., Steyer J. P., Escudie R., Torrijos M., Mehrotra I. and Kumar P. (2008). Experimental and modeling investigations of a hybrid upflow anaerobic sludge-filter bed (UASFB) reactor. *Water Science & Technology,* **58**(1) 109-117

Raskin L., Stromley J., Rittmann B. and Stahl D. (1994). Group-Specific 16S rRNA Hybridization Probes to Describe Natural Communities of Methanogens. *Applied and Environmental Microbiology,* **60**(4) 1232-1240

Raunkjear K., Hvitved-Jacobsen T. and Nielsen P. H. (1994). Measurement of pools of protein, carbohydate and lipid in domestic wastewater. *Water Research,* **28**(2) 251-262

Remigi E. U. and Foxon K. M. (2004). Modelling of anaerobic digestion processes: a review. *Proceedings: Research and Innovation in Wastewater Treatment,* Milan – Bologna, (24 – 26 November 2004)

Rockstrom J., Axberg G. N., Falkenmark M., Lannerstad M., Rosemarin A., Caldwell I., Arvidson A. and Nordstrom M. (2005). Sustainable pathways to attain the millennium development goals. Assessing the key role of water, energy and sanitation, a guideline report. *Stockholm Environment Institute*

Rodríguez J., Lema J. M., Van Loosdrecht M. C. M. and Kleerebezem R. (2005). Variable stoichiometry with thermodynamic control in ADM1. *Proceedings: The first international workshop on the IWA Anaerobic digestion model no. 1,* Lyngby, Denmark,

Ruiz I., Soto M., Veiga M. C., Ligero P., Vega A. and Blazquez R. (1998). Perfomance of and biomass characterisation in a UASB reactor treating domestic waste water at ambient temperature. *Water SA,* **24**(3) 215-222

Sallis P. J. and Uyanik S. (2003). Granule development in split-feed anaerobic baffled reactor. *Bioresource Technology,* **89** 255-265

Sasse L. (1998). Decentralised wastewater treatment in developing countries. BORDA www.borda-net.org/modules/wfdownloads/uploads/062%20BORDA_Dewats-Handbook.pdf (accessed 08/12/08)

Sayed S. K. I. A. and Fergala M. A. A. (1995). Two stage UASB concept for treatment of domestic sewage including sludge stabilization process. *Water Science & Technology,* **32**(11) 55-63 (*in:* Seghezzo *et al.*, 1998)

Schellinkhout A. and Osorio E. (1994). Long-term experience with the UASB technology for sewage treatment on large scale. *Proceedings: 7th International Symposium on Anaerobic Digestion,* Cape Town, South Africa,

Seghezzo L., Guera R. G., Gonzalez S. M., Trupiano A. P., Figueroa M. E., Cuevas C. M., Zeeman G. and Lettinga G. (2001). Removal efficiency and methanogenic activity profiles in a pilot-scale UASB reactor treating settled sewage at moderate temperatures. *Proceedings: 9th World Congress of Anaerobic Digestion - Anaerobic Conversion for Sustainability,* Antwerp, Belgium, (2-6 September 2001)

Seghezzo L., Zeeman G., Lier J. B. v., Hamelers H. V. M. and Lettinga G. (1998). A review: The anaerobic treatment of sewage in UASB and EGSB reactors. *Bioresource Technology,* **65** 175-190

Setiadi T., Husaini and Djajadiningrat A. (1996). Palm oil mill effluent treatment by anaerobic baffled reactors: Recycle effects and biokinetic parameters. *Water Science & Technology,* **34**(11) 59-66

She Z., Zheng X., Yang B., Jin C. and Gao M. (2006). Granule development and performance in sucrose fed anaerobic baffled reactors. *Journal of Biotechnology,* **122**(2) 198-208

Siegrist H., Renggli D. and Gujer W. (1993). Mathematical modelling of anaerobic mesophilic sewage sludge treatment. . *Water Science & Technology,* **27**(2) 25-36

Siegrist H., Vogt D., Garcia-Heras J. L. and Gujer W. (2002). Mathematical model for meso-and thermophlic anaerobic sewage sludge digestion. *Environ. Sci. Technol.,* **36** 1113-1123

Skiadas I. V. and Lyberatos G. (1998). The periodic anaerobic baffled reactor. *Water Science & Technology,* **38**(8-9) 401-408

Smith J. M. and Van Ness H. C. (1987). *Introduction to chemical engineering thermodynamics*. 4 edn. McGraw-Hill, Inc. Singapore.

Smith L. C., Elliot D. J. and James A. (1993). Characteristics of mixing patterns in an anaerobic digester by means of tracer curve analysis. *Ecological Modelling,* **69** 267-285

Sötemann S. W., Ristow N. E., Wentzel M. C. and Ekama G. A. (2005). A steady state model for anaerobic digestion of sewage sludges *Water SA,* **31**(4) 511-528

Speece R. E. (1996). Anaerobic biotechnology for Industrial Wastewaters. Archae Nashville, TN.

StatsSA (2005). Achieving a better life for all: Progress between Census '96 and Census 2001. Report No. 03-02-16 (2001) Statistics South Africa http://www.statssa.gov.za/Publications/Report-03-02-16/Report-03-02-16.pdf (accessed 03/12/2007)

Stumm W. and Morgan J. J. (1996). *Aquatic Chemistry: Chemical equilibria and rates in natural waters*. 3rd edn. John Wiley & Sons Inc. New York.

Tare V., Ahammed M. and Jawed M. (1997). Biomethanation in domestic and industrial waste treatment - an Indian scenario. *Proceedings: 8th International Conference on Anaerobic Digestion,* Sendai, Japan,

Tiwari M. K., Guha S., Harendranath C. S. and Tripathi S. (2006). Influence of extrinsic factors on granulation in UASB reactor. *Appl Microbiol Biotechnol,* **71**(2) 145-154

Tordiffe E. A. W., Botha B. J. V. and Loock J. C. (1985). The relationship between the geology and the groundwater quality of the Great Fish River catchment north of Kommadagga *Water SA,* **11**(2) pp. 99-105

United Nations (2005). Millennium Development Project Report 2005. http://www.unmillenniumproject.org/reports/fullreport.htm (accessed 03/12/2007)

United Nations (2007a). Millennium Development Goals Indicators http://mdgs.un.org/unsd/mdg/Metadata.aspx?IndicatorId=32 (accessed 30/11/2007)

United Nations (2007b). Millennium Development Goals Report 2007. http://mdgs.un.org/unsd/mdg/Resources/Static/Products/Progress2007/UNSD_MDG_Report_2007e.pdf (accessed 30/11/2007)

United Nations Development Group (2003). Indicators for monitoring the Millennium Development Goals: Definitions, Rationale, Concept and Sources. http://mdgs.un.org/unsd/mdg/Resources/Attach/Indicators/HandbookEnglish.pdf (accessed 30/11/2007)

USEPA (2002). *Onsite wastewater treatment systems Manual*. Report No. EPA/625/R-00/008

Uyanik S., Sallis P. J. and Anderson G. K. (2001a). The effect of polymer addition on granulation in an anaerobic baffled reactor (ABR) part-I: Process perfomance. *Water Research,* **36**(4) 933-943

Uyanik S., Sallis P. J. and Anderson G. K. (2001b). The effect of polymer addition on granulation in an anaerobic baffled reactor (ABR). II: compartmentalization of bacterial population. *Water Research,* **36**(4) 944-955

Van der Last A. R. M. and Lettinga G. (1992). Anaerobic treatment of domestic sewage under moderate climatic (Dutch) conditions using upflow reactors at increased superficial velocities. *Water Science & Technology,* **25**(7) 167-178

van Lier J. B. and Lettinga G. (1999). Appropriate technologies for effective management of industrial and domestic waste waters: the decentralised approach. *Wat. Sci. Tech,* **40**(7) 171-183

Vavilin V. A., Rytov S. V., Lokshina L. Y., Rintala Y. and Lyberatos G. (2001). Simplified hydrolysis models for the optimal design of two-stage anaerobic digestion. *Water Research,* **35**(17) 4247-4251

Verstraete W. and Vandeviviere P. (1999). New and broader applications of anaerobic digestion. *Critical Reviews in Environmental Science and Technology,* **29**(2) 151-173

Vieira S. M. M., Carvalho J. L., Barijan F. P. O. and Rech C. M. (1994). Application of the UASB reactor in domestic sewage treatment in a small community at Sumare São Paulo State. *Water Science & Technology,* **30**(12) 203-210 (*in:* Seghezzo *et al.*, 1998)

Vieira S. M. M. and Garcia A. D. J. (1992). Sewage treatment by UASB reactor. Operation results and recommendations for design and utilization. *Water Science & Technology,* **25**(7) 143-157 (*in* Seghezzo *et al.*, 1998)

Vossoughi M., Shakeri M. and Alemzadeh I. (2003). Perfomance of anaerobic baffled reactor treating synthetic wastewater influenced by decreasing COD/SO_4 ratios. *Chemical Engineering and Processing,* **4** 811-816

Wanasen (2003). Upgrading conventional septic tanks by integrating in-tank baffles.

Wang J., Huang Y. and Zhao X. (2004). Perfomance and characteristics of an anaerobic baffled reactor. *Bioresource Technology,* **93**(2004) 205-208

Weiland P. and Rozzi A. (1991). Treatment of low-strength wastewaters necessarily occurs at low organic loading rates except at very high hydraulic loading rates. . *Water Research,* **24**(8) 257-277

Wentzel M. C., Ekama G. A. and Sötemann S. W. (2006). Mass balance-based plant-wide wastewater treatment plant models – Part 1: Biodegradability of wastewater organics under anaerobic conditions. *Water SA,* **32**(3) 269-275

Wentzel M. C., Mbewe A., Lakay M. T. and Ekama G. A. (1999). Batch test for characterisation of the carbonaceous materials in municipal wastewater. *Water SA,* **25**(3) 327-336

Wikipedia (2007). Siderophore. http://en.wikipedia.org/wiki/Siderophores (accessed 13/12/07)

Witthauer D. and Stuckey D. C. (1982). Laboratory studies on anaerobic processes to treat dilute organic wastewater in developing countries. Study by IRCWD EAWAG Dübendorf, Switzerland.

Wu M. M. and Hickey R. F. (1997). Dynamic model for UASB reactor including reactor hydraulics, reaction and diffusion. *Journal of Environmental Engineering,* **123**(3) 244-252

Yu H. and Anderson G. K. (1996). Perfomance of combined anaerobic reactor for municipal wastewater treatment at ambient temperature. *Resources, Conservation and Recycling,* **17** 259-271

Zeeman G. and Lettinga G. (1999). The role of anaerobic digestion of domestic sewage in closing the water and nutrient cycle at community level. *Wat. Sci. Tech,* **39**(5) 187-194

Zhou W., Imai T., Ukita M., Li F. and Yuasa A. (2007). Effect of loading rate on the granulation process and granular activity in a bench scale UASB reactor. *Bioresource Technology,* **98**(7) 1386–1392

Zhou W., Imai T., Ukita M., Sekine M. and Higuchi T. (2006). Triggering forces for anaerobic granulation in UASB reactors. *Process Biochemistry,* **41**(1) 36-43

Zwietering T. N. (1959). The degree of mixing in continuous flow systems. *Chemical Engineering Science,* **11** 1-15

APPENDIX A1

SOUTH AFRICAN DISCHARGE STANDARDS

The authorisation for discharge allows a person who owns or lawfully occupies property registered in the Deeds Office or lawfully occupies or uses land that is not registered or surveyed outside of certain listed sensitive areas may on that property or land discharge up to 2 000 kℓ of wastewater on any given day into a water resource that is not a listed water resource provided that the discharge does not alter the natural ambient water temperature of the receiving water by more than 3Cº.

The authorisation for irrigation allows a person who owns or lawfully occupies property registered in the Deeds Office or lawfully occupies or uses land that is not registered or surveyed outside of certain listed sensitive areas may on that property or land irrigate up to 500 kℓ of domestic wastewater on any given day.

Table A1. 1:General Authorisations in terms of section 39 of the National Water Act, 1998 (Act no. 36 of 1998) for discharge to watercourse, or for irrigation

Component	Units	Target (discharge)	Target (Irrigation)
COD	mgCOD/L	75	400
Ammonia	mgN/L	3	
Nitrate	mgN/L	15	
Phosphorus	mgP/L	10	
TSS	mgTSS/L	25	
pH		5.5 - 9.5	6 - 9
Faecal Coliforms	cfu / 100 mL	1,000	100,000

Target values are obtained from the General Authorisations (DWAF, 1999) for *discharge* of waste or water containing waste into a water resource through a pipe, canal, sewer or other conduit and *irrigation* of any land with waste or water containing waste generated through any industrial activity or by a waterworks.

APPENDIX A2

METHODS OF SAMPLING AND ANALYSIS

This project spanned five years of experimentation. During this time, many people were involved in sampling and analysis, and a number of different techniques were used.

1 SAMPLING

Samples of inlet and outlet concentrations were obtained from the feed splitter box and the outlet pipe just before treated effluent was discharged back to the wastewater channel. During experimentation at Umbilo WWTP, samples were obtained from the sample valves supplied on the side of the reactor. The initial 100 mℓ drawn from each valve was discarded and the subsequent volume collected and stored for analysis.

The relative amounts of sludge and liquid in each compartment were measured using a *sampling stick* or *core sampler* (**Figure A2.** 1). This consisted of a Perspex tube with a 50 mm internal diameter, calibrated for height in metres, and fitted with a rubber bung attached to a steel rod. The rubber bung and rod was loosened from the outer tube and dropped into the ABR via the 75 mm port on the top of the compartment to be sampled. The Perspex rod was then dropped over the steel rod to land on the bung, capturing a *core* sample that would be withdrawn from the reactor. Initial sludge and liquid levels were recorded. A 5 min settling time was allowed before *settled* sludge levels were measured.

Figure A2. 1: Core Sampler filled with compartment 1 sludge (left) and compartment 8 sludge and supernatant (right)

During the Kingsburgh experimentation, samples of compartment contents were not obtained from the valves on the side of the reactor as it was believed that conditions near the wall of the reactor did not represent bulk conditions. Compartment samples were obtained using the core sampler. Once sludge levels had been recorded, the core sampler was balanced in a bucket and the bung worked loose so that the core sample flushed out into the bucket. Bucket contents were vigorously stirred and a sample withdrawn for storage and analysis.

2 SAMPLE STORAGE AND PREPARATION

Wherever possible, samples were transported immediately to a laboratory for analysis. Samples were stored in a cold room at the University of KwaZulu-Natal (Temperature varying between 4 and 10 ºC) or a refrigerator at Durban Institute of Technology. Where appropriate, samples were coarse filtered through Whatman No. 1 filter paper and micro-filtered through 0.45 µm acetate filter cartridges on site to reduce biological activity during transport and storage. For VFA measurements, samples were acidified using concentrated HCl. Samples for unstable analytes were transported in a cooler box filled with ice or ice-bricks.

3 ANALYTICAL METHODS

Where possible all analyses were conducted according to Standard Methods (APHA, 1998).

3.1 COD

Influent and effluent total COD concentrations were measured by the open reflux method; filtered or soluble COD concentrations were obtained by filtering samples through 0.45µm acetate filters and using the titrimetric closed reflux COD method (APHA, 1998).

3.2 Alkalinity

Alkalinity was determined by potentiometric titration using HCl to an end-point pH value of 4.5 according to Standard Methods (APHA, 1998)

3.3 Volatile Fatty Acids

Two methods were employed to measure VFA in samples:

Method 1-HPLC: Small samples (5 mℓ) were obtained from the influent and compartments 1 to 4 inclusive, and filtered through 0.45µm acetate filter cartridges on-site. These samples were transported on ice. A sample volume of 1 mℓ was passed through solid phase extraction cation exchange cartridges to extract VFA, and eluted with a sodium carbonate solution. Pretreated samples were analysed using high performance liquid chromatography for acetic, propionic, butyric, iso-butyric, valeric and iso-valeric acids.

Method 2-Titrimetric: VFA were determined as acetic acid in samples that were analysed titrimetrically for alkalinity using a five point titration according to Moosbrugger (1992).

3.4 Sulphate

Sulphate measurements were obtained on influent and effluent samples spectrophotometrically.

3.5 Phosphate

Phosphate measurements were obtained on influent and effluent samples spectrophotometrically.

3.6 Enumeration of total coliforms and *Escherichia coli*

Total coliforms and *E. coli* were simultaneously determined by the membrane filtration technique according to Standard Methods (APHA, 1998). Coliforms were enumerated as colony forming units (cfu) per 100 mℓ.

Samples were diluted (1: 10 000) and 10 mℓ volumes were filtered through a gridded 0.45 µm membrane filter (Schleicher and Schuell). Sterile phosphate buffer dilutions were done as controls at the beginning and at the end of filtrations.

Filters were aseptically placed on Chromocult Coliform Agar (Merck), and incubated at 35ºC for 18 – 24 h.

E. coli colonies appeared as dark-blue to violet colonies and total coliforms appear as salmon to red colonies. The absence of growth in controls indicated the sterility of the dilution water and filtration apparatus.

3.7 Enumeration of coliphages

Virus identification and isolation is difficult and expensive, and beyond the scope of most laboratories. For this reason, coliphages are routinely used as viral indicators. This technique involves enumerating the bacteriophage of host culture *E. coli* (ATCC 13706) using the double layer technique. Bacteriophages cause lysis on a lawn of *E. coli* host cells, forming clear plaques and were enumerated as plaque forming units (pfu) per 100mℓ.

3.8 Enumeration of helminth eggs

This was limited to a single helminth genus, namely *Ascaris*. Raw wastewater (1ℓ) and effluent (10ℓ) was collected on a weekly basis and allowed to sediment for 18 h. The supernatant of samples were discarded and the remaining sediments were centrifuged at 1 000 g for 15 min. The centrifuged supernatant was discarded and the enumeration of parasite eggs realised according to the modified Bailenger method (Ayres and Mara, 1996).

3.9 Scanning Electron Microscopy

Sludge samples from each compartment of the pilot-scale ABR were obtained during stable operation and prepared for SEM. Each sample was centrifuged for 5 min and the supernatant removed. Samples were washed three times in 0.1 M phosphate buffer at pH 7.2. Washed samples were decanted and fixed in 10% paraformaldehyde in 0.1M phosphate buffer. Samples were fixed for 16 h, decanted and washed three times with 0.1M phosphate buffer, and post-fixed with 1% osmium tetraoxide for 1 h at room temperature. Fixed samples were then repeatedly rinsed with distilled water to remove excess fixative, and dehydrated in a graded alcohol series (25, 50, 75 and 100%) of 10 min each. Samples were placed on Nucleopore filters (0.20 µm) and further dehydrated in a critical point drier (CPD).

Fixed samples were mounted on aluminium stubs, and sputter-coated with gold. The SEM graphs were taken on a Cleo 1450 instrument.

4 STATISTICAL METHODS

This section presents a summary of statistical methods used in this research. These statistical calculations are based on standard statistical theory, and were drawn mostly from Davies and Davies and Goldsmith (1977) and Brownlee (1966).

4.1 Definitions

The following terms are used:

α	:	Significance level
t	:	t-statistic for $Ø_n$ degrees of freedom at a (1-2×α) confidence level for a two-tailed problem or (1-α) for a one-tailed problem
n	:	Number of observations
Ø	:	Number of degrees of freedom
$\bar{x}$	:	Mean value of measurements variable *x*
σ	:	Standard deviation of a population
s	:	estimated standard deviation of a population calculated from a sample
V	:	Estimate of the variance of a population calculated from a sample

4.2 Confidence limits for a mean

The 100(1-2α)% confidence limits (i.e. for 95% confidence interval, α = 0.025) for a limited data set where σ is approximated by s may be described by (Davies and Goldsmith, 1997 p. 59)

$$\bar{x} \pm t_{(n-1,1-\alpha/2)} \frac{s}{\sqrt{n}} \tag{1}$$

4.3 Difference between two means

For two independent data sets with means $\overline{x_1}$ and $\overline{x_2}$, sample standard deviations of s_1 and s_2, and number of observations n_1 and n_2, the average difference between the means is $\overline{x_1} - \overline{x_2}$ and the 95% confidence interval of the difference is calculated from (Davies and Goldsmith, 1977 p. 61

$$\left(\overline{x_1} - \overline{x_2}\right) \pm t_{(\phi,\alpha)} \sqrt{\frac{s_1^2}{n_1} + \frac{s_2^2}{n_2}} \tag{2}$$

Where Ø is calculated from

$$\frac{1}{\phi} = \frac{1}{\phi_1}\left[\frac{s_1^2/n_1}{s_1^2/n_1 + s_2^2/n_2}\right]^2 + \frac{1}{\phi_2}\left[\frac{s_2^2/n_1}{s_1^2/n_1 + s_2^2/n_2}\right]^2 \tag{3}$$

Microsoft Office Excel will perform a t-test to compare two sets of data. A t-test calculates the t-statistic for a hypothesis (e.g. that $\overline{x_1} - \overline{x_2}$ =0, called the null hypothesis) from (2). The significance of the t-statistic can be obtained from the t-distribution with the appropriate degrees of freedom and describes the probability that the hypothesis is true. Thus if P=0.05, then there is only 5% probability

that the means are the same, or conversely that at the 95% confidence level, the hypothesis that the means are not the same cannot be rejected.

4.4 Ratio of two means: Fieller's theorem

Fieller's theorem states that the confidence limit of a/b, the ratio of two means a and b each with variance V(a) and V(b) is (Davies and Goldsmith, 1977 p. 236):

$$\frac{\frac{a}{b} - \frac{t^2 C(a,b)}{b^2} \pm \frac{t}{b}\sqrt{V(a) - \frac{2a}{b}C(a,b) + \frac{a^2}{b^2}V(b) - \frac{t^2 V(b)}{b^2}\left(V(a) - \frac{C(a,b)^2}{V(b)}\right)}}{1 - \frac{t^2 V(b)}{b^2}} \tag{4}$$

Where t is the appropriate t-statistic at a significance level α and for $Ø_n$ degrees of freedom. C(a,b) is the co-variance of variables a and b. Where a and b are independently determined (i.e. calculated from different experimental data sets) then C(a,b) = 0 and (4) reduces to

$$\frac{\frac{a}{b} \pm \frac{t}{b}\sqrt{V(a) + \frac{a^2}{b^2}V(b) - \frac{t^2 V(b)\cdot V(a)}{b^2}}}{1 - \frac{t^2 V(b)}{b^2}} \tag{5}$$

4.5 Linear Regression

Linear regressions can easily be performed by a variety of Microsoft Office Excel functions. The method for calculating a least squares slope to describe a linear relationship between two data sets is presented here since several of the quantities calculated in the process of calculating the least squares slope are used in determining confidence intervals of the regression coefficients and of values predicted by the regression (Davies and Goldsmith, 1977 pp. 185-206). For the independent data points x_i (e.g. time), and dependent data points y_i (e.g. concentration data for times x_i):

S_{xx} is the sum of squares about the mean of the independent data set:

$$S_{xx} = \sum_i \left(x_i^2\right) - \frac{\left(\sum_i x_i\right)^2}{n} - \sum_i \left(x_i - \bar{x}\right)^2 \tag{6}$$

S_{yy} is the sum of squares about the mean of the dependent data set:

$$S_{yy} = \sum_i \left(y_i^2\right) - \frac{\left(\sum_i y_i\right)^2}{n} - \sum_i \left(y_i - \bar{y}\right)^2 \tag{7}$$

S_{xy} is the sum of the product about the means:

$$S_{xy} = \sum_i \left(x_i \cdot y_i\right) - \frac{\left(\sum_i x_i\right)\left(\sum_i y_i\right)}{n} -$$

(8)

The equation of the regression is given by

$$y = bx + c \tag{9}$$

Where

$$b = \frac{S_{xy}}{S_{xx}} \tag{10}$$

It can be shown that

$$c = \bar{y} - b\bar{x} \tag{10}$$

4.6 Variation of the slope b

It can be shown that the variance of the slope b is

$$V(b) = \frac{\sigma^2}{\sum_i (x_i - \bar{x})^2} = \frac{\sigma^2}{S_{xx}} \tag{11}$$

Therefore the confidence limits of b are

$$b \pm t_{(\phi,\alpha)} \cdot \frac{s}{S_{xx}} \qquad (\phi = n - 2) \tag{12}$$

4.7 Variation about the regression

Analysis of variance around the regression is calculated using an ANOVA table:

Source of Variation	Sum of Squares	D of F	Mean Square
Due to regression	$b^2 \cdot S_{xx}$	1	$\sigma^2 + b^2 \cdot S_{xx}$
About regression	$S_{yy} - b^2 \cdot S_{xx}$	n-2	σ^2
Total	S_{yy}	n-1	

So the variance about the regression is

$$\sigma^2 = \frac{S_{yy} - b^2 \cdot S_{xx}}{n - 2} \tag{13}$$

The significance of the regression, i.e. the probability that the data can be described by this regression is determined using an F-test, where the F-statistic is the ratio

$$f = \frac{(\text{mean square due to the regression})}{(\text{mean square about the regression})}$$

And the significance of the regression is determined from the F-distribution.

The goodness of fit is calculated as

$$\frac{\left(\text{Sum of squares due to the regression}\right)}{\left(\text{total sum of squares}\right)} \times 100\% = \frac{b^2 \cdot S_{xx}}{S_{yy}} \times 100\% \tag{14}$$

The confidence interval around the regression is a minimum at the mean of the data set since this is where most information is known.

For an independent x value x_0, the regression predicts a value y_0,. The variance of the estimate y_0 is

$$V\left(y_0\right) = \sigma^2 \left[\frac{1}{n} + \frac{\left(x_0 - \bar{x}\right)^2}{\sum_i \left(x_i - \bar{x}\right)^2}\right] \tag{15}$$

This equation applies when the measured *x* values have negligible error, which is the often case when the x data set represents time.

The confidence limits for the value y_0 predicted by the regression are (Brownlee, 1966 p. 342)

$$y_0 \pm t_{(\alpha, n-2)} \cdot \sigma^2 \left[\frac{1}{n} + \frac{\left(x_0 - \bar{x}\right)^2}{S_{xx}}\right] \tag{16}$$

From (16) it is observed that the confidence limits change with x_0 and are smallest around the data mean $\bar{x}$.

This gives a typical curved shape to the confidence region of the regression prediction as shown in e.g. Figure 5.17.

4.8 Confidence interval of the ratio of two slopes

The confidence interval on the ratio of two slopes may be calculated from Fieller's theorem using the two slopes b_1 and b_2 and their variances $V(b_1)$ and $V(b_2)$ (Davies and Goldsmith, 1977 p. 236. The number of degrees of freedom in this case is $\sum_i \left(n_i - 2k\right)$ for k data sets (k = 2) since 2 degrees of freedom have been used in each regression (Davies and Goldsmith, 1977 p.218). Therefore confidence limits for b_1 / b_2 are

$$\frac{\frac{b_1}{b_2} - \pm \frac{t}{b_2}\sqrt{V\left(b_1\right) + \frac{b_1^2}{b_2^2}V\left(b_2\right) - \frac{t^2}{b_2^2}V\left(b_1\right) \cdot V\left(b_2\right)}}{1 - \frac{t^2}{b_2^2} \cdot V\left(b_2\right)} \tag{17}$$

4.9 Rank Sum Test

Two means may be compared using a t-test if the data described by the means is normally distributed. If not, either a transformation should be applied to the data in order to obtain normal distributions, or a distribution-free test should be used (Davies and Goldsmith, 1977 p. 74). The rank-sum test is a distribution-free test that uses the rank-order of the data :

For 2 data sets with n_1 and n_2 number of observations

- The data sets are combined and ranked in order of increasing value
- Rank 1 is assigned to the lowest value, rank 2 to the next etc. until the highest value which is assigned rank n_1+n_2.
- n_1 = smaller sample size; n_2 = larger sample size; $n = n_1+n_2$
- R is the sum of ranks of the smaller sample
- $R' = n_1 \cdot (n+1) - R$
- The critical value M is available in statistical tables or can be calculated from

$$M = \frac{n_1}{2}(n+1) - u\sqrt{\frac{n_1 n_2 (n+1)}{12}}$$

where $u = 1.96$ for $\alpha = 0.05$ (double - sided test)
and $u = 2.58$ for $\alpha = 0.01$ (double - sided test)

- If R or R'<M then a significant difference exists at a (100-α) confidence level.

APPENDIX A3

ADDITIONAL RESULTS

This appendix contains additional data that were obtained during operation of the pilot-scale ABR that was not included explicitly in the body of the thesis.

1 TABLE OF ANALYSES AND RESPONSIBLE PERSONS

Table A3. 1 presents lists of analyses and people responsible for supervising, sampling, analysis, data capture and data analysis for all of the data presented in this thesis.

2 PHASE I: OPERATION OF PILOT-SCALE ABR AT UMBILO WWTP

2.1 Influent characteristics

Time series concentration data showed considerable scatter, making it difficult to gain an impression of the distribution of the data when looking at a time vs. concentration plot. The rank probit plot presents the data ranked according to concentration and plotted against the equivalent ranked data point from a normal distribution with the same mean and standard deviation as the data set. If the data are normally distributed, the data points fall on a x=y 45º diagonal line. Deviations from the 45º line indicate that the distribution of the data deviates from the normal distribution. These plots make it very easy to understand the scatter of the data since percentile lines may be drawn in corresponding to values at selected percentiles determined from the ranked data. Rank probit plots are presented for COD, alkalinity, free and saline ammonia, phosphate, total solids and pH of the Umbilo influent wastewater.

Table A3. 1: Table of Responsibilities for analyses during pilot-scale ABR experimental study

		Phase I	Phase II	Phase III	Phase IV
COD, Alkalinity, pH, total solids, volatile solids	Supervisor	P Dama	K Foxon	K Foxon	K Foxon
	Sampler	Umbilo WWTP staff	DZ Mtembu & K Foxon	DZ Mtembu & K Foxon (+ EWS)	K Hudson, S Pillay & K Foxon (+ EWS)
	Analyser	Umbilo WWTP staff	DZ Mtembu & K Foxon	DZ Mtembu & K Foxon (+ EWS)	K Hudson & K Foxon (+ EWS)
	Data Capture	K Foxon	DZ Mtembu & K Foxon	DZ Mtembu & K Foxon	K Hudson & K Foxon
	Data Analysis	K Foxon	K Foxon	K Foxon	K Foxon
Ammonia, phosphate	Supervisor	P Dama	K Foxon	K Foxon	K Foxon
	Sampler	Umbilo WWTP staff	DZ Mtembu & K Foxon	DZ Mtembu & K Foxon	K Hudson, S Pillay & K Foxon
	Analyser	Umbilo WWTP staff	EWS	EWS	EWS
	Data Capture	K Foxon	DZ Mtembu & K Foxon	DZ Mtembu & K Foxon	K Hudson & K Foxon
	Data Analysis	K Foxon	K Foxon	K Foxon	K Foxon
VFA	Supervisor				K Foxon
	Sampler				K Hudson, S Pillay & K Foxon
	Analyser				D Moodley, K Foxon
	Data Capture				K Foxon
	Data Analysis				K Foxon

Table A3. 1 cont.

		Phase I	Phase II	Phase III	Phase IV
SO4	Supervisor	P Dama			K Foxon
	Sampler	Umbilo staff			K Hudson, S Pillay & K Foxon
	Analyser	Umbilo staff			D Moodley
	Data Capture	K Foxon			K Foxon
	Data Analysis	K Foxon			K Foxon
Na & K	Supervisor	P Dama			K Foxon
	Sampler	Umbilo staff			K Hudson, S Pillay & K Foxon
	Analyser	Umbilo staff			D Moodley
	Data Capture	K Foxon			K Foxon
	Data Analysis	K Foxon			K Foxon
E Coli, total coliforms, coliphage, Ascaris	Supervisor	P Dama			K Foxon
	Sampler	Umbilo staff			K Hudson, S Pillay & K Foxon
	Analyser	EWS			S Pillay
	Data Capture	K Foxon			S Pillay
	Data Analysis	K Foxon			S Pillay & K Foxon

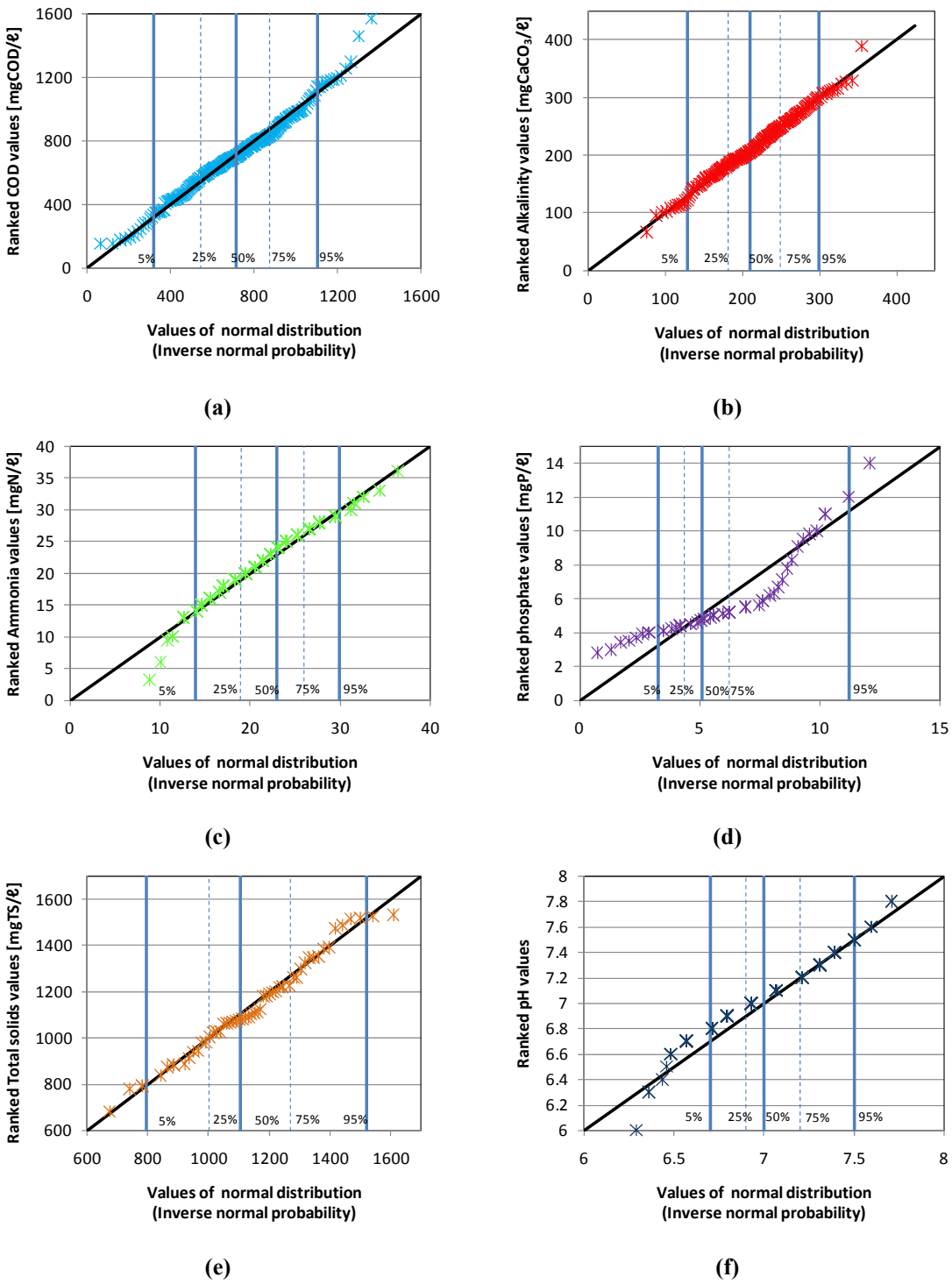


Figure A3. 1: Rank probit plots showing concentrations of determinands in inflow stream during Phase I at Umbilo WWTP (a) COD; (b) alkalinity; (c) free and saline ammonia; (d) total phosphate; (e) total solids; and (f) pH values.

2.2 Pathogen indicator organisms

On 23 April 2001, samples of influent and effluent were tested for total *Ascaris* spp., viable *Ascaris* spp., *E. Coli* and Total Coliforms. On 3 July 2001, samples of influent and effluent, and samples from

each of the eight compartments were analyses for *E. Coli*, total coliforms, *Pseudomonas* spp., *Salmonella* spp., *Vibrio* spp., and *Shigella* spp.

Table A3. 2: Pathogen indicator organisms detected in the influent and effluent of the pilot-scale ABR. Data are single measurements or averages of two measurements (coliforms only) on grab samples obtained on 23 April 2001 and 3 July 2001 during the 20 h T-HRT operating period under PLC control.

Pathogen indicator organism	Influent	Effluent
Total coliforms (cfu./100 mℓ)	> 4 000 000	46 500
E. coli (cfu./100 mℓ)	> 4 000 000	3 500
Total Ascaris spp. (/100 mℓ)	232	298
Viable Ascaris spp.(/100 mℓ)	83	5
Pseudomonas spp. (/100 mℓ)	0	1
Salmonella spp.(/100 m/ℓ)	0	0
Vibrio spp. (/100 mℓ)	0	0
Shigella spp. (/100 mℓ)	0	0

All analyses were performed by Durban Metro Wastewater (now eThekwini Water Services) Laboratories in Prior Road, Durban. **Table 4.1** presents average results for the two sampling days. The small sample numbers (in most cases, one analysis per pathogen indicator organism) meant that descriptive and comparative statistical calculations could not be performed. Comparison between single influent and effluent measurements show that a reduction of at least 2 log units was obtained for coliforms, and viable *Ascaris* spp. reduced from 83 viable eggs per 100 mℓ to 5 viable eggs per 100 mℓ representing a removal of 94%.

3 PHASE II-IV: OPERATION OF PILOT-SCALE ABR AT KINGSBURGH WWTP

3.1 Influent characteristics

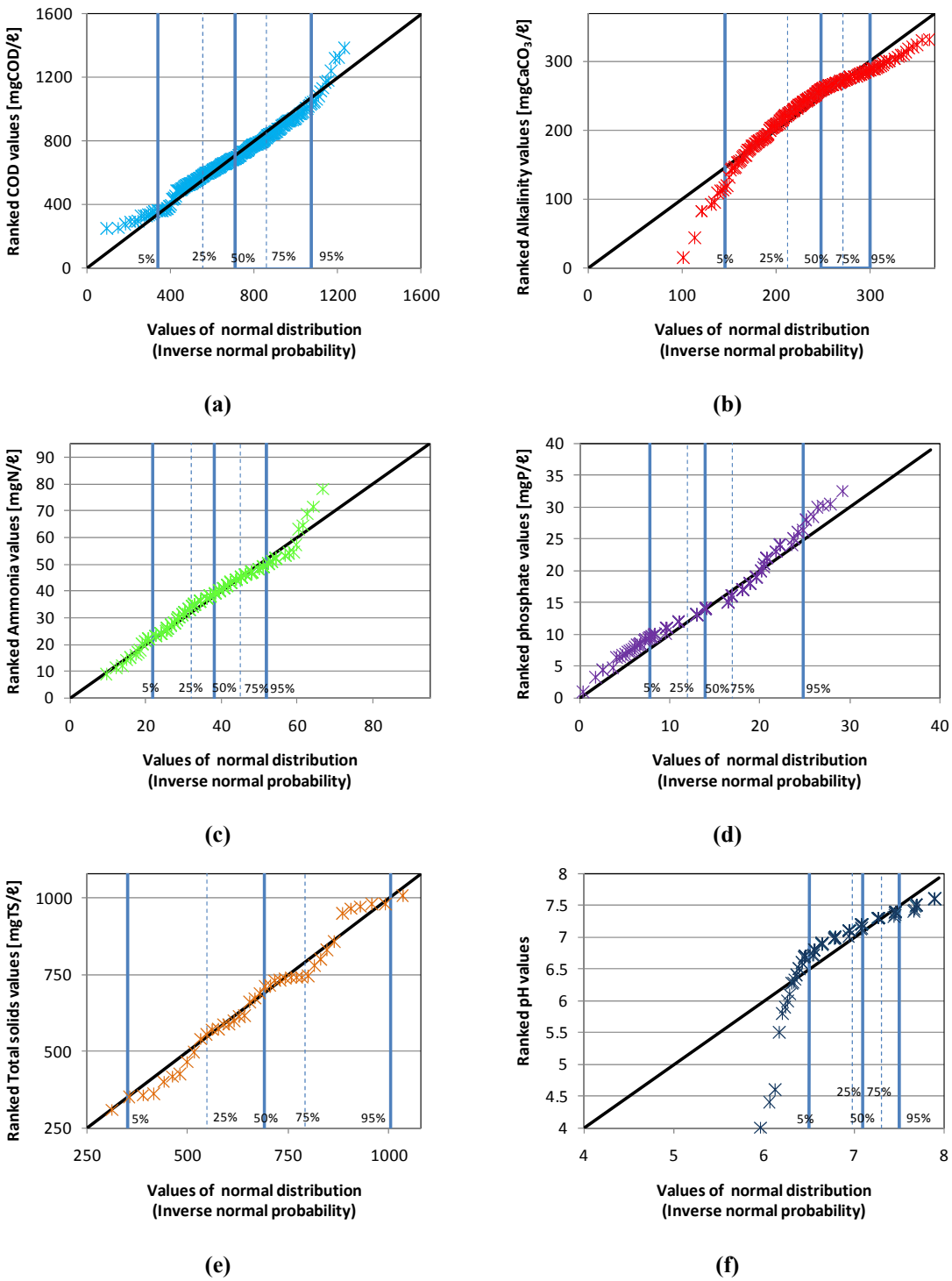


Figure A3. 2: Rank probit plots showing concentrations of determinands in inflow stream during Phase II-IV at Kingsburgh WWTP (a) COD; (b) alkalinity; (c) free and saline ammonia; (d) total phosphate; (e) total solids; and (f) pH values.

3.2 Pathogen indicator organisms

A portion of the MSc research undertaken by Pillay (2006) investigated the fate of pathogen indicator organisms in the pilot-scale ABR during Phase IV. Measurements of *Escherichia coli* (*E. coli*), total coliforms, coliphages and helminth eggs were made on samples obtained from the inflow and outflow of the pilot-scale ABR.

- Total coliforms and *E. coli* were simultaneously determined by the membrane filtration technique according to Standard Methods (APHA, 1998). Coliforms were enumerated as colony forming units (cfu) per 100 mℓ. Results provide an indication of the fate of general bacteria (total coliforms) and bacteria of human faecal origin (*E. coli*) in the pilot-scale ABR.

- Coliphages are routinely used as viral indicators. Measurement of coliphage incidence involves enumerating the bacteriophage of host culture *E. coli* (ATCC 13706) using a double layer technique; bacteriophages cause lysis on a lawn of *E. coli* host cells, forming clear plaques and were enumerated as plaque forming units (pfu) per 100 mℓ.

- Enumeration of helminth eggs was limited to a single helminth genus, namely *Ascaris*. Raw wastewater (1ℓ) and flow from the ABR outlet (10ℓ) were collected on a weekly basis and allowed to sediment for 18 h. The supernatant of samples was discarded and the remaining sediments were centrifuged at 1000 g for 15 min. The centrifuged supernatant was discarded and parasite eggs were counted using a modified Bailenger method (Ayres and Mara, 1996)

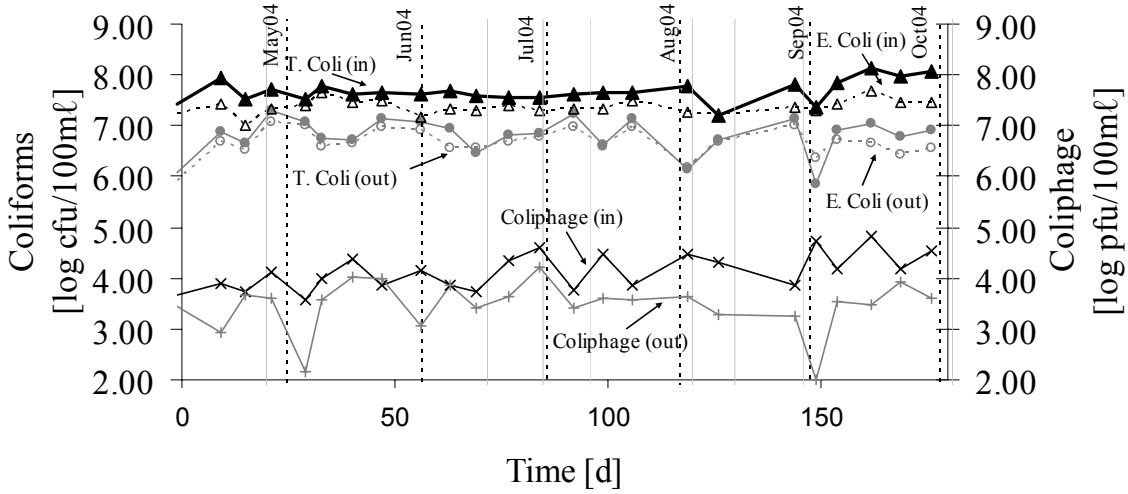


Figure A3. 3: Phase IV: Pathogen indicator organisms in the inflow and outflow. (Mean A-HRT approximately 42 h). Inflow and outflow total coliforms (-▲- and -●-), *E. Coli* (··Δ·· and ··○··) and coliphage (-×- and -+-) are shown.

Average *E. coli* concentration measurements in the pilot-scale ABR outflow ranged from 7 x 10^5 to 1 x 10^7 cfu/100 mℓ with an average reduction from inflow to outflow of 76 % (1.87 log reduction).

Total coliform concentrations in the outflow were in the range 1 x 10^6 to 2 x 10^7 cfu/100 mℓ with an average removal efficiency of 83 % (1.9 log reduction).

Measurements of coliphages in the inflow and outflow provide an indication of the fate of viruses in the pilot-scale ABR. Outflow concentrations of coliphages were between 1×10^2 and 2×10^4 pfu/100 mℓ in the outflow with a mean removal efficiency of 64 % (1.8 log reduction).

Total helminth eggs were counted in samples of inflow and outflow. Eggs counted were not assessed for viability. The concentration of eggs, assumed to be of Ascaris spp. varied substantially between different samples. Large numbers of eggs were found in the inflow, while the number in the outflow was low, but variable. The average egg concentration in the outflow was 17 eggs/ℓ (with a standard deviation of 15 eggs/ℓ). This corresponded to an average removal efficiency of 98%.

The removal efficiency for of the pathogen indicator organisms where all found to be significantly different to 0% (Student's t-test, $P < 10\text{-}3$). From all indicator organisms tested, the greatest reductions were observed for Ascaris eggs. The performance of the ABR in removing helminth eggs is probably attributed to eggs having a larger mean residence time within the reactor due to sedimentation.

Although the reduction of the various indicators was significant, none of the microbial and parasitic parameters met the requirements for discharge, either to water resources or irrigation agriculture. It is therefore likely that the outflow may harbour a wide range of microbial pathogens and parasites, which may present a potential health risk to humans and water supplies. These results indicate that outflow produced by a baffled septic tank or ABR would not be safe for discharge to water course or for agricultural use without a post-treatment step for the removal of pathogenic contaminants.

APPENDIX A4

APPLIED VS. APPARENT HYDRAULIC RETENTION TIME

The A-HRT quoted throughout this thesis were calculated from the applied flow rate and the empty reactor volume i.e.

$$\mathrm{HRT}[\mathrm{h}] = \frac{\text{reactor volume}\left[\mathrm{m}^3\right]}{\text{Flow rate}\left[\mathrm{m}^3/\mathrm{h}\right]} \qquad \textbf{Eq. A4- 1}$$

This implies that fluid spends the same amount of time in the ABR as it would if passed through a plug flow reactor in which the entire working volume was available for through flow.

It was observed that the upflow compartments in the pilot-scale ABR operated as fluidised beds; core samples withdrawn from the upflow section showed a settled sludge bed section, above which a fluidised bed existed. The latter in some cases extended to the top of the compartment, or when the total sludge load of the compartment was lower, extended partway up the compartment and was surmounted by a relatively clear liquid zone.

Presence of solids in each compartment would have resulted in a reduction of the volume of the compartment available for flow of the liquid phase. This concept is depicted graphically in **Figure** A4. 1 where reactor volume available for fluid flow is equal to the total compartment volume less the volume filled with sludge.

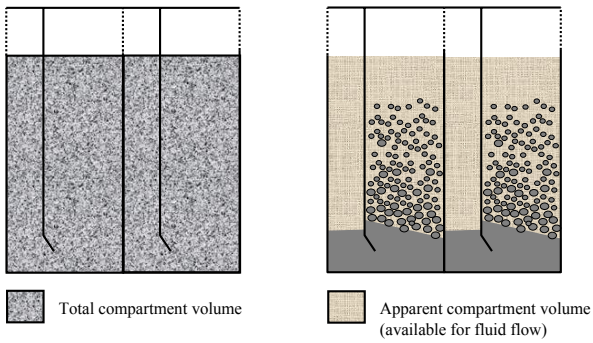


Figure A4. 1: Apparent compartment volume is related to the total compartment volume and the volume occupied by solids that are retained in the compartment.

Thus for a particular flow rate, the average amount of time spent by a package of fluid in any compartment should have been less than that predicted by the clean water CFD modelling undertaken by Dama (Dama et al., 2001).

The A-HRT calculated with Eq. A4- 1 therefore does not tell us the exact residence time of fluid in the reactor, but sets an upper bound for this value. The residence time distribution of both liquids and solids, and more importantly, the contact time distribution (**Section** 2.3.1.2) are important factors affecting the extent of treatment of a wastewater stream in digester. In this section, the hypothesis that an indication of residence time can be obtained by comparing inflow and outflow values of key determinands is tested.

1 RATIONALE

Figure 5.12 and **Figure 5.13** show the diurnal variation of inflow and outflow COD, alkalinity and pH value for the 24 h sampling campaign. Values of Alkalinity and COD concentration in day time inflow samples were higher than night time samples. Outflow COD values showed much less variation than corresponding influent values; however, outflow alkalinity values showed a similar deceasing and then increasing trend to the inflow alkalinity values. It was proposed that high inflow alkalinity corresponded to high outflow values since the baffled configuration of the ABR promotes flow characteristics that approach those of a plug-flow reactor (Barber and Stuckey, 1999) and thus by matching the shape of the inflow and outflow alkalinity curves the fluid retention time (and therefore maximum possible contact time) experienced by the bulk of the fluid could be estimated.[1] The apparent hydraulic retention time is the time lapse between similar alkalinity profiles in influent and effluent. The rationale for this approach is as follows:

- Outlet alkalinity is the sum of the inlet alkalinity and metabolism generated alkalinity (**Section** 2.1.6.3)

- Considerable attenuation of the organic load oscillation is observed between inlet and outlet COD (**Figure 5.13**)

- It is hypothesised that breakdown of biodegradable organic material and thus alkalinity generation proceeds at a relatively steady rate due to the first compartment acting as a buffering zone; in this compartment, solids retention reduces the amplitude of the oscillation in COD load to be treated in the remainder of the reactor

- Thus the *amount* of alkalinity generated will be relatively constant and determined by the steady rate of hydrolysis of particulate components in the buffering zone

- Thus the outlet alkalinity concentration will be the sum of the inlet alkalinity and a relatively constant value for metabolism generated alkalinity, and therefore the outlet alkalinity profile would have a similar shape to the inlet alkalinity profile.

2 COMPARISON OF INFLOW AND OUTFLOW DATA

Correlation coefficients for corresponding inflow and outflow data were determined for apparent retention times between 8 and 22 hours. The significance of the correlation coefficients was determined from the number of degrees of freedom for each case.

Correlation coefficients were calculated for comparison between inflow and outflow alkalinity values from the 24 h campaign (**Section 5.5.2.3**) at different apparent hydraulic retention times as shown in **Figure A4.** 2 below.

[1] This is not the same as the mean residence time, and cannot be used to infer the apparent compartment volume, since it does not take into account eddy volumes within the reactor, or active space associated with micro-organisms into and out of which components may diffuse, resulting in much longer retention times for fluid that passes through these zones.

19 h apparent HRT				14 h apparent HRT		
	n = 21				n = 16	
	Alkalinity values				Alkalinity values	
inflow sample time	inflow	outflow		inflow sample time	inflow	outflow
5/27/03 10:00	254.0032	299.3122		5/27/03 10:00	254.0032	
5/27/03 11:00	202.1441			5/27/03 11:00	202.1441	
5/27/03 12:00	185.905	280.0952		5/27/03 12:00	185.905	
5/27/03 13:00	173.9968	285.2678		5/27/03 13:00	173.9968	
5/27/03 14:00	171.2507	278.0478		5/27/03 14:00	171.2507	
5/27/03 15:00	169.1991	269.519		5/27/03 15:00	169.1991	299.3122
5/27/03 16:00	162.359	266.4764		5/27/03 16:00	162.359	
5/27/03 17:00	254.2003	271.4025		5/27/03 17:00	254.2003	280.0952
5/27/03 18:00	203.4527			5/27/03 18:00	203.4527	285.2678
5/27/03 19:00	187.6578	249.3755		5/27/03 19:00	187.6578	278.0478
5/27/03 20:00	181.7426	268.7387		5/27/03 20:00	181.7426	269.519
5/27/03 21:00	179.8516	261.2244		5/27/03 21:00	179.8516	266.4764
5/27/03 22:00	171.4792	272.247		5/27/03 22:00	171.4792	271.4025
5/27/03 23:00	174.3711	275.0962		5/27/03 23:00	174.3711	
5/28/03 0:00	175.9694	276.4672		5/28/03 0:00	175.9694	249.3755
5/28/03 1:00	173.6886	262.5183		5/28/03 1:00	173.6886	268.7387
5/28/03 2:00	163.8433	275.574		5/28/03 2:00	163.8433	261.2244
5/28/03 3:00	146.4356	265.2663		5/28/03 3:00	146.4356	272.247
5/28/03 4:00	149.2014	276.5242		5/28/03 4:00	149.2014	275.0962
5/28/03 5:00	164.3512	299.0592		5/28/03 5:00	164.3512	276.4672
5/28/03 6:00		286.8892		5/28/03 6:00		262.5183
5/28/03 7:00	208.4026	338.1573		5/28/03 7:00	208.4026	275.574
5/28/03 8:00	293.6795	300.4317		5/28/03 8:00	293.6795	265.2663
5/28/03 9:00	282.7434	286.2702		5/28/03 9:00	282.7434	276.5242
						299.0592
						286.8892
						338.1573
						300.4317
						286.2702

Figure A4. 2: Construction of data tables for correlation analysis for 19 h and 14 h apparent hydraulic retention time. Dot-dash lines indicate the data sets to be compared.

The correlation co-efficient r for each combination set was calculated according to Eq. A4- 2 (Davies and Goldsmith, 1977):

$$r = \frac{\sum_i^n \left(x_i - \bar{x}\right) \sum_i^n \left(y_i - \bar{y}\right)}{\sqrt{\sum_i^n \left(x_i - \bar{x}\right)^2 \sum_i^n \left(y_i - \bar{y}\right)^2}} \qquad \textbf{Eq. A4- 2}$$

Where x_i and y_i are values from the two data sets (x and y) to be compared and $\bar{x}$ and $\bar{y}$ are the average values of each of the data sets.

The significance of the correlation coefficients describes the probability that two data sets are correlated and was calculated from a Student's t-distribution: the t-statistic is determined from Eq. A4- 3, and the probability P was determined from the associated t-distribution with the appropriate number of degrees of freedom using Excel's TDIST function for a 2-tailed application.

$$t = \sqrt{\frac{r^2 \phi}{1 + r^2}} \qquad \textbf{Eq. A4- 3}$$

where ϕ is the degree of freedom and equals n-2.

Figure A4. 3 shows the significance of correlation for apparent hydraulic retention times between 12 and 18 h.

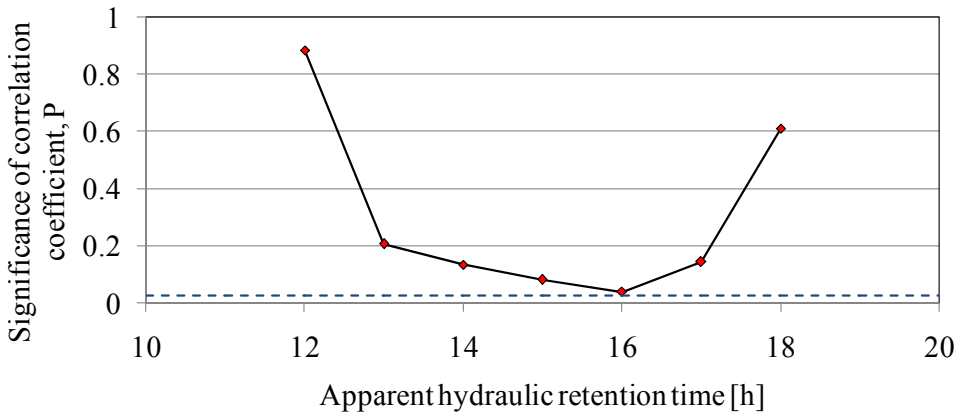


Figure A4. 3: Significance (P) of correlation coefficient between inflow and outflow alkalinity concentration measurements for apparent hydraulic retention times between 12 h and 18 h. Low values of P indicate increased correlation between inflow and outflow. The blue line indicates the two-tailed 95 % confidence level.

Assuming that this methodology for determining apparent hydraulic retention time is reasonable, the most probable correlation between inflow and outflow alkalinity are for an apparent hydraulic retention time of between 14 h and 17 h, with a minimum P value at 16 h. If the value for apparent hydraulic retention is set to less than 12 h or more than 18 h, the correlation coefficient becomes negative (i.e. an increase in inflow concentration is related to a decrease in outflow concentration); therefore these ranges were not considered.

3 CONCLUSION

Thus it was tentatively estimated that the apparent hydraulic retention time i.e. the amount of time that the bulk of the liquid flow spent in the pilot-scale ABR, was around 16 h. This value was used when superimposing outlet data over inlet data in **Figure 5.12** and **Figure 5.13** to show possible correspondence between the data sets. The A-HRT for this period was calculated to be 21.8 h, i.e. the proposed apparent HRT was approximately 75% of the applied HRT.

Ideally, these observations should have been confirmed by a tracer study.

APPENDIX A5

CALCULATION OF SOLIDS RETENTION TIME (SRT) IN STEADY-STATE AND ACCUMULATING SYSTEMS

The solids retention time of a system is often regarded as an indicator that can be used to assess the condition of digestion, and diagnose reactor ills. In this section, the concept of sludge age or solids retention time in an accumulating system is considered.

1 ZEEMAN AND LETTINGA MODEL OF SLUDGE AGE

Assuming the sludge concentration in the reactor (X), the fraction of the influent suspended solids that is removed in the reactor (R) and the fraction of R that is removed through hydrolysis (H) are known:

SRT is defined as the ratio of the sludge concentration in the reactor X [g COD/ℓ] to the (reactor) net sludge production X_P [gCOD/ℓ.d]

$$SRT = X / X_P$$

Eq. A5- 1

The net sludge production is calculated from the OLR [kg COD/m^3.d] and the fraction of the total COD that is associated with suspended solids $f_{SS} = COD_{ss}/COD_{total}$: [1]

$$X_P = OLR \cdot f_{SS} \cdot R \cdot (1 - H)$$

Eq. A5- 2

The OLR by definition is the ratio of COD influent concentration COD_{in} [g COD/m^3] to the hydraulic retention time HRT [d]:

$$OLR = COD_{in} / HRT$$

Eq. A5- 3

Eq. A5- 1, Eq. A5- 2 and Eq. A5- 3 may be solved simultaneously to show that for a desired SRT, the HRT may be calculated by Eq. A5- 4

$$HRT = \left(COD_{in} \cdot f_{SS} / X \right) \cdot R \cdot (1 - H) \cdot SRT$$

Eq. A5- 4

[1] The calculation of excess sludge production does not differentiate between influent suspended solids that accumulate in the reactor and solids that are biomass grown on the hydrolysed COD since these components are measured together in the sludge COD concentration measurement. (Zeeman and Lettinga, 1999)

2 SLUDGE AGE IN AN ACCUMULATING SYSTEM

Sludge age or SRT is commonly calculated for an anaerobic digester and is used as an indication of the maturity of the sludge and the probability of having developed a stable methanogenic population in the sludge (Zeeman and Lettinga, 1999). However, the concept of SRT or sludge age cannot be applied directly to an accumulation system since the sludge age increases with time. The model of Zeeman and Lettinga (1999) presented in **Appendix** A5.1 applies to a system with constant sludge load, i.e. where excess sludge is wasted at the same rate at which it is produced. The definition of SRT here is (from Eq. A5- 3)

$$SRT = X \Big/ X_P$$ **Eq. A5- 5**

With X_P defined as (from Eq. A5- 2 and Eq. A5- 3)

$$X_P = \frac{COD_{in} \cdot f_{SS} \cdot R \cdot (1 - H)}{HRT}$$ **Eq. A5- 6**

Thus for a constant feed composition and constant HRT, and assuming that the fractions of solids retained and hydrolysed do not change substantially with time, X_P can be regarded as constant for a specified feed and operating conditions.

If X_P is essentially constant, then the overall sludge load in the reactor should be calculated by simple integration:

$$X = X_0 + X_P \cdot t$$ **Eq. A5- 7**

where X_0 is the amount of seed sludge present at time 0 and t is the amount of time since start-up.

Using the definition of SRT (Eq. A5- 5):

$$SRT = \frac{X_0 + X_P \cdot t}{X_P} = \frac{X_0}{X_P} + t$$ **Eq. A5- 8**

This cannot be correct since $\frac{X_0}{X_P} > 0$ and therefore SRT > t; i.e. the "sludge age" calculated is longer than the period of operation.

Thus a new mathematical definition of sludge age is required. The concept that we are interested in is the average time that solids have spent inside the reactor. This can be defined mathematically as follows:

$$SRT_A = \sum_i SRT_i \cdot f_i = \sum_i \frac{SRT_i \cdot X_i}{X}$$ **Eq. A5- 9**

i.e. the average SRT of an accumulating system at a time t_e is the sum of the individual sludge ages SRT_i for each age category of sludge multiplied by the fraction of the total sludge load (f_i) with that characteristic age.

Let the entire time of operation since the reactor was commissioned $[t_0, t_e]$ be sliced into time intervals of length Δt. Consider a *package* of sludge generated in the time interval $[t_{i-\Delta t}, t_i]$ (by retention of influent solids or ultimately by generation of biomass through growth). At t_e the sludge age of this *package* of sludge is $(t_e - t_i)$, i.e.

$$SRT_i = t_e - t_i$$ **Eq. A5- 10**

The fraction of the total sludge that this package makes up depends on the rate of sludge production at other times of operation $X_P(t)$ and the initial amount of sludge in the reactor X_0.

If we assume once again that the sludge production rate X_P is approximately constant, then all *packages* of sludge produced after time t_0 will form the same fraction of the accumulated sludge at time t_e. The actual value will depend on the length of the time interval t_e and the value of X_0. The amount of sludge making up each package i will be:

$$COD_i = X_P \cdot \Delta t$$ **Eq. A5- 11**

And as before, the total COD in the reactor X is calculated from Eq. A5- 7. Substituting Eq. A5- 7, Eq. A5- 10 and Eq. A5- 11 into Eq. A5- 9 gives

$$SRT = \sum_i \frac{(t_e - t_i) X_P \cdot \Delta t}{X_0 + X_P \cdot t_e}$$ **Eq. A5- 12**

Note that X_P is not a specific biomass yield, but rather that amount of biomass generated or other solids retained that will accumulate over the long term, and cannot be practically determined for any time i, but may be inferred from long term sludge accumulation data.

Taking the limit of Eq. A5- 12 as $\Delta t \rightarrow 0$ gives

$$SRT = \frac{\int_{t_0=0}^{t_e} (t_e - \tau) X_P \cdot d\tau}{X_0 + X_P \cdot t_e} = \frac{\left(t_e^2 - \frac{1}{2} t_e^2\right) X_P}{X_0 + X_P \cdot t_e} = \frac{\frac{1}{2} t_e^2 X_P}{X_0 + X_P \cdot t_e}$$ **Eq. A5- 13**

Eq. A5- 13 is depicted graphically in **Figure A5.** 1

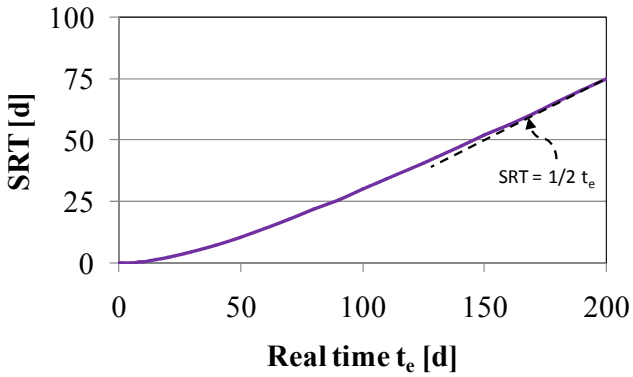


Figure A5. 1: Example of SRT vs. length of operating time for accumulating system model Eq. A5- 13 with X_P = 0.225 kg COD/d and X_0= 15 kgCOD

Near start-up where $X_0 >> X_P \cdot t_e$, Eq. A5- 13 becomes

$$SRT = \frac{t_e^2 X_P}{2 \cdot X_0} \text{ and } \frac{dSRT}{dt} = \frac{X_P}{X_0} t_e \qquad \textbf{Eq. A5- 14}$$

After extended period of operation when $X_P \cdot t_e >> X_0$, Eq. A5- 13 becomes

$$SRT = \frac{1}{2} t_e \text{ and } \frac{dSRT}{dt} = \frac{1}{2} \qquad \textbf{Eq. A5- 15}$$

i.e. in a system with little initial seeding, a sufficiently long time after commissioning, and provided that the solids accumulation rate does not change substantially with time we are left with the neat solution that the sludge age increases at half of actual time i.e. dSRT = 0.5 dt.

This model of sludge age is clearly fairly simplistic and only applies to a pseudo-steady-state condition where changes in overall sludge load do not affect the rate of excess sludge production X_P such as those that prevailed in the pilot-scale ABR.

- Sludge production should in fact be regarded as ultimate sludge production i.e. consist of inert particulate material originating from the influent and biomass produced from digestion of biodegradable COD that will be ultimately retained in the reactor. Thus a single value for excess sludge production can only be determined or applied when considering a long, stable period of operation in the accumulating system.

- The simplest method of determining X_P is from experimental data where X_P is calculated from overall sludge load data using Eq. A5- 7.

- Alternatively, X_P may be estimated using a model such as that of Zeeman and Lettinga (**Eq. A5-4**, Zeeman and Lettinga, 1999) or a similar construction e.g.

  X_P = amount of inert solids from influent that remain in the reactor unchanged + amount of COD which is ultimately converted to biomass

$$X_P = \frac{X_{Inert} \cdot \left(1 - f_{XI,eff}\right)}{HRT} + \frac{\left(COD_{in} - COD_{out} - X_{Inert} \cdot \left(1 - f_{XI,eff}\right)\right)}{HRT}$$

Eq. A5- 16

where X_{inert} is the amount of particulate inert COD in the influent [mgCOD/ℓ]

and $f_{XI,eff}$ is the fraction of the influent particulate inert COD that appears in the effluent during pseudo-steady-state operation.

- This model makes no allowance for solids that are initially retained and later released through e.g. physical displacement or sludge decay. These effects are implicitly contained in the single X_P value. This is not an unreasonable approach since it would be impossible to determine rate constants for these processes if modelled explicitly without long-term solids residence time distribution tests. However, the exclusion from the model may result in overestimation of sludge age if the ultimate sludge production from a package i of influent COD is overestimated from short-term experimental data.

- Finally, the sludge ages calculated from this model are fundamentally different to those calculated for a steady-state system since the probability distribution for solids leaving the reactor are fundamentally different, particularly when considering different micro-organism sub-populations with differing retention characteristics. Thus comparison of sludge ages calculated with the two different methods should be undertaken with care since the behaviour of sludge (e.g. in terms of methanogenic activity) from an accumulating system may well be substantially different to one of the same calculated sludge age from a steady-state system. Specifically, conclusions from other studies that methanogenesis establishes for sludge ages longer than a specific value will not necessarily be valid for the accumulating system with SRT calculated as above.

APPENDIX A6

DETERMINATION OF OUTFLOW CHARACTERISTICS USING THE STEADY-STATE MODEL OF SÖTEMANN ET AL. (2005)

This appendix presents a pseudo-steady-state model of anaerobic digestion system with accumulation of sludge.

Steady-state models are based on the principle of the *rate-determining step*: in a steady-state system, the overall rate of treatment will depend on the slowest process that occurs in the system. Provided the conditions of the system do not change such that another process becomes rate-limiting, a calibrated steady-state model will give a reasonably quick basis for designing a system and determining operating parameters, or estimating system performance under slightly different conditions.

Sötemann et al. (2005) developed a steady-state model for anaerobic digestion of sewage sludges based on the assumption that hydrolysis of macromolecules is the rate-limiting step in this process. This is a three step model consisting of (i) a kinetic part for determining COD removal and gas production, (ii) a stoichiometric part that calculates free and saline ammonia, alkalinity production and digester gas composition and (iii) a weak acid-base section that calculates the digester pH from the gas composition and alkalinity.

The COD of the feed in the steady-state model is assumed to be a combination of particulate biodegradable organic materials (SBCOD, **Section** 2.4.1) with a known average elemental composition $C_XH_YO_ZN_A$, VFA (represented by acetic acid) and a fraction of unbiodegradable material. All VFA is consumed in the process, and a portion of the remaining biodegradable organic material is converted to CH_4, CO_2, alkalinity and free and saline ammonia. The extent of biodegradation, i.e. the amount of SBCOD degraded, depends on the sludge age or length of contact time in the system.

1 THEORY OF STEADY-STATE MODEL

The purpose of this exercise was not to simulate the experimental data, but to determine what the probable outflow conditions would be for a range of inflow characteristics and at different values for the extent of treatment. Therefore, kinetics of biodegradation were not initially considered to be important. However, the kinetic part of the Sötemann steady-state model could not be ignored completely since in the original form, the sludge age and kinetic constants (hydrolysis rate and death rate) determines the apparent yield of the process, E, defined as the fraction of removed COD converted to sludge through microbial growth, and the value of E is necessary for determining overall stoichiometry.

1.1 Sötemann model of growth-death-regeneration

The original Sötemann model makes use of a growth-death-regeneration model, called the hydrolysis model; i.e. biodegradable particulate material (denoted S_{bP}) and VFA (S_{bVFA}) are converted to CH_4, CO_2 and biomass; biomass undergoes endogenous decay, where more S_{bP} is a product of the endogenous decay (**Figure A6.** 1). Thus the initial biomass yield from degradation of feed

biodegradable organic material may be high (Sötemann et al. (2005) recommend a value of 0.113 gCOD/gCOD for Y_{AD}, the yield co-efficient), but the overall yield calculated by mass balance over the system operated at steady state (E) may be significantly lower. Longer sludge ages and higher endogenous rates result in lower values of E, while for short sludge ages and low endogenous rates, E will be close to Y_{AD}.

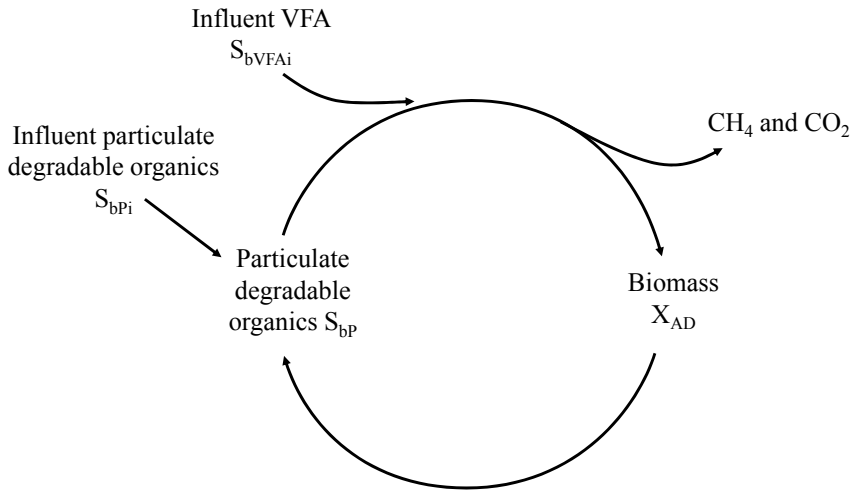


Figure A6. 1: Growth-Death-Regeneration scheme used in the hydrolysis part of the steady-state model of Sötemann et al. (2005)

Sötemann et al. (2005) reviewed a considerable amount of data on primary sludge digestion in order to determine values for rate constants for the hydrolysis model. As there was no experimental data available for calculating the apparent yield, E, it was proposed that the hydrolysis model with kinetic constants proposed by Sötemann et al. (2005) with appropriate corrections for temperature be employed.

However, the original steady-state model makes describes a CSTR with no unsteady sludge accumulation. Consequently, the HRT and sludge age are identical and are calculated from the ration of the volume of the reactor V and feed flow rate Q. As with Zeeman and Lettinga's definition of sludge age (**Section 6.3**), this definition is not appropriate for an accumulating system. Thus a new hydrolysis model had to be developed to describe the effect of sludge retention on apparent yield in an accumulating system.

1.2 Development of a hydrolysis model for an accumulating system

The same general methodology as published in Sötemann et al. (2005) was followed, with the following assumptions and changes:

- A pseudo-steady-state condition was assumed. This implies that the concentrations of reacting species in the digester do not change with time, i.e. biomass and SBCOD concentration are approximately constant. (The corollary of this assumption is that only inert solid material accumulates in the digester.)

- The concentration of solids exiting the digester (X_{Ade}) is a fixed fraction f_X of the total solids concentration in the digester (X_{Ad}):

$$X_{ADe} = f_X X_{AD}$$ **Eq. A6- 1**

- Instead of calculating the outlet concentration of biodegradable particulates, S_{bPe} from system kinetics, an input variable, the extent of treatment f_E was defined in order that outflow conditions may be calculated for ranges of treatment efficiency:

$$f_E = \frac{S_{bPi} - S_{bPe}}{S_{bPi}}$$ **Eq. A6- 2**

Mass balances were performed on biomass (X_{AD}) and biodegradable particulate components (S_{bP}):

$$dS_{bP} \cdot V = Q \cdot S_{bPi} \cdot dt - Q \cdot S_{bPe} \cdot dt - r_H \cdot V \cdot dt + b_{AD} \cdot X_{AD} \cdot V \cdot dt$$
$$\frac{dS_{bP}}{dt} = \frac{Q}{V}\left(S_{bPi} - S_{bPe}\right) - r_H + b_{AD} \cdot X_{AD}$$ **Eq. A6- 3**

$$dX_{AD} \cdot V = -Q \cdot X_{ADe} \cdot dt + Y_{AD} \cdot r_H \cdot V \cdot dt - b_{AD} \cdot X_{AD} \cdot V \cdot dt$$
$$\frac{dX_{AD}}{dt} = \frac{-Q}{V} X_{ADe} + Y_{AD} \cdot r_H - b_{AD} \cdot X_{AD}$$ **Eq. A6- 4**

(These balances assume that the overall rate of hydrolysis is a (first order) function of biomass concentration, and not biodegradable particulate substrate (S_{bP}) concentration.)

The pseudo-steady-state condition is that

$$\frac{dS_{bP}}{dt} = \frac{dX_{AD}}{dt} = 0$$ **Eq. A6- 5**

Setting Eq. A6- 3 and Eq. A6- 4 to 0, rearranging to solve for r_H and equating to eliminate r_H gives

$$\frac{Q}{V} \cdot Y_{AD} \cdot \left(S_{bPi} - S_{bPe}\right) + Y_{AD} \cdot b_{AD} \cdot X_{AD} = \frac{Q}{V} \cdot X_{ADe} + b_{AD} \cdot X_{AD}$$ **Eq. A6- 6**

Solving Eq. A6- 6 for X_{AD} yields

$$X_{ADe} = Y_{AD} \cdot \left(S_{bPi} - S_{bPe}\right) + \frac{V}{Q}\left(Y_{AD} - 1\right) \cdot b_{AD} \cdot X_{AD}$$ **Eq. A6- 7**

Substituting Eq. A6- 1 in Eq. A6- 7 gives

$$X_{ADe} = \frac{f_X \cdot Y_{AD}}{f_X + \frac{V}{Q} \cdot b_{AD} \cdot (1 - Y_{AD})} (S_{bPi} - S_{bPe})$$

Eq. A6- 8

The apparent system yield is the fraction of removed COD converted to sludge through microbial growth, which from Eq. A6- 8 is

$$E = \frac{f_X \cdot Y_{AD}}{f_X + \frac{V}{Q} \cdot b_{AD} \cdot (1 - Y_{AD})}$$

Eq. A6- 9

In the Sötemann model, the CH_4 is produced from digestion of SBCOD and VFA. The amount of CH_4 produced from SBCOD is related to the rate of hydrolysis of SBCOD. It is assumed that the VFA are converted entirely to CH_4, since the yield co-efficient for methanogens is small, and the error introduced would be small compared to the yield of hydrolytic processes. This assumption is valid for a hydrolysis rate-limited system, where influent VFA are a small portion of the overall influent COD.

The overall rate of CH_4 production can be described by

$$S_M = \frac{(1 - f_X \cdot Y_{AD})}{f_X + \frac{V}{Q} \cdot b_{AD} \cdot (1 - Y_{AD})} \cdot (S_{bPi} - S_{bPe}) + S_{bVFAi} = (1 - E) \cdot (S_{bPi} - S_{bPe}) + S_{bVFAi}$$

Eq. A6- 10

where

f_X	=	Fraction of total solids concentration in reactor that leaves with the outflow
Y_{AD}	=	Yield co-efficient for acidogenic micro-organisms
V	=	Volume of digester [m^3]
Q	=	Feed flow rate [m^3/d]
b_{AD}	=	Endogenous decay rate [d^{-1}]

1.3 Stoichiometry of digestion

The stoichiometric expression describing anaerobic conversion of $C_XH_YO_ZN_A$ to CO_2 and CH_4 used by Sötemann et al. (2005) is given in Eq. A6- 11

$$C_X H_Y O_Z N_A + \left(2X + A - Z - \frac{9ED}{20} - \frac{(1-E)D}{4}\right) H_2O + \left(X - A - \frac{ED}{5} - \frac{(1-E)D}{8}\right) CO_2 +$$
$$\left(\frac{(1-E)D}{8}\right) CH_4 + \left(\frac{ED}{20}\right) C_5H_7O_2N + \left(A - \frac{ED}{20}\right) NH_4^+ + \left(A - \frac{ED}{20}\right) HCO_3^-$$

Eq. A6- 11

where D is the electron equivalence of SBCOD with characteristic stoichiometric representation of $C_XH_YO_ZN_A$:

$$D = 4X + Y - 2Z - 3A \quad (\text{e}^- \text{ eq./mol}) \qquad \textbf{Eq. A6- 12}$$

$C_XH_YO_ZN_A$ has a COD of

$$COD = 8 \cdot [Y + 2 \cdot (2 \cdot X - Z) - 3 \cdot A] \quad (\text{gCOD}/\ell) \qquad \textbf{Eq. A6- 13}$$

And a molar mass

$$MM = 12 \cdot X + Y + 16 \cdot Z + 14 \cdot A \quad (\text{gVSS/mol}) \qquad \textbf{Eq. A6- 14}$$

Then if ν_j is the stoichiometric coefficient of component j in Eq. A6- 11, then for all components *except VFA and CH4* the amount of component j produced in the steady-state model is calculated as

$$\Delta S_j = \nu_j \cdot (S_{bPi} - S_{bPe}) \qquad \textbf{Eq. A6- 15}$$

and the amount of CH_4 produced (S_M) is calculated from Eq. A6- 10.

The outflow concentration of each component j is determined as

$$S_{je} = S_{ji} \cdot \Delta S_j \qquad \textbf{Eq. A6- 16}$$

1.4 Weak acid-base chemistry

The digester pH value in a methanogenic reactor rate-limited by hydrolytic processes is dominated by the partial pressure of CO_2 in the headspace, P_{CO2}, the bicarbonate concentration and the ionic strength in solution. The partial pressure of CO_2 is calculated from the ratio of the production of CO_2 and CH_4. The dependence of pH on P_{CO2} and $[HCO_3^-]$ is expressed in Eq. A6- 17 (Sötemann et al., 2005)

$$P_{CO_2} = \frac{[HCO_3^-] \cdot \left(1 + 10^{pK'_{C1} - pH} + 10^{pH - pK'_{C2}}\right)}{10^{-pK'_{HCO2}} \cdot \left(1 + 10^{pH - pK'_{C1}} + 10^{2pH - pK'_{C1} - pK'_{C2}}\right)} \qquad \textbf{Eq. A6- 17}$$

where

$[HCO_3^-]$	=	bicarbonate concentration (which is approximately equal to total alkalinity) [mol/ℓ]
P_{CO2}	=	partial pressure of CO_2 in the gas phase [atm or mol fraction]
pK'_{HCO2}	=	$-\log_{10}$ of apparent Henry's law constant for CO_2
pK'_{C1}, pK'_{C2}	=	$-\log_{10}$ of apparent dissociation constants for carbonate system corrected for ionic strength

1.5 Implementation of steady-state model

The model structure presented in **Sections** 1.2 to 0 is essentially the same as that in Sötemann et al. (2005) except that a sludge retention factor f_X has been applied such that the concentration of biomass X_{AD} and SBCOD S_{bP} are not the same as in the effluent. This allows much lower effective yield co-efficients (E) to be calculated at the operating HRT used in the pilot-scale ABR study.

The steady-state sludge retention model was applied as follows:

Hydrolysis model

- For a fixed extent of treatment, f_E the outlet SBCOD concentration was calculated from Eq. A6- 2
- The fraction of solids retained in each compartment was estimated from experimental data
- The value of E was calculated from Eq. A6- 9
- The amount of CH_4 produced was calculated from Eq. A6- 10

Stoichiometry Model

- The value of E calculated above was used in Eq. A6- 11 to determine the amount of CO_2, HCO_3^-, NH_4^+ and biomass formed
- The bicarbonate alkalinity was calculated from Eq. A6- 15 and Eq. A6- 16 for HCO_3
- The P_{CO2} was determined from Eq. A6- 15 and Eq. A6- 16 for CO_2 and from the CH_4 production calculated in the hydrolysis model

Weak acid-base model

- The digester pH value was determined using Eq. A6- 17

This model is limited to applications where the digester pH falls in the range 6.5 – 8.5, and where the bicarbonate concentration dominates the total alkalinity concentration (or where bicarbonate concentration data are available).

2 INPUTS INTO THE STEADY-STATE MODEL OF THE ABR

Anaerobic digestion feedstocks are conventionally described in terms of their carbohydrate, protein and lipid components (Batstone et al., 2002) since each of these categories may be represented by characteristics elemental compositions; i.e. carbohydrates have elemental compositions similar to $(CH_2O)_n$; proteins contain nitrogen, and lipids have high C:O and H:O ratios. Although it is not usually practical to characterise the feed by measuring these constituents, it is useful to represent the feed in terms of these since it is easy to visualise changes in feed composition in terms of the relative contribution of each of these categories. Pillay (Pillay, 2006) undertook a limited number of carbohydrate measurements using the method of Dreywood, (Dreywood, 1946) with minor modifications (Raunkjear et al., 1994) using an anthrone reagent, and protein measurements using the Lowry method (Lowry et al., 1951).

The average elemental composition of generic carbohydrate, lipid and protein compositions taken from Henze et al. (Henze, 1992) were used to calculate the overall average elemental composition of the biodegradable organic material in the feed as follows:

Carbohydrate:	$C_{10}H_{18}O_9$	fraction = i (mol %)
Lipid:	$C_8H_6O_2$	fraction = j (mol %)
Protein:	$C_{14}H_{12}O_7N_2$	fraction = k (mol %)

The average composition of biodegradable organics (SBCOD) can be represented by $C_XH_YO_ZN_A$

where $X=(i\cdot 10 + j\cdot 8 + k\cdot 14)/100$
$Y=(i\cdot 18 + j\cdot 6 + k\cdot 12)/100$
$Z=(i\cdot 9 + j\cdot 2 + k\cdot 7)/100$
$A=(k\cdot 2)/100$

Eq. A6- 18

A filter converting feed SBCOD composition from the carbohydrate-lipid-protein characterisation to the elemental composition ($C_XH_YO_ZN_A$) used in the steady-state model was developed as follows:

Consider that data are available as carbohydrate (C), lipid (L) and protein (P) as a proportion of total SBCOD in COD units, e.g.

$$\%C = \frac{100\cdot C}{SBCOD}\frac{[gCOD/\ell]}{[gCOD/\ell]}\%; \quad \%L = \frac{100\cdot L}{SBCOD}\frac{[gCOD/\ell]}{[gCOD/\ell]}\%; \quad \%P = \frac{100\cdot P}{SBCOD}\frac{[gCOD/\ell]}{[gCOD/\ell]}\%$$ **Eq. A6- 19**

The COD content of each of the component k can be calculated from the following expression:

$$COD_k = X_k + \frac{Y_k}{4} - \frac{Z_k}{2} - \frac{3A_k}{4}\left[molCOD/mol\right]$$ **Eq. A6- 20**

i.e.

Carbohydrate	(C)	$C_{10}H_{18}O_9$	;	COD_C = 10 mol COD/mol
Lipid	(L)	$C_8H_6O_2$	;	COD_L = 8.5 mol COD/mol
protein	(P)	$C_{14}H_{12}O_7N_2$	;	COD_P = 12 mol COD/mol

The composition in mole units for each of the components k is calculated as follows:

$$mol\%k = \frac{\%k\cdot COD_k}{\%C\cdot COD_C + \%L\cdot COD_L + \%P\cdot COD_{=P}}$$ **Eq. A6- 21**

Finally, the composition of a composite SBCOD component with COD composition according to Eq. A6- 19 is calculated using Eq. A6- 18 such that the value of Y is 7 and the other three coefficients, X, Z and A are scaled accordingly, i.e.

$$X = 7 \cdot \frac{(10 \cdot mol\%C + 8 \cdot mol\%L + 14 \cdot mol\%P)}{(18 \cdot mol\%C + 6 \cdot mol\%L + 12 \cdot mol\%P)}$$

$$Y = 7$$

$$Z = 7 \cdot \frac{(9 \cdot mol\%C + 2 \cdot mol\%L + 7 \cdot mol\%P)}{(18 \cdot mol\%C + 6 \cdot mol\%L + 12 \cdot mol\%P)}$$

$$A = 7 \cdot \frac{(2 \cdot mol\%P)}{(18 \cdot mol\%C + 6 \cdot mol\%L + 12 \cdot mol\%P)}$$ **Eq. A6- 22**

(The representation of composite organics with a Y coefficient with a value of 7, i.e. $C_XH_7O_ZN_A$ is historical and originates from the conventional representation of biomass as $C_5H_7O_2N$.)

APPENDIX A7

PUBLICATIONS ARISING FROM THIS PROJECT

This section contains a list of publications emanating from this project. Publications

1 RESEARCH REPORT

Foxon KM, Buckley CA, Brouckaert CJ, Dama P, Mtembu Z, Rodda N, Smith M, Pillay S, Arjun N, Lalbahadur T and Bux F. (2006) Evaluation of the Anaerobic Baffled Reactor for Sanitation in Dense Peri-urban Settlements. Water Research Commission Report No. 1248/1/06, ISBN 1-77005-371-9

2 JOURNAL ARTICLES

Dama P., Bell J., Foxon K., Brouckaert C. Huang T., Buckley C. Naidoo V., and Stuckey D. (2002) Pilot-scale Study of an Anaerobic Baffled Reactor for the Treatment of Domestic Wastewater *Wat. Sci. Technol.* **46** (9), pp. 263-270

Foxon K.M., Brouckaert C.J., Remigi E. and Buckley C.A. (2004) The anaerobic baffled reactor: An appropriate technology for onsite sanitation. Water SA, **30**(5) pp.44-50 (special edition).

Lalbahadur T, Pillay S, Rodda N, Smith M, Buckley CA, Holder F, Bux F, and Foxon KM (2005) Microbiological Studies of an Anaerobic Baffled Reactor: Microbial Community Characterisation Deactivation of Health-related Indicator Bacteria. Water Science and Technology, 51 (10), pp. 155-162

3 CONFERENCE PROCEEDINGS

Dama P., Govender K., Huang T., Foxon K.M., Bell J., Brouckaert C.J., Buckley C.A, Naidoo, V. and Stuckey D. (2001). Flow Patterns in an Anaerobic Baffled Reactor. *Proceedings Part 1. 9th World Congress on Anaerobic Digestion*, Antwerp, Belgium, 2-5 September 2001, pp. 793 - 798.

Bell, J., Dama, P., Govender, K.M., Buckley, C.A. and Stuckey, D.C. (2001) Performance characterisation and microbial populations associated with the start-up of a laboratory-scale and a pilot-scale anaerobic baffled reactor. *Proceedings Part 2. 9th World Congress on Anaerobic Digestion*, Antwerp, Belgium, 2-5 September 2001, pp. 389-392 (poster).

Dama P, Bell J, Naidoo V, Foxon KM, Brouckaert CJ, Buckley CA and Stuckey D (2001). The Anaerobic Baffled Reactor for the Treatment of Domestic Wastewater in Dense Peri-Urban Communities. *Proceedings Part 2. 9th World Congress on Anaerobic Digestion*, Antwerp, Belgium, 2-5 September 2001, pp. 385-388 (poster).

Dama P., Bell J., Foxon K.M., Brouckaert C.J., Huang T, Buckley C.A., Naidoo V. and Stuckey D. (2001) Pilot-scale study of an anaerobic baffled reactor for the treatment of domestic wastewater. *Proceedings. IWA Conference on Water and Wastewater Management in Developing Countries*, Kuala Lumpur, Malaysia, 29-31 October 2001.

Dama, P., Bell J., Naidoo, V., Foxon, K.M., Brouckaert, C.J., Buckley, C.A. and Stuckey, D. (2001) The anaerobic baffled reactor for the treatment of domestic wastewater in dense peri-urban communities. *Proceedings. IWA Conference on Water and Wastewater Management in Developing Countries*, Kuala Lumpur, Malaysia, 29-31 October 2001.

Foxon K.M., Dama P., Bell J. and Buckley C.A. (2001) Design considerations for the implementation of an anaerobic baffled reactor in low-income settlements in KwaZulu-Natal. *Proceedings. IWA Conference on Water and Wastewater Management in Developing Countries*, Kuala Lumpur, Malaysia, 29-31 October 2001.

Dama P., Bell J., Foxon K.M., Brouckaert C.J., Huang T., Buckley C.A., Naidoo V. and Stuckey D. (2002) Pilot-scale study of an anaerobic baffled reactor for the treatment of domestic wastewater. *Proceedings. 2002 Water Institute of Southern Africa (WISA) Biennial Conference and Exhibition*, Durban, South Africa, 19-23 May 2002 (poster).

Foxon, K.M., Dama, P. and Buckley C.A. (2002). Application of aquatic modelling of the design of an anaerobic baffled reactor for peri-urban sanitation. *Proceedings. 2002 Water Institute of Southern Africa (WISA) Biennial Conference and Exhibition*, Durban, South Africa, 19-23 May 2002 (poster).

Foxon, K.M., Brouckaert, C.J. and Buckley, C.A. (2003) The application of an Anaerobic Baffled Reactor (ABR) to sanitation in peri-urban communities. *Proceedings. WWWest 2003 meeting of WEST Scientific Support Centres,* 14 - 16 April 2003.

N. Arjun, M. Smith, N. Rodda and C. Buckley (2003) Determination of quality of drinking water from standpipes and ground tanks in the Cato Manor peri-urban settlement 16th SASM-KZN Symposium, Durban 2003

S. Pillay, N. Rodda, M. Smith, K. Foxon, C. Buckley (2003) Pathogen Deactivation and Microbial Community Analysis of an Anaerobic Baffled Reactor. *Proceedings 16th Annual Symposium of the South African Society for Microbiology –KZN*, 2003.

Mackay, N., Smith, A., Rodda, N. and Buckley, C.A. (2003) Feasibility study of the use of effluent from an Anaerobic Baffled Reactor (ABR) for irrigation in a peri-urban community (Cato Manor, Durban) *Proceedings. International Water Association Symposium on Health-Related Water Microbiology*, Cape Town, South Africa, 14-19 September 2003. (poster).

Lalbahadur T., Pillay S., Rodda N., Smith M., Buckley C. Holder, F., Bux F. and Foxon K. (2004) Microbiological Studies of an Anaerobic Baffled Reactor : Microbial Community Characterisation Deactivation of Health-related Indicator Bacteria *Proceedings*: 1st International Conference on Onsite Wastewater Treatment and Recycle, Esplanade Hotel Fremantle, Western Australia, 11-13 February 2004.

Foxon K., Mtembu D., Pillay S., Rodda N., Smith M. and Buckley C. (2004) The Anaerobic Baffled Reactor (ABR): An Appropriate Technology for On-site Sanitation *Proceedings*: 1st International Conference on Onsite Wastewater Treatment and Recycle, Esplanade Hotel Fremantle, Western Australia, 11-13 February 2004

Foxon, K.M. Brouckaert, C.J. Remigi, E.U. and Buckley, C.A. (2004) Biochemical modelling of the anaerobic baffled reactor. *Proceedings. 2004 Water Institute of Southern Africa (WISA) Biennial Conference and Exhibition*, Cape Town, South Africa, 2-6 May 2004.

Foxon K.M., Brouckaert C.J., Remigi E. and Buckley C.A. (2004) The anaerobic baffled reactor: An appropriate technology for onsite sanitation. *Proceedings. 2004 Water Institute of Southern Africa (WISA) Biennial Conference and Exhibition*, Cape Town, South Africa, 2-6 May 2004.

Arjun N., Joubert J-P, Smith M., Rodda N. and Buckley C. (2004) Water Quality at Point for Users Supplied by Standpipes and Groundtanks in a Peri-urban Community *Proceedings: Sp*ecialist Conference on Water and Wastewater Management for Developing Countries (WAMDEC 2004) : Innovative And Sustainable Approaches For Serving The Poor, Elephant Hills Hotel, Victoria Falls, Zimbabwe, 28-30 July 2004

Pillay S., Foxon K., Smith M., Rodda N., and Buckley C. (2004) Treatment performance and effluent quality from an anaerobic baffled reactor treating domestic wastewaters : Implications for water reuse *Proceedings: Sp*ecialist Conference on Water and Wastewater Management for Developing Countries (WAMDEC 2004) : Innovative And Sustainable Approaches For Serving The Poor, Elephant Hills Hotel, Victoria Falls, Zimbabwe, 28-30 July 2004

Remigi, E.U. and Foxon K.M. (2004) Modelling of anaerobic digestion processes: A review. *Proceedings. Research and Innovation in Wastewater Treatment*. Workshop dedicated to the memory of Prof. Alberto Rozzi. Milan – Bologna, 24 – 26 November 2004, Politecnico di Milano, pp. 92 – 112.

Pillay S, Foxon K., Rodda N., Smith M. and Buckley C., The use of effluent from an anaerobic baffled reactor (ABR) for irrigation in a peri-urban community. *Proceedings*: Third International Conference on Ecological Sanitation, International Convention Centre, Durban, South Africa, 23-27 May 2005. (Paper)

Foxon K. and Buckley C., The ABR Concept: A holistic approach to managing wastewater from low-income communities and institutions. *Proceedings*: Third International Conference on Ecological Sanitation, International Convention Centre, Durban, South Africa, 23-27 May 2005. (Paper)

Arjun N., Foxon K., Smith M., Rodda N., and Buckley C. (2005) Characterisation of the wastewater produced by a developing community. *Proceedings*: Third International Conference on Ecological Sanitation, International Convention Centre, Durban, South Africa, 23-27 May 2005. (Poster)

Foxon K.M. and C.A. Buckley (2006) Guidelines for the implementation of anaerobic baffled reactors for on-site or decentralized sanitation. *Proceedings: WISA Conference 2006*, Durban 21 -24 May 2006 (Paper)

Foxon KM and C.A Buckley (2007) Analysis of a pilot-scale anaerobic baffled reactor. *Proceedings: IWA 11th Congress on Anaerobic Digestion 2007*, Brisbane 23 -27 September 2007 (Paper)

4 THESES

Mtembu Z. (2005) *The anaerobic baffled reactor for sanitation in dense peri-urban settlements.* MScEng Thesis, University of KwaZulu-Natal

Lalbahadur T. (2004) *Characterisation of the microbial communities present in an anaerobic baffled reactor utilising molecular techniques*. MTech Thesis, Durban Institute of Technology

Pillay S. (2006). *Microbiological studies of an anaerobic baffled reactor (ABR) treating domestic wastewater*. MSc Thesis. School of Biological and Conservation Sciences. University of KwaZulu-Natal. Durban.